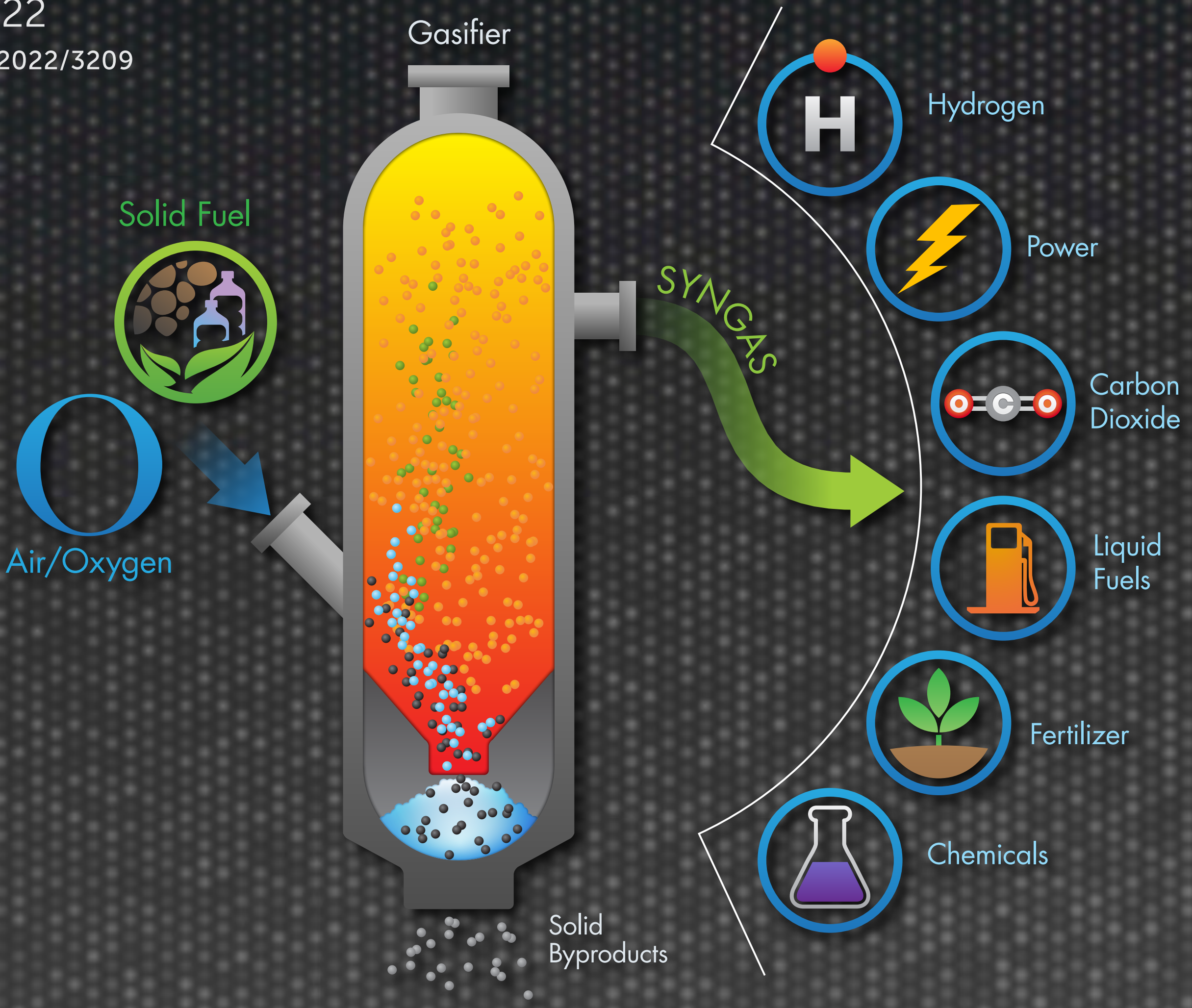
GUIDELINES/HANDBOOK FOR THE DESIGN OF MODULAR GASIFICATION SYSTEMS

JULY 2022

DOE/NETL-2022/3209



U.S. DEPARTMENT OF ENERGY

NATIONAL ENERGY TECHNOLOGY LABORATORY

DISCLAIMER

This project was funded by the United States Department of Energy, National Energy Technology Laboratory, in part, through a site support contract. Neither the United States Government nor any agency thereof, nor any of their employees, nor the support contractor, nor any of their employees, makes any warranty, express or implied, or assumes any legal liability or responsibility for the accuracy, completeness, or usefulness of any information, apparatus, product, or process disclosed, or represents that its use would not infringe privately owned rights. Reference herein to any specific commercial product, process, or service by trade name, trademark, manufacturer, or otherwise does not necessarily constitute or imply its endorsement, recommendation, or favoring by the United States Government or any agency thereof. The views and opinions of authors expressed herein do not necessarily state or reflect those of the United States Government or any agency thereof. Images in this report were created by NETL, unless otherwise noted.

AUTHORS

**Henry Allen Long III,[1,2] Barbara Arnold,[1,2] Bob Brown,[1,2]
David Crouse,[1,2] Santosh Gangwal,[1,2] Steve Knudsen,[1,2]
Chris Munson,[1,2] Norman Shilling,[1,2] Bob Slettehaugh,[1,2]
Jim Sorensen,[1,2] Brian Turk,[1,2] Massood Ramezan[1,2]**

CONTACT

David Lyons[1*]

[1]National Energy Technology Laboratory (NETL)
[2]NETL support contractor
*K.Lyons@netl.doe.gov, 304.285.4379

The authors wish to acknowledge the excellent guidance, contributions, and cooperation of the following NETL support contractors:
**David Denton, Duane Myers, Brad Piggott,
Andrew Sexton, Douglas Todd**

Suggested Citation:

National Energy Technology Laboratory (NETL), 2022, “Guidelines/Handbook for the Design of Modular Gasification Systems,” U.S. Department of Energy, Pittsburgh, PA, DOE/NETL-2022/3209, July 2022.

TABLE OF CONTENTS

LIST OF TABLES

LIST OF FIGURES

ACRONYMS AND ABBREVIATIONS

$/kW	Dollars per kilowatt
°C	Degrees Celsius
°F	Degrees Fahrenheit
A	Amperes
A&E	Architecture and engineering
AC	Alternating current
ACC	Air-cooled condenser
ACHE	Air-cooled heat exchanger
Aeros	Aeroderivatives
AGR	Acid gas removal
Al	Aluminum
Al_2O_3	Aluminum oxide
ALARP	As low as reasonably practicable
ANSI	American National Standards Institute
API	American Petroleum Institute
Ar	Argon
As	Arsenic
ASF	Anderson-Schultz-Flory
AsH_3	Arsine
ASI	Air-side integration
ASME	American Society of Mechanical Engineers
ASTM	American Society for Testing Materials
ASU	Air separation unit
AVT	All-volatile treatment
Ba	Barium
BACT	Best available control technology
Be	Beryllium
BESS	Battery energy storage system
BFB	Bubbling fluidized bed
BGL	British Gas Lurgi
bpd	Barrels per day
BPM	Best Practices Manual
Br	Bromine
BTL	Biomass-to-liquids
Btu	British thermal units
C	Carbon
C_2H_4	Ethylene
Ca	Calcium
CAA	Clean Air Act
CaO	Calcium oxide
CBTL	Coal/biomass to liquid
CCPS	Center for Chemical Process Safety
CCS	Carbon capture and storage
CCUS	Carbon capture, utilization, and storage
CCW	Closed cooling water system
Cd	Cadmium
CEMS	Continuous emissions monitoring system
CFB	Circulating fluidized bed
CFD	Computational fluid dynamics
CFR	Code of Federal Regulations
CGA	Compressed Gas Association
CH_3OH	Methanol
CH_4	Methane
CHP	Combined heating and power
CI	Compression ignition
CIM	Common Information Model
Cl	Chlorine
Co	Cobalt
CO	Carbon monoxide
CO_2	Carbon dioxide
COE	Cost of electricity
COPE	Claus Oxygen-Based Process Expansion
COS	Carbonyl sulfide
CPUC	California Public Utilities Commission
Cr	Chromium
CS_2	Carbon disulfide
CTL	Coal-to-liquids
Cu	Copper
CuO	Copper oxide
CWA	Clean Water Act
CWS	Circulating water system
DC	Direct current
DCS	Distributed control system
DEA	Diethanolamine
DER	Distributed energy resource
DLE	Dry low emission
DLN	Dry, low-NOx
DME	Dimethyl ether
DMPEG	Dimethyl ethers of polyethylene glycol
DNP3	Distributed Network Protocol 3
DOE	U.S. Department of Energy
DSIRE	Database of State Incentives for Renewables & Efficiency

ECN	Energy Research Centre of the Netherlands
EIA	U.S. Energy Information Administration
EIGA	European Industrial Gas Association
EOR	Enhanced oil recovery
EPA	U.S. Environmental Protection Agency
EPACT 2005	Energy Policy Act of 2005
EPC	Engineering, procurement, and construction
EPRI	Electric Power Research Institute
F	Fluorine
F-T	Fischer-Tropsch
FBR	Fixed-bed reactor
Fe	Iron
FEED	Front-end engineering design
FERC	Federal Energy Regulatory Commission
FGRU	Flare gas recovery unit
FSNL	Full-speed, no-load
FTS	Fischer-Tropsch synthesis
GAN	Gaseous nitrogen
GC	Gas chromatograph
GE	General Electric
GHG	Greenhouse gas
GJ	Gigajoule
GOX	Gaseous oxygen
GSA	U.S. General Services Administration
Gt	Gigaton
GT	Gas turbine
GTL	Gas-to-liquids
GWh	Gigawatt-hours
GWth	Gigawatt-thermal
H_2	Hydrogen
H_2O	Water
H_2S	Hydrogen sulfide
HAP	Hazardous air pollutant
HAZOP	Hazard and operability
HCl	Hydrogen chloride
HCN	Hydrogen cyanide
HDGT	Heavy-duty gas turbine
He	Helium
Hg	Mercury
HHV	Higher heating value
HMI	Human machine interface
HP	High pressure
HRSG	Heat recovery steam generators
HTDP	High-temperature desulfurization process
HTFT	High-temperature Fischer-Tropsch
HTS	High-temperature shift
HVDC	High-voltage direct current
Hz	Hertz
I	Iodine
I/O	Input/output
IAPWS	International Association of the Properties of Water and Steam
IC	Internal combustion
ICE	Internal combustion engine
ICI	Imperial Chemical Industries
IEA	International Energy Agency
IEC	International Electrotechnical Commission
IEEE	Institute of Electrical and Electronics Engineers
IGCC	Integrated gasification combined cycle
IGV	Inlet guide vane
IMPCA	International Methanol Producers and Consumers Association
IoT	Internet of Things
IOU	Investor-owned utility
IP	Intermediate pressure
IPP	Independent power producer
ISA	International Society of Automation
ISO	International Organization for Standardization
JHA	Job hazard analysis
JM	Johnson Matthey
JV	Joint venture
K	Kelvin
K	Potassium
KBR	Kellogg, Brown, and Root
KEPCO	Kansai Electric Power Company
kg	Kilogram
kV	Kilovolt
kVA	Kilovolt-amperes
kW	Kilowatt
kWe	Kilowatt-electric
kWh	Kilowatt-hour
kWth	Kilowatt-thermal

LAR	Liquid argon
lb	Pounds
LHV	Lower heating value
Li	Lithium
LIN	Liquid nitrogen
LNG	Liquified natural gas
LOPA	Layers of Protection Analysis
LOTO	Lock Out/Tag Out
LOX	Liquid oxygen
LP	Low pressure
LPG	Liquified petroleum gas
LRSR	Liquid redox sulfur recovery
LSTK	Lump sum turnkey
LTFT	Low-temperature Fischer-Tropsch
LTS	Low-temperature shift
MAC	Main air compressor
MATS	Mercury and Air Toxics Standards
MCH	Methylcyclohexane
MCFC	Molten carbonate fuel cell
MDEA	Methyldiethanolamine
MEA	Monoethanolamine
Mg	Magnesium
MgO	Magnesium oxide
MHI	Mitsubishi Heavy Industries
MJ	Megajoule
Mn	Manganese
ModPlant	Modular Plant
MTBE	Methyl tert-butyl ether
MTG	Methanol-to-gasoline
MTO	Methanol-to-olefins
MTR	Membrane Technology and Research, Inc.
MVA	Multi-volt amps
MW	Megawatt
MWe	Megawatt-electric
MWh	Megawatt-hour
MWth	Megawatt-thermal
N_2	Nitrogen
Na	Sodium
Na_2O	Sodium oxide
NAAQS	National Ambient Air Quality Standards
NCCC	National Carbon Capture Center
NDA	Non-disclosure agreement
NETL	National Energy Technology Laboratory
NFPA	National Fire Protection Agency
NGCC	Natural gas combined cycle
NH_3	Ammonia
NH_4Cl	Ammonium chloride
Ni	Nickel
Nm^3	Normal cubic meters
NMP	N-methyl-pyrrolidine
NO_2	Nitrogen dioxide
NO_x	Nitrogen oxide
NPDES	National Pollutant Discharge Elimination System
NSPS	New Source Performance Standards
NSR	New Source Review
NWC	Net working capital
O&M	Operation and maintenance
O_2	Oxygen
OLGA	Oil-based gas washer
OOIP	Original oil-in-place
OpenFMB	Open Field Message Bus
ORP	Oxygen reduction potential
OSHA	Occupational Safety and Health Administration
P	Phosphorous
P&ID	Piping and instrumentation diagram
Pb	Lead
PC	Pulverized coal
PCC	Point of common coupling
PFD	Process flow diagram
PHA	Process hazard analysis
PI	Process intensification
PLC	Programmable logic controllers
PM	Particulate matter
PNNL	Pacific Northwest National Laboratory
POI	Point of interconnection
PPA	Power purchase agreement
ppb	Parts per billion
ppbv	Parts per billion by volume
ppm	Parts per million
ppmv	Parts per million by volume
ppmw	Parts per million by weight
PSA	Pressure swing adsorption
PSDF	Power Systems Development Facility
psia	Pounds per square inch absolute
psig	Pounds per square inch gauge
PSM	OSHA Project Safety Management

PTSA	Pressure and temperature swing adsorption
PUC	Public utility commission
PURPA	Public Utility Regulatory Policies Act
QGESS	Quality Guidelines for Energy System Studies
R&D	Research and development
RCRA	Resource Conservation and Recovery Act
REE	Rare earth element
REMS	Radically Engineered Modular Systems
RGGI	Regional Greenhouse Gas Initiative
RH	Relative humidity
ROI	Return on investment
RRI	Rich reagent injection
RTI	Research Triangle Institute
RTO	Regional Transmission Organization
RTU	Remote terminal unit
Ru	Ruthenium
S	Sulfur
SAIC	Science Applications International Corporation
Sb	Antimony
SBCR	Slurry bubble column reactor
SCADA	Supervisory control and data acquisition
scf	Standard cubic feet
SCOT	Shell Claus off-gas treating
SCPC	Supercritical pulverized coal
SCR	Selective catalytic reduction
Se	Selenium
SEWGS	Sorbent enhanced water-gas shift
SGIP	Smart Grid Interoperability Panel
SGS	Sour-gas shift
Si	Silicon
SIS	Safety Instrumented System
SIL	Safety Integrity Level
SiO_2	Silicon dioxide
SIS	Safety Instrumented System
SMR	Steam methane reforming
SNCR	Selective non-catalytic reduction
SNG	Substitute natural gas
SO_2	Sulfur dioxide
SO_3	Sulfur trioxide
SOFC	Solid oxide fuel cell
SO_X	Sulfur oxide
SURE	Sulfur Recovery
SWGS	Sour water-gas shift
Syngas	Synthesis gas
T&C	Terms & Conditions
tpd	Tons per day
TEA	Techno-economic analysis
TEU	20-foot equivalent unit
TGT	Tail gas treatment
THM	Toluene-hydrogen-methylcyclohexane
Ti	Titanium
TiO_2	Titanium oxide
TREMP	Topsøe Recycle Energy-efficient Methanation Process
TRIG™	Transport Integrated Gasification
TRL	Technology Readiness Level
TSA	Temperature swing adsorption
TVC	Total variable cost
UL	Underwriters Laboratory
USACE	U.S. Army Corps of Engineers
V	Volts
V	Vanadium
VOC	Volatile organic compound
VSA	Vacuum swing adsorption
W	Watts
WGS	Water-gas shift
Zn	Zinc
ZnO	Zinc oxide
ZnS	Zinc sulfide

1.0 BACKGROUND: CRITICAL ELEMENTS FOR MODULAR GASIFICATION SYSTEMS

Modular gasification process plants and systems are an option for consideration by project developers, particularly those interested in the production of local electrical power and value-added products, such as hydrogen and liquid fuels. The decision about which plant or system to adopt for a given intended use is based on the careful consideration of multiple factors. Fundamentally, the system under consideration must have technical viability and potential to compete in the marketplace. Early phases of development include preliminary or conceptual designs with estimates of system performance and cost to address questions of technical and market viability. However, not all project developers have the capability or resources to readily establish preliminary design and perform techno-economic analysis (TEA) to determine system feasibility.

This handbook was developed to formalize, define, and automate the design process for developing modular gasification process plants and systems. The handbook includes detailed definitions of the elements found in these systems, defines the scope of the system/process design, and provides specific design guidelines for the various modules or unit operations that commonly figure in gasification process plants (specifically the gasifier, fuel/feedstock handling and pretreatment procedure, air/oxygen supply, syngas cleanup system, syngas conversion process, power and/or steam generation block, grid connection module, chemical conversion module, product recovery/purification systems, environmental treatment/control technologies, carbon dioxide (CO_2) capture/storage technologies, and various ancillary systems).

1.1 BREAKING THE TRADITIONAL ECONOMY-OF-SCALE PARADIGM

The following sections provide an historical overview of gasification plants in the United States, the economy-of-scale factors associated with these plants, and the benefits and opportunities for modular gasification systems/applications.

1.1.1 LARGE GASIFICATION PLANTS AND THE ECONOMY OF SCALE

The most important drivers for gasification in the United States emerged during the energy crises of the 1970s, coincident with the general environmental movement, which resulted in increased awareness of the need to reduce the environmental impacts of fossil fuel-based energy. Integrated gasification combined cycle (IGCC) electricity generation offered the possibility of higher efficiency and lower emissions than conventional coal-based power generation methods. Large-scale gasification technologies had been successfully commercialized by that time, while gas turbine and combined cycle technology was also well established (e.g., General Electric [GE] turbines). Combining gasification with a syngas-burning combustion turbine was considered a tractable engineering challenge, given that the major pieces of technology had already been advanced to high levels of technological readiness. This set the stage for the early implementations of IGCC in the United States, including the Cool Water Generating Station (California, USA), which was followed by larger-scale commercial plants like the Tampa Electric IGCC plant in Polk County, Florida (USA) and Wabash River Gasification Repowering Project (USA). Noteworthy instances of IGCC plants from the mid-1970s to recent years are shown in Figure 1-1.

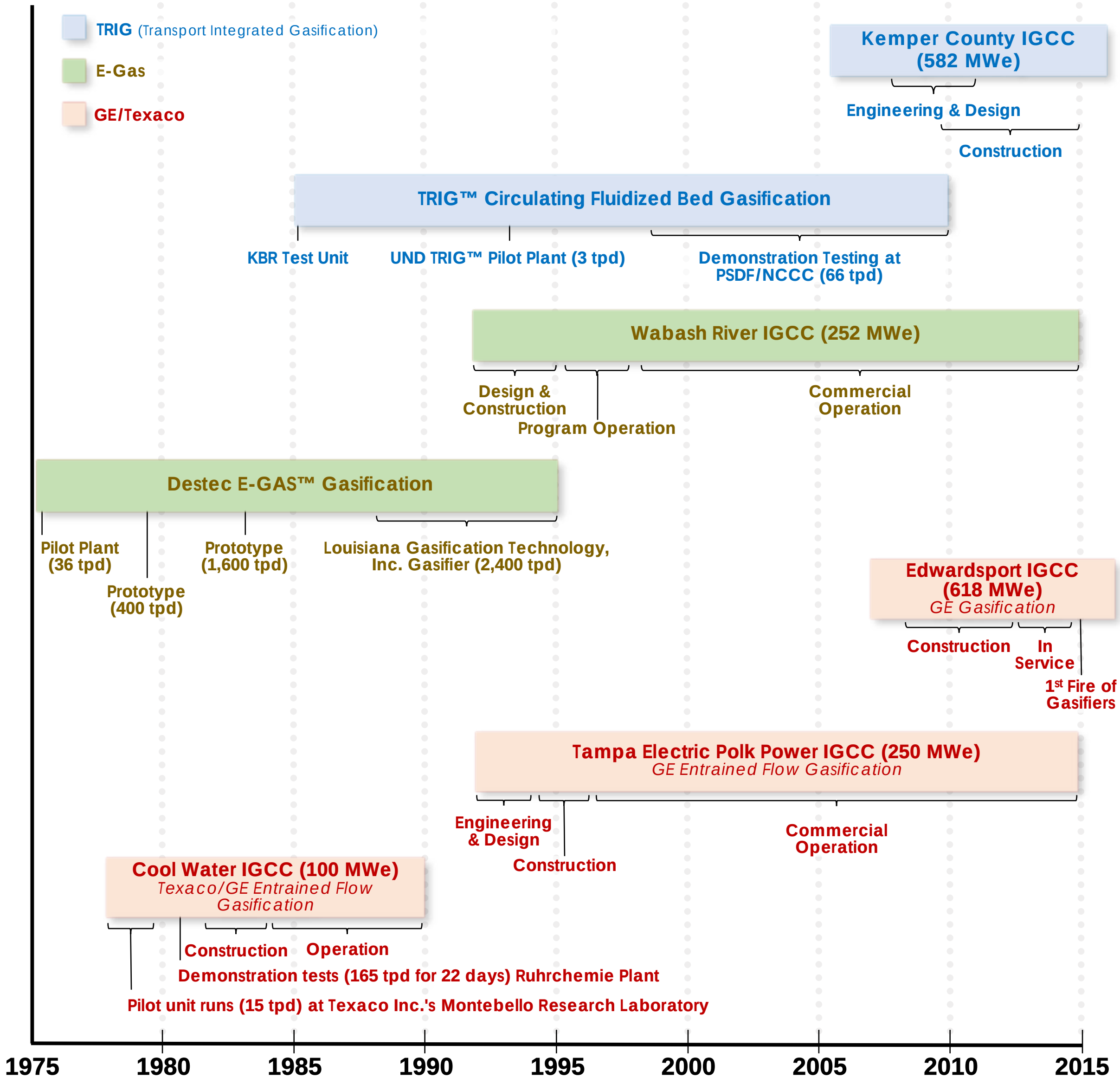


Figure 1-1. IGCC projects timeline

The figure shows a trend from smaller to larger units, irrespective of the type of gasification technology, with the newer plants over twice the size of the approximately 250-megawatt (MW) Wabash River and Tampa units. This was caused by simple economics facing power plants of this type: these plants needed to be large to take advantage of economy-of-scale factors that applied to most of the major plant systems. The sizes of the turbines, gasifiers, and conventional cryogenic air separation units (ASUs) to produce oxygen for gasification needed to be large to access the best efficiency ratings and lowest unit costs. For example, in IGCC, the steam turbine produces roughly half of the gross power generation, and the most efficient steam turbine examples are large in size (i.e., they typically require two gasification trains in parallel upstream to provide sufficient exhaust gas flow through the heat recovery steam generator (HRSG) to raise sufficient steam). As for cryogenic ASUs, they are not cost effective unless a certain size threshold is exceeded.

Experiences with the most recent large plants have proven unfavorable. Although Wabash River and Tampa (and the latter's forerunner Cool Water) were technically successful and had come in essentially on schedule and on budget, the Edwardsport IGCC plant (Indiana, USA) and Kemper County IGCC plant (Mississippi, USA) both experienced delays and incurred large budget overruns. Gasification capacity/reliability at the Edwardsport IGCC plant has been a challenge and the Kemper County IGCC plant eventually abandoned gasification and now runs as a large, natural gas combined cycle (NGCC) plant.

The causes of the schedule delays and cost escalations for these projects are well studied and documented. Many of these issues resulted from market factors, permitting impacts, and planning, wherein the project developers failed to anticipate the high costs of materials and differences between initial design and ultimate design, resulting in more expensive engineering solutions than initially anticipated. For example, the initial design/costing for Edwardsport IGCC plant was heavily reliant on the Tampa Electric IGCC plant (Polk County, Florida, USA) design and cost experience, which resulted in major underestimates of both cost and requirements in the much larger units of the Edwardsport IGCC plant relative to those of the Tampa Electric IGCC plant. In the case of the Kemper County IGCC plant, the scale-up from Kellogg, Brown, and Root's (KBR) Transport Integrated Gasification (TRIG™) gasifier[1] size to the Power Systems Development Facility (PSDF) TRIG™ units was so ambitious (factor of 200 larger in terms of throughput) that it is a likely root cause for the nearly three-fold increase in plant cost over the initial estimate.

These IGCC experiences resulted in the realization that a traditional, economy of scale driven gasification project is burdened with significant cost and investment risk, which is difficult to justify to utilities, customers, and financial institutions.

1.1.2 BENEFITS OF MODULARITY

The solution to many of these issues is to adopt gasification technology in smaller scale or modular systems, reducing absolute levels of cost and risk. This concept is the rationale behind the current focus on gasification technology development by the U.S. Department of Energy's National Energy Technology Laboratory (DOE/NETL). This approach has many benefits, including the following:

Mass Production Advantage—Small, modular reactors and plants with common components could be quickly mass-produced, prefabricated, and dropped in at a desired site, which reduces build and maintenance costs compared to constructing a traditional power or chemical plant.

Availability and Redundancy—Modular plants with multiple parallel trains have higher availability, because equipment failure only slows production, rather than halting it, as is often the case in conventional plants with limited equipment redundancy.

Flexibility in Siting and Market—The "right sizing" of plants to local markets by coupling several small, prefabricated, modular reactors or units can allow one to service markets that were previously too small to be economically attractive for large, traditional plants.

Enabling Opportunity Fuel Use—Smaller systems and distributed plants could enable the utilization of a higher proportion of opportunity fuels like waste plastics, municipal solid waste (MSW), and local biomass.

Enabling Innovative Technologies—Modular systems may include innovative gasification and conversion technologies, including those with increased ability to accept marginal feedstocks (as noted above). Traditional technology development involves time-consuming scale-up over many orders of magnitude, but small-scale modular systems should allow faster development and reduced costs of development, which translate into lower financial risks, especially for first-of-a-kind systems.

The flexibility in gasification system process configuration and capability, ability to site a system close to a variety of feedstocks, opportunity to tailor the system to local markets, and modest capital demand to construct modular systems all create significant opportunities for modular gasification. These advantages provide a route to viable commercialization of smaller-scale gasification applications without relying on traditional economy of scale.

1.2 MODULAR PROCESS DESIGN, FABRICATION/ CONSTRUCTION, AND TRANSPORTATION COST

In attempting to apply modularization principles to process facilities like gasification-based plants, the primary consideration is modular fabrication and construction, which involves fabricating preassembled equipment blocks or modules at a factory, transporting them to the plant site, and completing final interconnection of the modules in the field. Most of the assembly and fabrication work is conducted at the factory under controlled conditions, requiring only minimal plant site construction effort compared to field-erected (i.e., stick-built) structures. There is great flexibility in terms of the modules, which may comprise one conventional unit operation, several unit operations, or only part of a unit operation depending on plant size and complexity. The literature provides numerous examples of modular construction in the process industries, including catalytic cracking units and steam methane reformers, suggesting that modular design and construction for gasification-based plants is feasible.

The process design for a facility constructed of modules ensures that the functionality of the process is the same as that of a conventional facility. Connections for piping, tubing, wiring, and control are all completed at the module level. Modules communicate via junction boxes for power and control wiring and instrument air is fed from a header. In many cases, the pipe rack is integrated into the module, with the top portion of the module carrying the ducts and piping.

The detailed engineering and mechanical design of each module should consider the module's weight and dimensions, which are constrained by the intended transportation method and size of the installation location, the necessary strength of the frame materials to support the equipment weight, and the need for a foundation. Transportation options typically include road, rail, and barge. For road transportation, module size is most often constrained by the skid size accommodated by conventional tractor/trailers. Rail and barge-transported modules may allow for larger module sizes.

Modular construction offers several advantages that can outweigh potential (or perceived) increases in upfront cost, including the following:

Enhanced Quality Control and Safety—Assembly in a factory setting allows for the control of factors that affect product quality and employee safety. Shop personnel can work in an environment with a controlled flow of personnel and materials, as well as access to overhead cranes. Personnel remain in a relatively small area and work is brought to their workstations. Workers become more familiar with this limited environment, including locations of tools, industrial gases, and electrical supply points needed to perform their jobs. Personnel also become more familiar with their coworkers and develop working relationships that promote safety and efficiency. In contrast, an employee in the field must move around a construction site or plant environment and move necessary equipment. Elevated work on scaffolds and work with mobile cranes are common and carry inherent safety risks. Personnel may also be subject to inclement weather conditions that could increase risk. Teams are frequently assembled for each individual job, hindering workforce cohesion.

Capital Cost Savings—Modular units are less capital-intensive largely due to more efficient labor. For many projects, large portions of the capital costs are related to welding. A welder in a shop is considerably more efficient than a welder in the field, because a shop welder can remain in the same location and spend a higher portion of their work time doing the actual welding. The cost of field fabrication is typically at least twice that of shop fabrication and could increase by up to a factor of six, depending on the skill level and availability of the workers.

Material Cost Savings—Modules can require extra steel, but in most cases, the cost is offset by eliminating the need for a foundation. Pipe racks built into the modules also help to minimize costs by reducing the amount of piping required at the plant level.

Deployment Cost and Project Timeline Savings—Assembling modules requires fewer hours of skilled labor than building a plant in the field. The amount of work that can be performed in the shop relative to the field varies from project to project, but shifting most of the work into the shop can be beneficial, especially when qualified labor and expertise are in short supply. In the case of modular construction of a refinery plant, approximately 65% of the labor is performed in the shop. Considering the labor savings and the economics of these initial stages, the total cost of a project can be reduced by approximately 30%. Projects that employ extensive modularization can be completed faster. For example, a project that would typically take three years can be completed in around two years, enabling an early startup, reducing overhead costs, and shortening the financing period.

Challenges of Modular Construction—Transportation is one of the challenges of modular construction. As module size increases, so do the logistic difficulties of delivering modules to the plant site. Modules must be shippable (i.e., road and logistics constraints) and withstand travel rigors. They must be maneuvered and fit into place at the field site. A shipping study is usually necessary to ensure delivery and installation feasibility. Transporting modules can increase engineering and shipping costs but, in most cases, these challenges and costs are more than offset by the advantages gained.

Modules may also require more upfront engineering than conventional unit operations. Modules that are ordered from external fabricators typically require highly detailed design packages. Generally, effective module design does require more planning and early engineering and procurement than onsite development. For modular systems to succeed, the owner and contractor must be united from the start of all design work and owner operations personnel must be involved as early as possible. Modularization requires full and early commitment from all stakeholders to ensure appropriate business planning, scope definition, engineering, and procurement.[2, 3]

1.3 STANDARDIZATION OF COMPONENTS WITHIN AND ACROSS MODULES

Process modules and their corresponding unit operations have distinctly varied functions within the overall gasification-based process system. Although the chemical and physical transformations occurring in different process streams in various reaction vessels and equipment items handling process streams are varied, all modules generally have similar process control needs, features, and functionality managed by the control systems, which can (in theory) be implemented via similar kinds of sensors, actuators, and valves. These can be standardized to streamline multiple aspects of systems engineering, application, and operations and maintenance (O&M) of these components within and across plant modules.

When a new non-modular, large-scale plant is built, it traditionally has a bespoke, plant-wide central control system. Also, in engineering and design, there is a strong opportunity to standardize many components to minimize required spares and streamline future maintenance of all plant systems. In the case of modular plants, process modules are very likely to be built by different vendors, with the possibility of a lack of standardization among modules and questions about how plant-wide control is accomplished. Any such disadvantages of plants built from separate modules can be greatly reduced by planning, consistent specifications, and following certain principles, including module system boundaries and equipment component standardization.

Module system boundaries should be chosen so that they have as few external interfaces as possible. To achieve this, functions must be integrated instead of being added by external periphery. This reduces the number of module interface elements. Each module forms a self-contained, autonomous unit, minimizes the complications arising from cross-module interfaces.[4] This leads to the concept of a completely modular design, where the various plant modules act with full automation and autonomously, with communication existing only via interfaces and communication protocols.[5] This contrasts with traditional centralized control where no modules are automatic, and all control is from a central hub.

Equipment component standardization can reduce equipment spares inventories and O&M training requirements and procedures, saving expenses over the lifetime of the plant. The idea is to standardize equipment items throughout the plant as much as possible to reduce the number of different kinds of equipment items and parts that need to be maintained and for which the staff need to be trained to service. This does require time and effort up front to establish and enforce standards, but it results in ongoing reduction of O&M costs.

In gasification plants, the obvious targets for equipment standardization are common items, such as motors, pumps, variable frequency drives, gearboxes, valves, and programmable logic controllers (PLCs). Rather than sourcing these items from many different manufacturers (requiring a multitude of vendor-specific spares, manuals, and troubleshooting procedures), sourcing from a limited number of manufacturers/vendors (possibly through a single capable and trusted supplier) can simplify inventories and procedures. This approach can extend to the size and specifications of the equipment items themselves. For example, rather than inventory spares for different sizes of motors/drives, there is benefit in standardizing to a single, larger drive in each frame size (and/or the size most carried by distributors), greatly reducing inventory on the shelves and facilitating restocking. Similarly, more durable materials can be set as the standard for equipment (e.g., stainless-steel shafts in rotating equipment versus carbon steel). Larger, standardized equipment items with better materials tend to be more expensive initially, but the simplicity in ongoing repairs and maintenance result in cost savings. Another example of equipment standardization recommended is to adopt interchangeable International Organization for Standardization (ISO) dimension valves, solenoids, and relays to be able to use any ISO-compliant manufacturer's product (if needed). In the area of controls hardware, it is recommended to standardize a single vendor/brand and format of PLCs to reduce spares and limit training requirements for controls systems O&M (Schindler, 2011).[6]

1.4 PLUG-AND-PLAY DESIGN FOR MULTIPLE TECHNOLOGY BLOCK MODULES

One recommended approach for improving plant and process performance and control is to require vendor-uniform interfaces between control systems and process modules, which enables modularization of control functions and avoids the issue of forcing control system manufacturers to assure compatibility. As a result, process modules from any manufacturer could be obtained and integrated into the control system in a "plug-and-play" approach, with the process module functionally independent from the control system type.[6] However, doing so is challenging, because it requires the cooperation of all vendors/manufacturers to standardize interfaces and could try the technical or economic limits of the manufacturers.[7]

More generally, the principle of "plug-and-play" for a process plant is the idea that the overall plant process may be modularized by dividing a complete plant into functional units. The functional units or modules each perform a subprocess and, within each subprocess, mechanical and automation components are devised to enable standalone operation. These functional units or modules are combined to produce specific process plants, which can be extended almost indefinitely by adding modules, enabling immediate adaptation to market and production requirements. The capacity is increased by adding units instead of scaling up individual ones. The modules have a requirement for standalone operation, so they can be relatively easily "plugged in" or grouped or sequenced in the plant for desired output without the traditional need for a highly centralized and controlled plant design/engineering and operation approach.[8]

The concept for flexible, process-encapsulated modules for process plants has attracted interest from specialty chemical and pharmaceuticals industries, where it is thought that the process module plug-and-play approach can enable faster time to market, energy savings, lower operating costs, and less capital expenditure.[9] Modular gasification plants are expected to share many characteristics with smaller chemical plants, so the same principles should apply.

1.5 UTILIZATION OF EXISTING INFRASTRUCTURE TO THE MAXIMUM EXTENT POSSIBLE

The construction and operation of a process plant require the availability of multiple types of infrastructure. Plant construction and operation costs should be lower if every opportunity to leverage existing infrastructure relevant to the plant and its site is taken, and the higher costs of building that infrastructure are avoided.

Infrastructure (as commonly defined) is a part of the set of factors that warrant consideration in the evaluation and selection of plant sites. Some of these factors are as follows,[10] with discussion of how they pertain to infrastructure and possibilities for leveraging existing infrastructure and associated savings.

Raw Materials—Raw material staged onsite (e.g., fuels, catalysts, and chemical reagents) requires infrastructure for storage and handling. Existing areas permitted for or amenable to storage and handling of bulk raw materials of these kinds should be used whenever and wherever possible.

Markets—Markets for products and byproducts are needed (ideally in close proximity), while feedstocks and supplies will be most conveniently obtained through locally sourced markets. Although the market is not infrastructure per se, an existing local market able to cost effectively meet supply needs and take products is a valuable asset to leverage.

Energy Availability—Gasification plants require electrical power and steam is likely to be needed in quantity for most plants of this type. Existing power supply lines and transformers should be used if available and sized appropriately for the gasification plant. Steam can be generated through a unit operation in the plant, but if an existing steam generation plant happens to be onsite (or in proximity) it can be leveraged for cost savings and possible economies of scale.

Climate—Plants located in cold climates may need protective shelters around process equipment, while special cooling towers or air conditioning equipment may be required in hot climates. Plants located or engineered to take advantage of any such existing equipment or infrastructure would benefit.

Transportation Facilities—Water routes, railroads, and highways are the essential transportation infrastructure of concern at a plant. Kinds and amounts of products and raw materials that need to be moved tend to dictate the most suitable transportation types. Existing rail spurs, waterways/canals, and barge dock would greatly save on costs of building those if water or rail transportation is needed. If not, access roads and highways suitable to receiving the required truck traffic must be in place or constructed. The high capital and permitting costs of road building should be avoided by utilizing existing road infrastructure (if possible).

Water Supply—Gasification plants use water for cooling, washing, steam generation, and as a feedstock. The plant must have access to a dependable water supply of the required quantity. Existing access to a river or lake water supply (or wells if water requirement can be accommodated by well sourcing) with piping, pumps, and storage tanks is highly desirable and would save the otherwise inevitable cost of the water supply infrastructure on the plant site and from the water source to the site. There is the additional benefit of possible existing permitting on any water supply system already in place/operation, which may help alleviate the burden of water permitting.

Waste Disposal—Adequate capacity and facilities for correct waste disposal are required for process plants. Whether they are onsite or offsite, leveraging use of facilities offering low-cost solid waste disposal and water treatment is beneficial to project economics.

Site Characteristics—Land characteristics at the plant site have important effects on construction costs. If the site for the plant in question has pre-existing grading, preparation, or other improvements along these lines, it is favorable for the project moving forward. Also, pre-existing availability of space at the site for possible expansion is a plus.

Flood and Fire Protection—Existing community infrastructure in the form of emergency services and fire protection is beneficial to the proposed project if they can be leveraged with cost savings, because such services must be made available at any cost.

Community Factors—Existing community infrastructure in the form of varied housing, cultural facilities, and recreational opportunities is beneficial in enabling quality of life for plant personnel. In the absence of this infrastructure, project developers may be obligated to provide living quarters (which can be a significant burden), while the lack of these intangible resources can be difficult to overcome. Planning a project to take advantage of such community factors is good practice.

1.6 MINIMIZING COMPONENT STEPS AND REDUCING EQUIPMENT SIZE

Technological innovations are expected to play an important role in supporting viable modular gasification systems development. They help to streamline component development and design, enable inexpensive fabrication of equipment components, identify and advance new high-performing materials for gasification applications, and enable design and development of overall systems with a high degree of integration and performance.

Simulation-Based Reactor Optimization—Massive computing power at lower costs is expected to allow increasingly realistic computer models to simulate particle behavior during fuel conversion processes. Leveraging this innovation, simulation-based reactor design and optimization will be used to better predict reactor behavior and results and advanced manufacturing techniques (see below) will enable rapid prototyping of concepts. This approach should lead to reduced times for new reactor development. Sophisticated modeling using optimization-based computational fluid dynamics (CFD) and kinetics studies, along with the generation of validation data, will be used to develop tools to provide accurate predictions of spatial reaction behavior and allow researchers to virtually test feed- and product-specific reactor designs. This should lead to unprecedented new reactor configurations that would allow manipulation of pyrolysis, gasification, and volatile cracking in different regions of the reactor—or allow combining of conversion and purification steps into one process, greatly increasing control and process efficiency.

Process Simulation and Cost Analysis—Process simulation and cost analysis help to determine which breakthrough technologies have the potential to thrive in the commercial market and bring lasting benefit. The integration of systems engineering and analysis of energy processes, infrastructure, markets, energy process simulation, and life cycle environmental impact analyses can enable more effective design of advanced energy systems. Both costs and environmental impacts analyses are needed to determine full life cycle environmental footprint and the potential of advanced energy systems in the commercial marketplace.

Reaction Intensification and Novel Reactor Innovation—Improved catalysis using nanostructured catalysts for intensifying and finely controlling reaction chemistry, more carefully controlled heat transfer and management, and use of non-traditional energy sources (e.g., plasmas and microwaves) are some elements enabling reaction intensification. These could contribute to increased reactor efficiency with reduced capital expenses, particularly at smaller, modular scales.

Advanced Manufacturing and Reactor Materials Development—Breakthroughs in advanced manufacturing can help to reduce the capital costs for small-scale reactors and modular plants. Rapid prototyping, such as inexpensive, 3D plastics printing of cold flow reactors, can be used to quickly test unique reactor designs. Additive manufacturing and other advanced manufacturing techniques for metals and ceramics are used to rapidly translate proven cold-flow unique reactor designs into reactors for hot-flow validation and final manufacturing of reactors for deployment. Additive manufacturing can also be used to fabricate some system components, such as gasifier feed injectors, enabling more complex designs at substantially lower costs, and faster manufacturing cycle times.

1.7 PERFORMANCE TRADE-OFF CONSIDERATIONS WITH OVERALL COST AND SCALE REDUCTIONS

The principle of reaction intensification—in which existing processing schemes are reworked into ones that are both more precise and more efficient—includes the following possible benefits:

- Smaller equipment.
- Reduced number of process steps.
- Reduced plant size and complexity.
- Modularity cost reduction may compensate for lost economy of scale.
- Reduced feedstock consumption.
- Reduced pollution, energy use, and capital and operating costs.

The ideas of reduced number of process steps and reduced complexity associated with process intensification are illustrated in Figure 1-2.

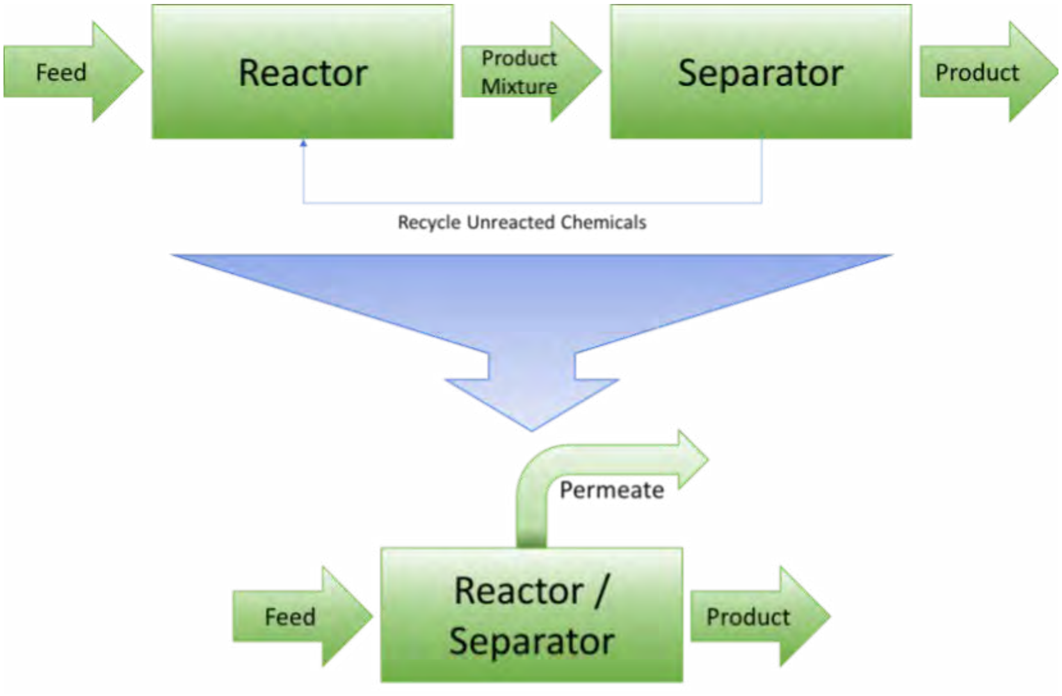


Figure 1-2. Process intensification

Ideally, the smaller equipment and reduced process complexity translates directly into cost reductions. In practice, moving from established technology to a process intensified scheme requires considerably more sophisticated technology (e.g., reactor, integrated separations) to realize the simplified schemes, resulting in potential cost tradeoffs that should be carefully evaluated in project development.

It is also possible that smaller-scale, modular processes may face tradeoffs in the area of product quality or the range of suitable product application (e.g., a single-stage modular process, even while utilizing some advanced new technology, may still have to sacrifice on the product specification that would otherwise be obtainable with a traditional multi-stage process). Again, the product value/quality tradeoff should be carefully evaluated in project development.

Sometimes there are no tradeoffs from different scales in processes in certain key areas. One example is the reactor itself, which is often a point of concern in scaleup. However, if the gasification reactor is a packed-bed reactor with a heat-transfer jacket, its tube diameter and length could be defined at development scale (pilot scale) and the pilot plant could vet this fixed tube diameter and length with a single large tube. The full-size production scale might require 10 tubes (maybe in a module) or 100 tubes (maybe two larger 50-tube modules or 10 10-tube modules), but no matter the scaled-up aggregation, each individual reactor tube should function like the one in the pilot plant, which would give high confidence to the anticipated reactor performance at any range of scales.[11, 12]

1.8 ENGINEERING, PROCUREMENT, AND CONSTRUCTION PARTNERING AND CONTRACTING

Engineering, procurement, and construction (EPC) contracts are commonly used by project developers to undertake large-scale and complex infrastructure or plant projects. They are entered into by a project developer and a contractor. Under the contract terms, the contractor is obligated to deliver a complete facility to the project developer, who need only "turn a key" to start operating the facility (from which the terminology of turnkey construction contract, or lump sum turnkey contract, is derived). EPC agreements are complex, but, at the most basic level, they obligate the contractor to deliver the plant or facility for a guaranteed price by a guaranteed date, and the plant/facility must perform according to specifications. Failure to comply with contract requirements usually results in the contractor incurring monetary liabilities. The costs of turnkey contracts and the inherent risks behind them can be reduced by choosing modules that have already been commercially demonstrated. Further reduction of costs is possible if the module vendors are willing to share in the performance guarantee, which is particularly beneficial for first-of-a-kind plants/modules. The EPC contractor coordinates all design, procurement, construction, and commissioning work and ensures that the entire project is completed as required and on time.

Gasification-based plants/facilities are relatively complex, combining aspects of chemical process facilities and, if electricity is to be generated, the aspects of a power generation plant. It is probable that most project developers contemplating a gasification facility will choose to partner with an experienced engineering firm through an EPC contract to design, construct, and commission their facility. The EPC lead contractor (often an architect-engineering firm with experience in energy and chemical process fields) will manage the work and execute designs, but typically subcontract many aspects of the plant development and construction to construction and specialty companies. All the details of the engineering and construction plans and project development costs will be spelled out in the proposals that prospective lead contractors will submit for consideration to the project developer.

Entering into EPC contracts and partnering through facility development works best when the project developer and potential contractors interact with ample communication from the earliest planning stages and maintain flexibility as the engineering and construction move forward.[13] This may include involving engineering firms prior to the detailed design of a gasification-based facility/plant during the concept phase or front-end engineering design (FEED) phase of a project. Decisions affecting plant design and cost are usually made well in advance of detailed design of a facility. Changes after completion of these initial design phases usually add cost and schedule overruns to the overall project. Ideally, the modular construction approach eliminates the majority of the chemical and mechanical engineering required for each project, since modules are “stamped” out one after the other to increase the capacity of the facility and the amount of engineering required for detailed design is limited to site-specific issues and site-specific engineering disciplines. It is recommended that project developers reach out in advance of the issuance of their formal Request for Proposals (RFPs) to potential bidders to help define project specifications and maintain communication throughout the project development to work through the many issues that arise during the design, building, and commissioning phases of the project.

1.9 MAINTENANCE NEEDS, INCLUDING ACCESS TO EQUIPMENT AND INSTRUMENTS

Best practices for the maintenance of process plants have changed as technology enabling more sophisticated approaches has advanced. Traditionally, maintenance took the reactive/corrective approach where the equipment or instrument was fixed only when it broke. Preventive maintenance is a more advanced approach, where maintenance is provided on a routine schedule based on whatever information is available on the life expectancy of the class of the equipment or system assets in question. Preventive maintenance can save 10 to 15% in down time and repair costs compared to reactive maintenance. However, if preventative maintenance is based on less than complete and accurate information, it is possible that time and resources will be spent to inspect/clean/service equipment that is otherwise fully functional and does not need maintenance.

A newer approach is predictive maintenance, which is a continuous process based on the current condition of equipment (as opposed to the scheduled intervals of preventative maintenance). Predictive maintenance is made possible through the increasing availability of technology, such as smart sensors, data analytics, the industrial Internet of Things (IoT), and machine learning (ML). Predictive maintenance leverages increased data and analytics availability to precisely determine maintenance needs and wisely use the available maintenance budget.[14, 15] As the project developer considers proceeding with a process plant, it is beneficial to be aware of the options for maintenance best practices. The project developer should strike an appropriate balance on systems capabilities to support the most cost-effective maintenance approach for the plant or process unit.

One of the most important aspects affecting plant maintenance is proper plant layout. A well-designed layout is important in many ways; in terms of maintenance, it provides safe, convenient access for the maintenance of items and the removal or in situ repair of components or process equipment.[16] The whole process of determining plant/systems layout is relatively complex and considers many factors other than facilitation of maintenance. When access is available for vehicles and personnel during construction, operation, maintenance, and emergencies, and adequate separation around plant units and equipment items is in place to ensure proper hazard containment, many plant specifications will have been met (e.g., safety, operational, regulatory), in addition to making provisions for ease of maintenance.

1.10 PRACTICAL CONSIDERATIONS OF EQUIPMENT AND PLANT LAYOUT FOR IMPROVED MAINTENANCE

Project developers should consider the following to help ensure simple, fast, and uncomplicated access to individual items of the plant and equipment to perform maintenance and improve plant availability and productivity.[17]

Equipment Positioning Priority—Equipment should be arranged in such a way to minimize the need to remove any such equipment that is properly operating to access the part needing repair. Careful thought should go into optimal equipment arrangements to best accommodate expected maintenance and make access easier for more frequent jobs.

Attention to Electric Motors and Mechanical Drives—Lifting access to motors is needed for their almost inevitable replacement over the lifetime of a plant. Maintenance access for pump set alignment, adjustment, and repairs is likely needed several times over their lifetime. Drive belts, chains, pulleys, and sprockets wear out and need replacing, so space is required around such equipment to remove the guard or cover over the drive, mount the pulley or sprocket puller, and remove the sprocket and/or pulley from the shaft. Motors on equipment with shaft seals, packed glands, or mechanical seals need to be mounted for quick access to those wearing parts.

Safe Access to Equipment—Safe access to any equipment or unit that may need maintenance should be considered in the design and construction to avoid hazardous and troublesome situations (e.g., makeshift scaffolding) as workers try to reach elevated or otherwise inaccessible places in the plant.

Cranage and Lifting—Cranes or forklifts are needed to move heavy equipment and, if such heavy items’ lift points are located away from the crane, the required lift capacity is much greater and incurs more expense. Project designers should consider placing any items that need lifting with close, convenient access to a crane or forklift.

Locating Connections and Instruments—Flanges, valves, assemblies, and all electrical or instrument items must be accessible for repair, which is difficult if they are located in awkward or dirty places (e.g., underneath equipment) or if there is any other type of reduced access or insufficient space. Design features, such as extended nozzles, may help so that valves and instruments can be fitted beside tanks and vessels. In all cases, adequate space and clearance should be provided to give the tradesman quick, easy, and safe access.

Ergonomic Working Around Equipment—Repairs involve people getting their bodies to every location around the piece of equipment to be fixed, while maintainers and operators must get tools into position to make alignments and adjustments. Electricians have to fault-find equipment. If the freedom of access to do these jobs is restricted, the jobs may be difficult and poorly completed.

Maintenance Access Over Time—Once a heavy item is mounted in the plant, it typically remains in place. If neighboring equipment is later added without regard to the access originally established, this likely results in significant losses of availability and high costs of disassembly to make repairs.

Overall, the best design principles for later ease of maintenance are based on building in safe areas with plentiful access to equipment and preserving space for personnel to complete maintenance.

1.11 SUMMARY

Modern use of gasification in the United States was mainly spurred by the need for higher efficiency and less polluting power generation from coal. The primary focus in the late 20th century was large IGCC electricity generation plants, which took advantage of economies of scale and were successful at the time. Recently, large gasification-based plants have been less cost effective. The adoption of modular gasification technology in smaller scale, using advanced technologies, opportunity feedstocks, and accessing niche markets should reduce costs and risks, enabling a route to commercialization of gasification applications.

Various strategies help the viability of modular gasification. Modular fabrication and construction, standardization of equipment units and control systems, and use of “plug and play” process modules can greatly simplify design, reduce unit construction costs, and enable assembling plants from relatively low cost, interchangeable, and standardized process building blocks. Siting of modular gasification sites, taking advantage of favorable site characteristics and infrastructure elements and considerations, can enable a successful market implementation. Advanced technologies, such as modeling and simulation, process intensification, and advanced manufacturing methods and materials, can be leveraged for higher performance and lower unit costs of the plant. In various aspects of engineering and design of the plant, procurement, and construction, best practices for partnering and contracting result in faster and cheaper establishment of the operable plant, while careful attention to ongoing needs of plant O&M in planning and engineering features of equipment and plant layout yields significant benefits for the plant over its entire life cycle.

Endnotes

1 NETL (undated) “KBR TRANSPORT GASIFIERS”, *https://www.netl.doe.gov/research/coal/energy-systems/gasification/gasifipedia/kbr*.

2 M. Baldea, T.F. Edgar, B. Stanley, A.A. Kiss (2018) “Modularization in Chemical Processing”, AIChE, *https://www.aiche.org/resources/publications/cep/2018/march/modularization-chemical-processing*.

3 C. Rentschler, M. Mulrooney, G. Shahani (2016) “Modularization: The Key to Success in Today’s Market”, Hydrocarbon Processing, *https://www.hydrocarbonprocessing.com/magazine/2016/december-2016/special-focus-plant-design-engineering-and-construction/modularization-the-key-to-success-in-today-s-market*.

4 J. Friedrich, S. Scheifele, A. Verl, A. Lechler, (2015) “Flexible and Modular Control and Manufacturing System”, 9th CIRP Conference on Intelligent Computation in Manufacturing Engineering - CIRP ICME ‘14, Procedia CIRP 33:115-120.

5 T. Bieringer, C. Bramsiepe, S. Brand, A. Brodhagen, C. Dreiser, C. Fleischer-Trebes, N. Kockmann, S. Lier, D. Schmalz, C. Schwede, A. Schweiger, F. Stenger (2016) “Modular Plants—Flexible chemical production by modularization and standardization – status quo and future trends”, DECHEMA, December 2016, Frankfurt am Main, Germany.

6 B. Schindler (November 7, 2011) “How to Reap the Benefits of Parts Standardization”, Reliable Plant, Noria Corporation, *http://www.reliableplant.com/Read/28608/benefits-parts-standardization*.

7 ZVEI: Die Elektroindustrie (2015) “Module-Based Production in the Process Industry – Effects on Automation in the “Industrie 4.0” Environment, Recommendations of the Modular Automation Working Group Following Namur Recommendation NE 148”, ZVEI - German Electrical and Electronic Manufacturers’ Association, *https://dam-mdc.phoenixcontact.com/asset/156443151564/124fe9b2921442779803e4b98b85d10f/ZVEI-White-Paper-Modulare-Automation_EN.pdf?asis=true*.

8 N. Franz (2015) “Plants based on the Lego principle-Modular automation with valve terminals”, Festo AG & Co., *https://www.festo.com/net/SupportPortal/Files/374863/WhitePaper_Ventilinseln_en_V02_L.pdf*.

9 Hannover Messe (February 15, 2017) “Modular Production: Plug & Play in Chemical Plants”, Hannover Messe, *https://www.hannovermesse.de/en/news/news-articles/modular-production-plug-play-in-chemical-plants*.

10 K.D. Timmerhaus, M.S. Peters, R.E. West (2003) Plant Design and Economics for Chemical Engineers, 5th edition, McGraw-Hill Book Company.

11 S. Benbrook (January 12, 2018) “Reduce Pilot Plant Capital Costs”, Chemical Processing, *https://www.chemicalprocessing.com/articles/2018/reduce-pilot-plant-capital-costs/*.

12 D. Ozokwelu (September 2014) “High Efficiency Modular Chemical Processes (HEMCP)”, Modular Process Intensification, Framework for R&D Targets, DOE, *https://www.energy.gov/sites/prod/files/2014/10/f18/hemcp-topic-overview.pdf*.

13 R. Walton (December 17, 2019) “EPC Best Practices: Planning, Flexibility and Communication”, Power Engineering, *https://www.power-eng.com/2019/12/17/epc-best-practices-planning-flexibility-and-communication/*.

14 Chemical Engineering (July 1, 2018) “The Changing Face of Maintenance”, Chemical Engineering, *http://www.chemengonline.com/changing-face-maintenance-2/*.

15 Inspectioneering (n.d.) “Refinery & Plant Maintenance”, Inspectioneering Journal, *https://inspectioneering.com/tag/maintenance*.

16 S. Moran (December 1, 2016) “Process Plant Layout — Becoming A Lost Art?”, Chemical Engineering, *https://www.chemengonline.com/process-plant-layout-becoming-lost-art/*.

17 M. Sondalini (n.d.) “Equipment and Plant Layout for Improved Maintenance: 7 Consideration for Equipment and Plant Layout for Improved Maintenance”, Accendo Reliability, *https://accendoreliability.com/equipment-plant-layout-improved-maintenance/*.

2.0 SCOPE DEFINITION

2.1 INTRODUCTION

Chapter 2 provides an overview of the many design concepts and considerations the user will encounter when proceeding with the design of the gasification system, as well as introduces the user to the general design philosophy and procedure for the two general classes of gasification-based systems. These concepts are further explored in later chapters.

Gasification is a robust, proven, and versatile technology that produces syngas (carbon monoxide [CO] + hydrogen [H_2]) from reactions between carbon, oxygen, and steam at low to very high pressure (up to 1,500 pounds per square inch gauge [psig]) and temperatures ranging from 1,250°F to more than 2,600°F. Gasification is compatible with many different applications that fall within two major categories: power generation and chemical/fuel production. Figure 2-1 shows a partial list of potential applications for gasification, all of which fall into two major categories: power generation and chemical/fuel production. In addition to fuel cells, note that the hydrogen pathway can be used to produce ammonia. The primary competition for products from gasification is syngas production from natural gas.

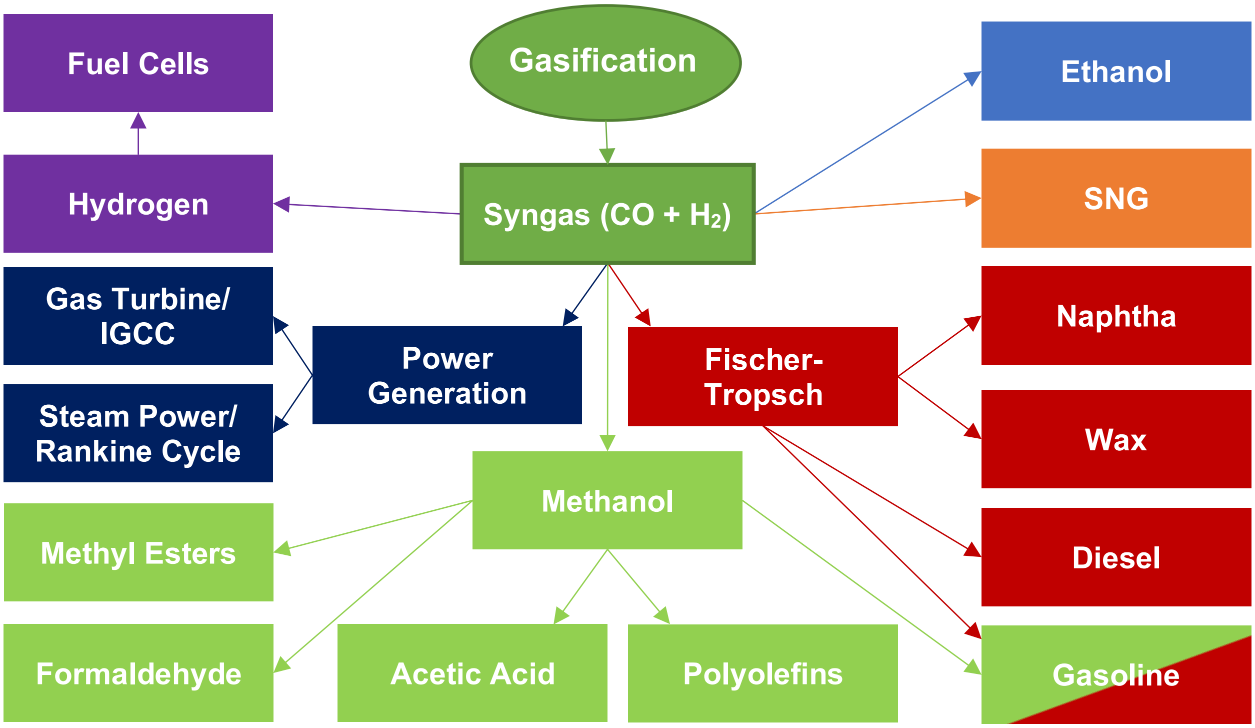


Figure 2-1. A partial list of the many uses for gasification

It is also important to note that many of the applications shown above can coexist; for example, it is possible to construct an integrated gasification combined cycle (IGCC) plant where some of the syngas is bled from the main fuel stream and used to make hydrogen for fuel cells. In some chemical/fuel production plants, the complete conversion of syngas to a chemical or fuel is not practical, and the remaining syngas may be combusted for power generation as an additional revenue stream—albeit for additional capital investment. The remainder of this chapter is dedicated to introducing the user to both general “paths” for the gasification-derived syngas, the general design philosophy behind each “path,” and some of the major design decisions to consider.

2.2 END-PRODUCT SELECTION AND GENERAL DESIGN PROCEDURE

The first (and most important) decision that the user must make when designing a gasification plant is the desired end product(s). Is the plant going to produce electricity, chemicals, or both? The standard design procedure changes depending on the answer to this question. A basic flowchart of both procedures in shown in Figure 2-2. In many cases, restrictions on the plant may limit or completely define certain design aspects before the design process begins. For example, in circumstances where the plant location is known, the feedstock may already be determined. The design procedure given in Figure 2-2 is intended as a "blank-slate" process to guide new designs toward their eventual goals. In reality, plant designing is an iterative process that may require revisiting certain steps multiple times during the planning and simulation phases.

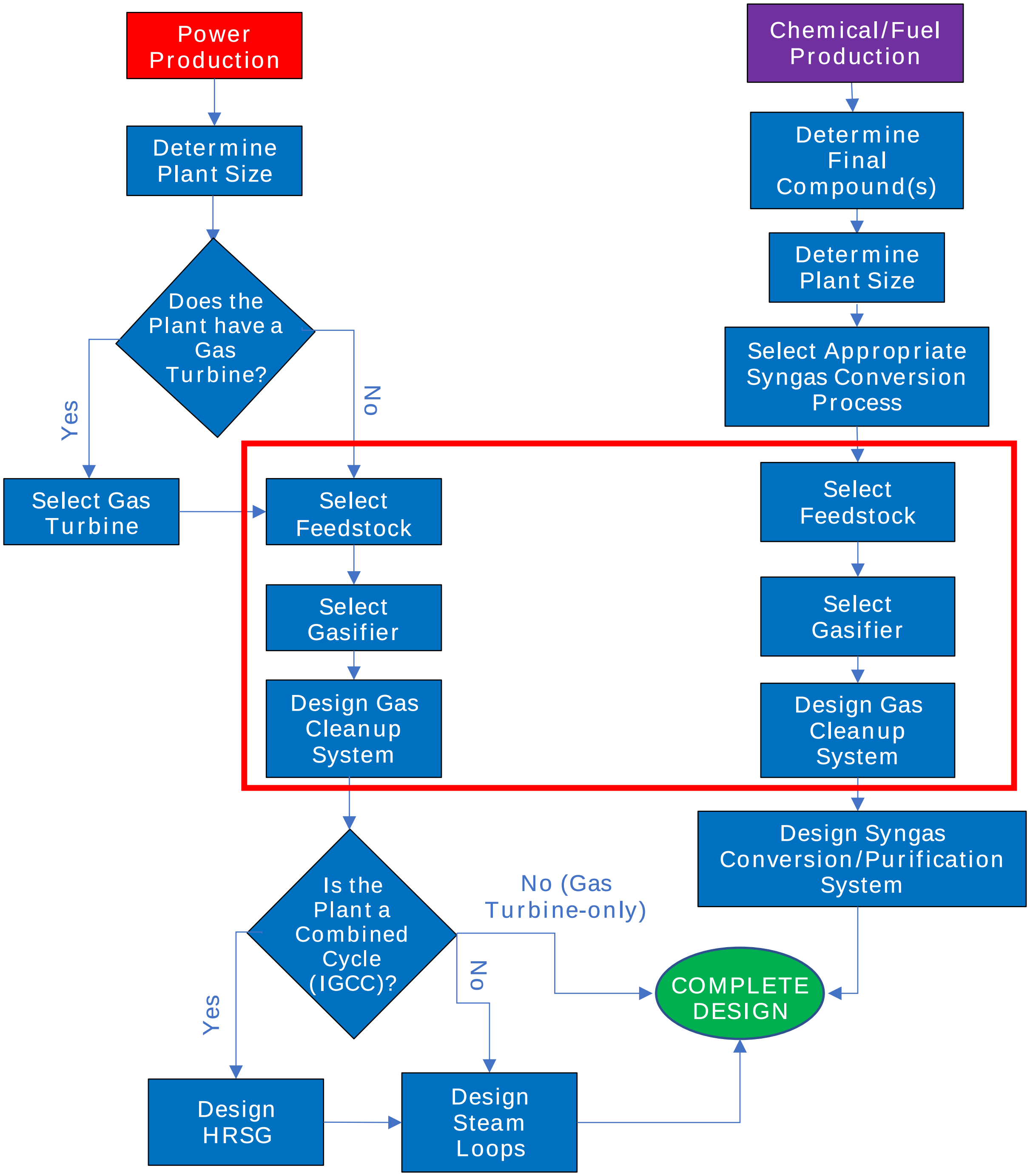


Figure 2-2. General design flowcharts for power and chemical production plants

If a plant is designed for both chemical and power production, the user will have to simultaneously follow both logic flowcharts. The two paths briefly overlap when the feedstock is to be determined and diverge again after designing an appropriate cleanup system. The remainder of this chapter and subsequent chapters assume that the user intends to choose only one of these two design paths for any given project and that the integration of the two plants can be achieved by splitting the final (clean) syngas into two streams: one to be converted into whatever chemical compounds the user has selected and one to be used in the power cycle of the user's choosing. A word of caution: a more efficient (or additional) syngas cleanup system will generally be required for the fuel/chemical path.

For a power plant, the designer should establish the plant power output then select the number of gas turbines required to meet that output. For an IGCC plant, the designer should understand that each selected gas turbine will likely end up providing an additional 50% of its rated power through downstream steam generation via a heat recovery steam generator (HRSG). All gas turbines selected should be of the same model. For Rankine cycle projects, a preliminary study should be performed based on an assumed syngas composition with roughly 15 to 20% of the higher heating value of natural gas (typically 22,453 to 23,726 British thermal units [Btu]/lb). Syngas power generation equipment may also have a substantially higher capital investment cost than comparable natural gas-fired power generation equipment due to their relatively low volume of production and specialty components required for syngas applications.

For a chemical production plant, the first (and most important) decision the user must make is what product(s) the plant is to produce. This is a key factor in choosing the correct syngas conversion process. For example, naphtha and other higher-order hydrocarbons are best created using a Fischer-Tropsch (F-T) reactor, while methanol and its derivates can be created via a methanol catalytic converter. The user should consider the answers to the following questions before proceeding with the design:

- What product(s) will the completed gasification plant produce?
- If the plant produces power, how large (in terms of megawatts [MW]) does the plant need to be? If the plant has a gas turbine, what and how many of each model would be best for meeting the power output requirements?
- If the plant produces chemicals or fuels, what conversion technology is most appropriate to reach the desired end result?

2.3 FEEDSTOCK SELECTION

The next major step is selecting an appropriate feedstock. Theoretically any hydrocarbon can be gasified, ranging from uncommon fuels like biomass and petcoke to niche compounds like plastic and leather.[1] In choosing a fuel, the designer should be aware of five key aspects: energy content (i.e., heating values), proximate analysis, ultimate analysis, sulfur analysis, and ash composition. In addition, the price of fuel and its transportation costs and logistics should be considered for the plant economic analysis. This section details each of these aspects.

2.3.1 FEEDSTOCK ENERGY CONTENT

The energy content for all potential fuels is measured in terms of a *Heating Value* or *Calorific Value*. This number is generally calculated based on stoichiometric complete combustion of the fuel, assuming all carbon (C) becomes carbon dioxide (CO_2), all hydrogen (H) becomes water (H_2O), and all sulfur (S) becomes sulfur dioxide (SO_2). The heating value will differ depending on whether water is assumed to be in the liquid or the vapor state when complete combustion occurs, which leads to two distinct values. The smaller number (assuming all water vapor) is called the *lower heating value (LHV)* or *lower calorific value (LCV)* and the larger (assuming all liquid) is called the *higher heating value (HHV)* or *higher calorific value (HCV)*. The actual heating/calorific value is dependent on the specifics of the application in question and falls somewhere between these two idealized extremes. As an example, the LHV of natural gas (assumed 100% methane [CH_4]) is around 21,433 Btu/lb and its HHV is around 23,811 Btu/lb. Larger heating values mean that the same amount of energy can be provided for the overall process using less fuel input; or, the same fuel input can be used to provide more energy to the process. This is an important consideration for power plant applications.

2.3.2 PROXIMATE AND ULTIMATE ANALYSES

All solid fuels can be broken down into four key components: moisture, ash, volatile matters, and fixed carbon. A listing of the fuel content in terms of these four categories is called the *fuel proximate analysis*. An example of this is shown in Table 2-1, which shows the proximate analysis of a typical willow wood species.

Table 2-1. Typical proximate analysis of willow wood[2]

COMPONENT	AMOUNT
Fixed Carbon	16.371%
Volatile Matters	75.330%
Moisture	7.106%
Ash	1.193%
TOTAL	**100%**

Related to the proximate analysis is the *fuel ultimate analysis*, which breaks the fuel down into *individual elemental species*, such as C, oxygen (O), H, S, and nitrogen (N). A typical ultimate analysis for willow wood is shown in Table 2-2.

Unlike proximate analysis, ultimate analysis can be specified on a "dry" or "wet" basis (Table 2-2 is on a dry basis), meaning that water content can be excluded from or included in the analysis. In either case, all H and O content shown will specifically be elemental hydrogen or oxygen contained in *non-water* compounds. "Wet" ultimate analysis will have an extra row or column specifically for moisture. It is important to consider whether the ultimate analysis is wet or dry before proceeding with design. Sulfur and nitrogen in most fuels exist as heteroatoms that lead to hydrogen sulfide (H_2S), carbonyl sulfide (COS), ammonia (NH_3), hydrogen cyanide (HCN), and other undesirable contaminants in syngas. In biomass, these compounds will exist as parts of complex molecules like proteins. Chlorine is also often considered, as it tends to become hydrogen chloride (HCl) after gasification, which will need to be removed to protect downstream equipment from acid attack.

Table 2-2. Typical dry, ash-free ultimate analysis for willow wood[2]

COMPONENT	AMOUNT
Carbon	45.492%
Hydrogen	6.358%
Oxygen	47.798%
Nitrogen	0.189%
Sulfur	0.163%
Chlorine	~0.0%
TOTAL	**100%**
Higher Heating Value (Btu/lb)	7,876
Higher Heating Value (MJ/kg)	18.32

2.3.3 SULFUR ANALYSIS

Sulfur exists in many forms. In solid fuels, it can be classified in one of three different ways. Most industries use the following naming convention for each of the three classifications of sulfur-containing compounds: *Pyritic Sulfur* (technically metal-sulfides), *Sulfates* (anything of the form XSO_4), and *Organic Sulfur* (sulfur bonded to carbon atoms, generally a part of the volatiles).

"Sulfur Analysis" refers to a general breakdown of the amount of the various forms of sulfur in a particular feedstock. In general, a sulfur analysis is obtained through various methods intended to measure the total sulfur content of a fuel (since each type of sulfur is responsive to different forms of measurement). However, knowing the sulfur analysis of a given fuel can be helpful, especially if some form of pre-processing sulfur removal is to be used, such as molten caustic leaching. For example, fossil fuels tend to be high in inorganic sulfur, while biomass tends to be high in organic sulfur.[3, 4, 5] Sulfur-containing compounds derived from gasification are generally undesirable and must be removed from the syngas via an acid gas removal (AGR) system (discussed in Section 2.8.2).

2.3.4 ASH ANALYSIS

Ash is a very diverse group of compounds, mostly consisting of metal oxides. It contains aluminum (Al), silicon (Si), titanium (Ti), manganese (Mn), and trace amounts of heavy metals (e.g., arsenic [As], selenium [Se], mercury [Hg], and rare earth elements [REEs]). Ash is highly relevant to gasification applications as it needs to be removed from the syngas stream before the syngas can be further processed. Gasification results in three forms of ash depending on whether the gasifier temperature is high enough to melt the ash. *Slag* is the term used to describe ash that has melted. Slag tends to be low in pores and high in density with a wide size range. *Fly ash* on the other hand is porous and much lower in density and is fragmented ash left from the gasification process. There may only be fly ash, only slag, or both depending on gasifier temperature. Molten slag will usually collect on the sides of the gasifier and exit from the bottom. However, smaller slag particles can entrain out of the gasifier with the fly ash and syngas. Larger ash particles will not entrain and will instead sink down to the bottom of the gasifier where they can be removed. This is called *bottom ash* or *clinker ash*, which is less common. Bottom ash, a congregate of fused ash, can reach sizes measured in inches (and perhaps feet) and forms either due to gasifier mal-operation or the fouling nature of ash. Bottom ash will also sink and is removed from the gasifier bottom. Due to its large size, clinker ash can block the gasifier's bottom outlet.

Ash entrained with syngas is removed either in dry form using cyclones and/or filters or in wet form using wet scrubbing. Depending on the pressure of the gasifier, the dry ash removal system may require extensive equipment to cool and inert the ash before discharge. The wet ash removal system results in the formation of black water, which may require extensive water treatment processing.

Slag and fly ash are useful byproducts of gasification, because they can be key components in concrete and road building materials.[6] Fly and clinker ash can also be useful in applications like roofing materials or soil enrichment (depending on composition). If the designer wishes to export the sequestered ash for profit, knowing its composition is a necessity. However, if ash is considered a waste product, it will have disposal costs that negatively impact the economics of the plant. The ash leaching properties must also be determined for environmental reasons. Table 2-3 shows a typical ash analysis for willow wood.

Table 2-3. Ash analysis of willow wood[7]

SPECIES	AMOUNT IN ASH
Phosphorus (P)	1.80%
Potassium (K)	20.12%
Calcium (Ca)	73.18%
Sulfur (S)	2.55%
Copper (Cu)	0.07%
Iron (Fe)	1.43%
Manganese (Mn)	0.49%
Zinc (Zn)	0.21%
Nickel (Ni)	0.02%
Chromium (Cr)	0.02%
Lead (Pb)	trace
Arsenic (As)	~0.00%
TOTAL	**100%**

Although most of the ash components in the fuel are collected as solid ash products (e.g., slag, fly ash, or bottom ash), some of the trace heavy metals (e.g., mercury, arsenic, and selenium) are converted to gaseous species during the gasification process. The mercury is typically reduced to predominantly elemental mercury, The arsenic and selenium are reduced in the reducing environment of the gasifier to arsine and hydrogen selenide. These trace element compounds must be removed by the syngas cleanup system, as these materials are strong poisons for many of the catalysts used for syngas conversion and are toxic compounds whose emissions are regulated.

To ensure that the syngas cleanup system is adequately designed to handle the removal of these trace elemental species to protect downstream catalyst systems and meet emission requirements, reasonable estimates for mercury, arsenic, and selenium should be obtained for the fuel source selected. This information about trace contaminants becomes even more important for other potential and blended fuel sources. The increase in these trace elements in the gasification fuel impacts the trace element composition of ash and syngas downstream of the gasifier.

2.3.5 COST

Cost is a major factor that should be considered in any choice of fuel. One cost factor is the fuel's *location* compared to where the finished plant is to be constructed, as this will affect transportation costs and the required storage capacity. Also, the fuel (especially fuels like biomass) may require pretreatment (e.g., drying, grinding, torrefaction, pelletization) before it can be used, which adds to the plant operating costs. Some fuels may have high *fouling factors*, a measure of how much residue the fuel will leave behind on the gasifier walls during operation over the course of one year,[8] which can affect maintenance costs. Finally, fuels high in ash, sulfur, or chlorine will add to downstream cleanup costs.

In choosing a feedstock, the previously mentioned six metrics (i.e., energy content, proximate analysis, ultimate analysis, sulfur analysis, ash analysis, and cost) need to be considered when selecting a fuel, depending on the application. Overall, the user should be prepared to answer the following questions at this stage when trying to design a new gasification plant:

- What available fuels (that are within the budget) have the highest heating values (for power plants)?
- What available fuels have proximate and ultimate analyses most conducive to what I want to produce (for chemical plants)?
- How much will it cost to clean the syngas once the fuel has been gasified? Is there a potential tradeoff to the cost of purchasing the fuel, such as the added efficiency from using a fuel with more energy content?
- Ultimately, what fuel that is available will result in the maximum possible throughput (i.e., profit) at the lowest possible TOTAL operating cost?

2.4 SITE SELECTION: GREENFIELD OR BROWNFIELD

A *Greenfield Project* is a project that is on essentially undeveloped land. That is, there is no existing infrastructure to retrofit or remodel: everything must be constructed from scratch. By contrast, a *Brownfield Project* is a project where the land has already been utilized for some purpose in the past, and there is some infrastructure remaining that the designer may wish to use. Greenfield projects generally have a higher risk factor, because the land is undeveloped, and costs tend to be high. However, since there is no existing infrastructure, the designer usually has considerable flexibility on how to design the plant to meet the owner's requirements. Brownfield projects will usually require the designer to work around the existing infrastructure and the associated constraints. For example, if an old steam power plant is to be repowered, the existing steam turbines will have design point mass flow rates that need to be adhered to (+/- 15%). On the plus side, this means that the project will usually be cheaper than an equivalent greenfield project, because some plant components will already be available onsite. This, however, may be offset by unforeseen added costs, such as the removal/demolition of unusable equipment or required upgrades for the plant to meet current codes. In some cases, this can result in higher costs than an equivalent greenfield project.

For this stage of the project, the following questions should be considered:

- Is this a greenfield or brownfield project?
- If there are both brownfield and greenfield sites available, which one will be the most economically viable for the project?
- If greenfield, what are the financial and environmental risks involved with developing the land?
- If brownfield, what infrastructure is available and how can it be incorporated into the design? What constraints does it impose on the design itself?

2.5 PERMITTING REQUIREMENTS AND LEGAL REGULATIONS

Any power or gasification plant constructed in the United States is subject to the Clean Air Act (CAA) (42 U.S.C. §7401, 1970) and all of its amendments, the Clean Water Act (CWA) (33 U.S.C. §1251, 1972) and related water regulatory laws, and the Solid Waste Disposal Act (42 U.S.C. §82, 1965), which was later amended by the Resource Conservation and Recovery Act (RCRA) (42 U.S.C. §6901, 1976). Under the CAA, the U.S. Environmental Protection Agency (EPA) publishes a list of national ambient air quality standards (NAAQS) periodically that covers emissions of six key pollutants: ozone (O_3), CO, SO_2, nitrogen dioxide (NO_2), particulate matter (covers PM-10 and PM-2.5), and lead (Pb). All new industrial plants, including gasification plants, must adhere to any state or local air pollution regulations to maintain an overall statewide air quality that meets the NAAQS standards.[9]

Table 2-4. Regulations affecting gasification-based power plants[10]

ENVIRONMENTAL MEDIA AND APPLICABLE REGULATIONS	POLLUTANTS REGULATED	REGULATORY BASIS
Air Pollution • National Ambient Air Quality Standards (NAAQS) • Federal New Source Performance Standards (NSPS) • Federal New Source Review (NSR) • Title IV, 1990 CAAA—Acid Deposition Control • Title III 1990 CAAA—Hazardous Air Pollutants • Title I, 1990 CAAA—Attainment Maintenance of NAAQS, Regional Programs—NO_X SIP Call • State Implementation Plans • Local Standards (air quality, emission limits, control methods)	SO_2, NO_X, PM10, Pb, O_2, CO, HAPs	Clean Air Act, Clean Air Act Amendments, State and local laws
Water Pollution • Federal Safe Drinking Water Standards (SDWS) • National Pollutant Discharge Elimination System Limits (NPDES) • State Pollutant Discharge Elimination System (SPDES) • Toxic and Hazardous Waste Regulations (Federal and State) • State and Local Standards (stream quality, effluent limits, treatment methods)	arsenic, benzene, cyanide, mercury, naphthalene, selenium, other organics, and trace metals	Clean Water Act, Safe Drinking Water Act, Resource Conservation and Recovery Act (RCRA) State and local laws
Solid Waste Discharge • RCRA Subtitle C Toxic and Hazardous Waste Regulations • RCRA Subtitle D Non-Hazardous Waste Regulations • State and Local Stands (Classification, Disposal Methods)	Fly Ash, Bottom Ash, Slag, Pollution Control Waste, Byproducts	Solid Waste Disposal Act as amended by the Resource Conservation and Recovery Act (RCRA)

The CWA establishes federal standards for wastewater quality. Under the CWA, any facility that has a wastewater stream must obtain a National Pollutant Discharge Elimination System (NPDES) permit to discharge the wastewater to any standing water source in the United States.[11] The RCRA provides a framework for the safe treatment and disposal of solid waste created by industrial processes.[12] In addition to these regulations, power-based applications (e.g., IGCC) are also subject to the standards of chemical plants, as the creation of syngas is considered a chemical production process all on its own. A list of regulatory standards and statutes for various pollutants affecting IGCC and other gasification-based systems (at the time of publication) is provided in Table 2-4. Note that state regulations may be more stringent than federal requirements.

Finally, in the case of power plants, the type of market needs to be addressed. There are two types of markets in the United States: regulated and deregulated. In a *regulated market*, a single vertically-integrated monopoly controls all of the electrical power output in the state, all end-users are required to purchase electricity from this single provider, and a state utility commission controls how much profit the provider can make, which essentially sets fixed electricity prices. In a *deregulated market*, the state has less control over electric utilities and customers can choose which company they wish to obtain their electricity from. In the continental United States, the only deregulated electricity market states are California, Connecticut, Delaware, Illinois, Maryland, Massachusetts, Michigan, New Hampshire, New Jersey, Oregon, Ohio, Pennsylvania, Rhode Island, Texas, and Virginia (at the time of publication).[13]

It is the responsibility of the project director and designer to comply with all these key pieces of federal legislation, as well as state and local laws, and apply for and obtain a permit for the plant to be constructed and operated. The designer and director should consider the following questions:

- What state and local laws apply to the type of plant that will be constructed? Can the appropriate state and local agencies be convinced to grant permission to begin construction?
- What emissions targets need to be achieved to satisfy existing federal laws (e.g., CAA, CWA, RCRA), and can the plant continue to meet more stringent targets if the EPA changes them?
- If the plant produces electricity, is the construction site located in a regulated or deregulated market? How does this affect economic decisions going forward?

2.6 SYNGAS CONVERSION TO END PRODUCTS (POWER/FUELS/CHEMICALS)

The first major decision of the design process is to determine the desired end products (i.e., power or chemicals/fuels). An introduction to several of the technologies follows, and additional discussion is available in Chapters 6 and 7.

2.6.1 POWER GENERATION

For the generation of electrical power, the primary technology for gasification systems is IGCC. It makes use of combined cycle technology, where clean syngas is burned in a gas turbine to provide power and then the exhaust from the gas turbine is sent through a large heat exchanger called a Heat Recovery Steam Generator (HRSG), which produces steam to drive a set of steam turbines for additional power. The HRSG fulfills the role of a boiler in a standard Rankine cycle that receives its heat energy from a standard gas turbine Brayton cycle. In the case of IGCC, the fuel for the gas turbine will be syngas created by a gasifier and purified via a *Gas Cleanup System* (see Chapter 6). A basic schematic of a typical IGCC system is shown in Figure 2-3.

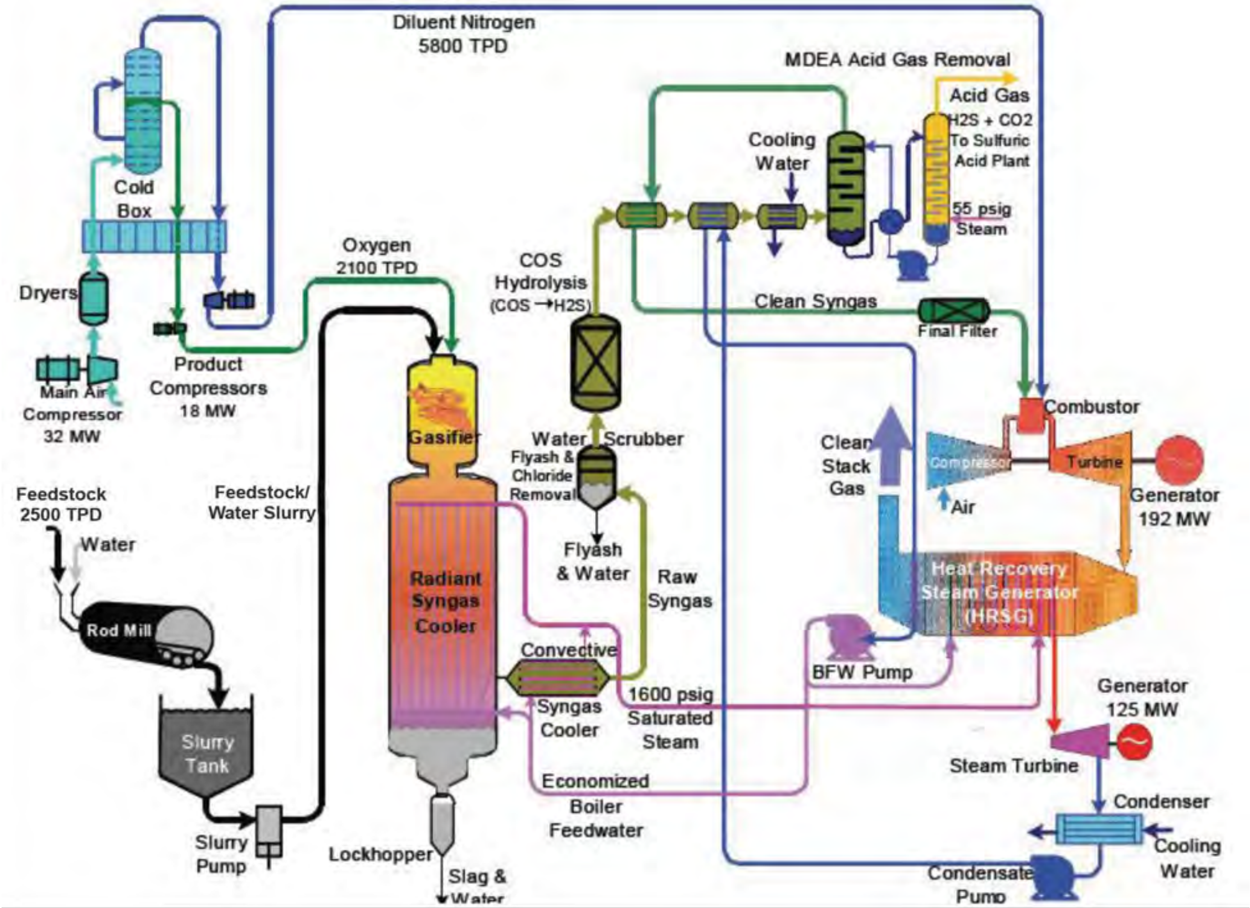


Figure 2-3. Typical IGCC power plant

An IGCC system is challenging to design and operate, as it requires the integration of a gas turbine with a gasifier. The gas turbines generally must be purchased “off-the-shelf” and generally cannot be custom designed; turbine manufacturers generally have pre-modified versions of their existing inventory designed specifically for operation with syngas as a fuel (e.g., the GE-7FB model). The gasifier output, meanwhile, is very sensitive to fuel input, geometry, gasifier temperature and pressure. Thus, for IGCC, it is best to first select an appropriately sized type and number of gas turbines and then design the rest of the plant to accommodate the gas turbine block’s fuel and air demands. Some additional considerations include the following:

1. Syngas tends to have high amounts of hydrogen, which has a very high flame speed and adiabatic flame temperature. Conventional gas turbines cannot handle fuels with high hydrogen content without modifications to safely handle it. The high flame temperature is generally handled by diluting the syngas with either nitrogen, CO_2, water/steam. It is important to choose a gas turbine designed with the use of hydrogen in mind (Some suppliers have modifications for syngas or other high-hydrogen fuels [e.g., most turbines with "FB" frames and above can handle this consideration (e.g., the GE7FB-series)]). In all cases, the designer should consult the manufacturer to ensure that the turbine they choose can work with syngas.
2. Syngas has very low density compared to other fuels, meaning the same mass flow rate takes up more volume than a standard gas turbine fuel system can typically accept. Some suppliers provide separate fuel nozzles for the same gas turbine to accommodate this issue. As always, the designer should consult the manufacturer about using this option (if it is available).
3. Syngas density and heating value may also force additional mass flow through the gas turbine itself. As with the other issues, many suppliers offer modifications that can accommodate this issue. The designer should also be aware of the increased potential for *compressor surge*, a condition in which the energy of the gas exceeds what the rotors in the compressor can sustain, resulting in rapid expansion of the gas during compression, which can lead to reverse flow and potential damage to the compressor. The designer should be aware of this potential issue and have a contingency in place that can allow the gas turbine to properly shut down in the event of a surge.

As an alternate to IGCC, a standard Rankine cycle with a boiler designed to operate with clean syngas from the gasifier may be used to generate power. Rankine cycle boilers are far less sensitive to changes to fuel composition and throughput than gas turbines, though the high flame temperature and flame speed of hydrogen is still of some concern around the fuel injectors. The component designers must pay special attention to the boiler fuel injection zone and ensure that the combustion flame will not extinguish and that measures are taken to protect the injectors from a high-temperature hydrogen flame. Steam turbines are much more customizable compared to gas turbines and can be scaled to fit the situation. Finally, for power plants designed for *load following* (changing power output over time to suit immediate demand needs), expect the gas and steam turbines to run at part-load for a significant portion of the plant's operating time. Modular or smaller-scale gasification power plants almost certainly will be required to load follow. For IGCC plants, the startup and shutdown times for each of the components tends to vary, so, in the case of load following, the designer should consider another power design structure or consider instead the purchase of a backup unit (such as a lone gas turbine simple cycle that burns natural gas or syngas from a smaller gasifier train) to boost power production during peak hours while the main plant is designed to operate at either full-load or 80 to 90% part load during off-peak hours.

2.6.2 CHEMICAL AND FUEL PRODUCTION

For chemical production facilities, there are four key technological pathways: methanol, substitute natural gas (SNG), Fischer-Tropsch (F-T) synthesis, and hydrogen. The details of these and other technologies are included in Chapter 9. Further conversion of these products into other compounds (like methyl esters) is not included in this chapter.

2.6.2.1 METHANOL

The production of methanol (CH_3OH) from syngas is a well-established commercial process. A highly selective Cu-Zn/Al_2O_3 catalyst is typically used and selectivities typically exceed 98% (based on syngas input). The process proceeds via the following sequence of reactions:

Methanol Production

$2H_2 + CO \rightarrow CH_3OH$ (R2.1 – Methanol Synthesis)

$CO + H_2O \leftrightarrow CO_2 + H_2$ (R2.2 – Water-gas shift reaction)

$CO_2 + 3H_2 \rightarrow CH_3OH + H_2O$ (R2.3 – Methanol Synthesis 2)

These reactions are highly exothermic and are generally performed in a high-pressure, fixed-bed reactor system.[14] Methanol is a versatile chemical that can be further upgraded through other processes, such as carbonylation to acetic acid, methanol esterification with acetic acid to methyl acetate, and methyl acetate carbonylation to acetic anhydride. Catalytic synthesis of ethanol from syngas produces a mixture of C_2+ alcohols and hydrocarbons,[15] but has poor selectivity/yield of any specific desired alcohol. Thus, this route has not been commercially applied to date.

2.6.2.2 SUBSTITUTE/SYNTHETIC NATURAL GAS

The primary component of natural gas is methane, which tends to appear in very low concentrations in most syngas mixtures that result from gasification. However, it is possible to design the gasification process that focuses on the production of methane for use as a substitute (or synthetic) natural gas (SNG). Selective catalytic methanation of syngas using a nickel-based catalyst is a well-established commercial process. The economic feasibility of SNG production is highly dependent on the prices of natural gas in the region where the SNG is to be exported.[16] In places like the United States, which has benefitted from hydraulic fracturing, natural gas tends to be very cheap, while in areas that are relatively poor in natural gas resources (e.g., Japan, China, India) prices will be higher and SNG production is much more economically viable. An example SNG plant is shown in Figure 2-4.

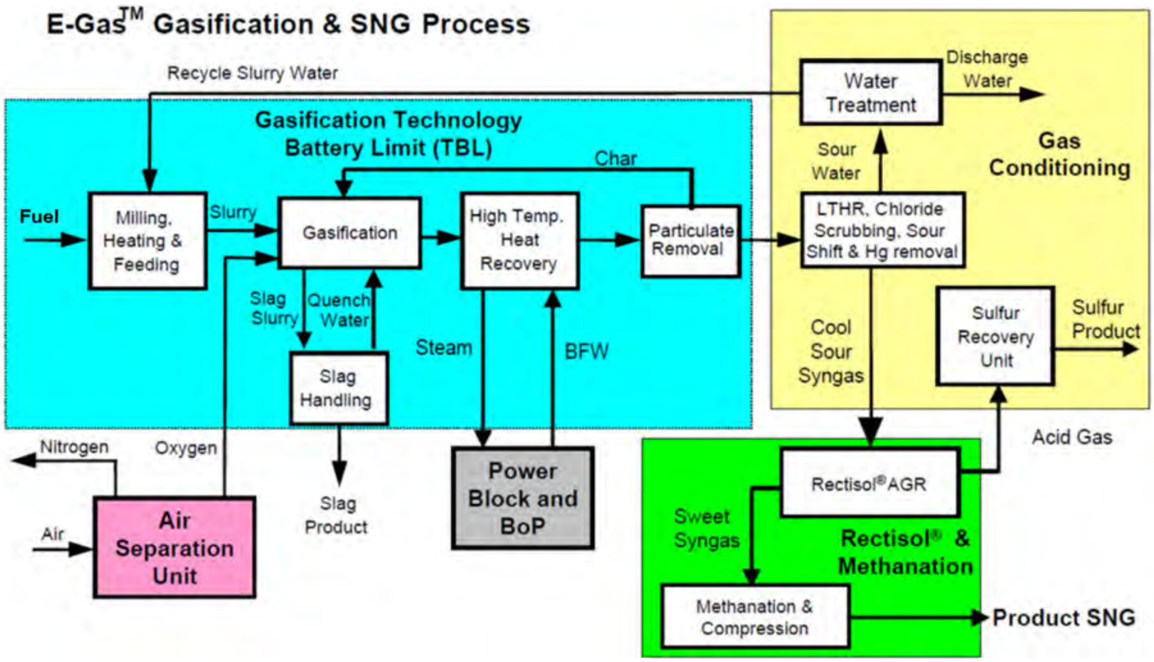


Figure 2-4. Typical SNG conversion process using E-Gas™ gasifier

The centerpiece of this plant is the methanation system, which consists of a sequence of reactors of varying temperature arranged in order from highest to lowest equilibrium temperature. Like methanol synthesis, SNG production is highly exothermic. SNG reactors typically require an H_2-to-CO ratio of roughly 3.0 and require the use of an air separation unit (ASU) during gasification to achieve the highest CH_4 purity possible. Finally, because of the highly exothermic nature of the SNG production process, it is possible to use the excess heat from the reactors to generate steam for export to other processes or to produce electrical power on the side.[17]

2.6.2.3 FISCHER-TROPSCH (F-T) SYNTHESIS

Liquid fuels have several advantages that make their use attractive in the power industry. These fuels are easier to store and transport than gaseous fuels due to their higher density and less stringent container requirements. Higher density also means that they take up less space for the same amount of mass. They have zero ash content and more uniform composition compared to solid fuels, which means the flame temperatures are easier to maintain during combustion and there is less wear-and-tear on reactor walls.[18] Liquid fuels can also be used in some gas turbines, unlike solids.[19]

The *Fischer-Tropsch (F-T) Process* transforms syngas, ideally with an H_2-to-CO ratio of 2.0, into a spectrum of liquid hydrocarbons, which can be used more efficiently and conveniently with downstream applications. Gasification is typically the first step of the F-T process, followed by a syngas cleaning system and a Fischer-Tropsch reactor, which can be fixed-bed, fluidized bed, or slurry bed type. In addition to higher-order alkane liquid fuels, the F-T process can also be used to produce waxes, gasoline, diesel, and low molecular weight olefins.[20] Wax byproducts can be used to produce lubricants or lightly hydrocracked to produce additional diesel and naphtha fuel.[21] A schematic of a typical F-T plant can be seen in Figure 2-5.

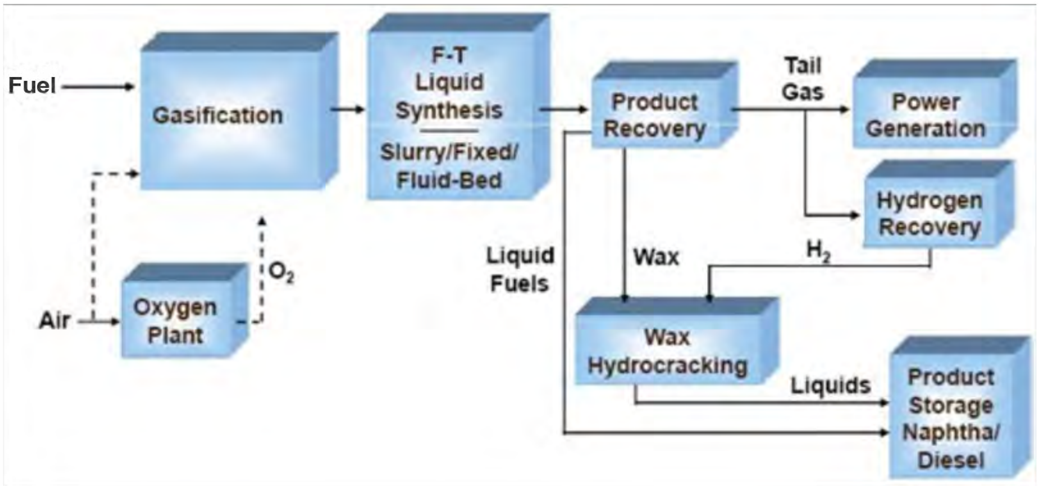


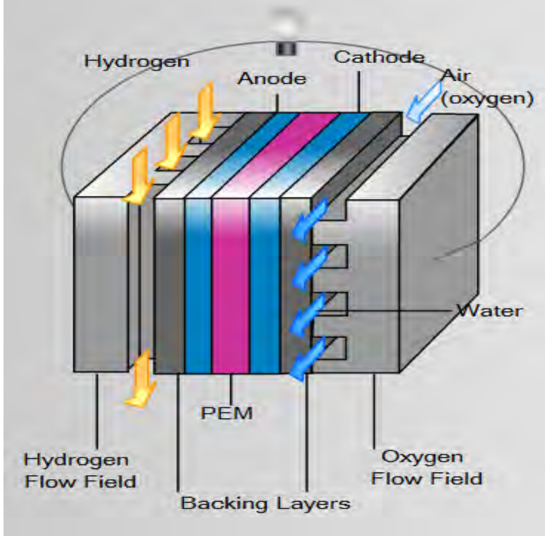
Figure 2-5. Example of a Fischer-Tropsch production scheme[20]

The shear versatility and flexibility of the F-T process make it ideal for fuel or chemical production plants that aim to produce a wide variety of compounds. The F-T process also leads to the creation of a very dense fuel gas as a byproduct. This "tail gas" consists of leftover carbon monoxide and hydrogen and smaller hydrocarbon byproducts (e.g., ethane, butane), which can be burned in either a gas turbine or a steam boiler to generate electricity.[21]

2.6.2.4 HYDROGEN

Hydrogen has several applications, including direct use as a fuel for hydrogen-powered vehicles or for power generation (either via combustion or via fuel cells), upgrading of refinery streams via hydrogenation, and production of ammonia (via chemical reaction with nitrogen). The required end-product purity of the hydrogen depends on the ultimate use of the hydrogen.

A *fuel cell* is an electrochemical energy storage device that uses a stored fuel (usually hydrogen) to produce electricity. Fuel cells typically have higher energy conversion efficiency than combustion engines,[22] but additional monetary and energy production costs. Hydrogen fuel cells work by converting hydrogen fuel into hydrogen cations (lone protons) by using an anode to strip the hydrogen atoms of their electrons. This process is augmented by a metal catalyst, like platinum. The electrons then exit into an external electric circuit to do useful work. The free hydrogens move across a *proton exchange membrane (PEM)*, see Figure 2-6, and contact the cathode, which accepts oxygen from the environment to react exothermically with the hydrogen anions and the electrons that were stripped from them earlier. The only byproducts from this process are heat and water.[22] Aside from PEMs, other types of fuel cells that are compatible with hydrogen include solid oxide fuel cells (SOFCs) and molten carbonate fuel cells (MCFCs). SOFCs and MCFCs have an advantage in that they are more suitable for more traditional syngas (with CO, H_2, and CH_4 as primary fuels) rather than pure hydrogen. Each type of fuel cell is named for the electrolyte used to transfer chemical ions from one side of the cell to the other (e.g., solid oxides for SOFCs).



Used with permission from Hydrogenics[22]

Figure 2-6. Hydrogen PEM fuel cell

In summary, the designer should be aware of the numerous technologies available that are compatible with gasification and consider costs and emissions when determining what technological pathway to choose to reach the desired end products. The designer and all project leaders should consider the following questions during this stage of development:

- What technology is most appropriate for the end products that will be produced?
- If a power plant, what type of power system can be built with the given budget, and is the potential return worth the investment? (IGCC is the most efficient, but also has the highest upfront costs.)
- If the plant has a gas turbine in it, is it compatible with current design decisions (e.g., grid frequency) and how does it affect other design decisions (e.g., fuel source, air separation/integration)?
- For a chemical/fuel production plant, what is the desired end-use of the product to be made, and what is the most optimal technological pathway to facilitate it?

2.7 GASIFIER SELECTION, AIR SUPPLY, AND FEEDSTOCK HANDLING

Once the desired end products are determined and a framework has been established, it is time to design the plant's gasification block. The following sections cover the design of the gasifier itself, the air delivery system, and any handling/pretreatment necessary for the desired feedstock.

2.7.1 GASIFIER SELECTION

Gasifiers come in many varieties, each with a unique operational design. The types of gasifier that have enjoyed the most widespread commercial success include *down-draft, up-draft, fluidized bed, entrained flow,* and *transport*. A summary of the typical specifications of each of these gasifier types is presented in Table 2-5. The choice of gasifier aligns with the type of fuel chosen.

Table 2-5. Various gasifier types[23, 24, 25, 26]

TYPE	DOWN-DRAFT	UP-DRAFT	FLUIDIZED BED	ENTRAINED FLOW	TRANSPORT
Average Operating Temperature (°F)	790 to 1,250	790 to 1,250	1,700 to 1,900	>2,000	1,750 to 1,825
Required Fuel Grain Size (mm)	<50	<50	<6	<0.10	<0.40
Fuel Preparation Generally Needed	Drying	None	Grinding	Pulverization	Grinding
General Feedstock Requirements	Low moisture content	Low tar content	High ash fusion temperature (avoid slag)	None	None
Throughput ("flow rate")	Low	Low	Medium	High	Moderate-High
Ash Condition	Slag/Fly Ash	Slag/Fly Ash	Clinker Agglomeration	Slag	Fly Ash
Feed Conditions	Dry	Dry	Dry	Dry or slurry	Dry
Other Requirements	None	None	Requires skilled operator	High oxygen	Requires transport agent

The *down-draft gasifier* is designed to accept fuel and oxidant from the top of the gasifier and both flow co-currently toward the bottom while chemical reactions occur. This has several implications. First, the highest gas temperatures in the entire reactor will occur in the pyrolysis zone.[24] Since the pyrolysis zone is above the combustion zone, all volatiles and tars pass through the high temperatures (more than 1,800°F) of the combustion zone, and result in their destruction via cracking and oxidation. The high temperature gas entering the reduction zone promote the endothermic gasification reactions in the reduction zone, cooling the gas to 1,100 to 1,200°F. There is a fair amount of waste heat associated with this type of gasifier and it generally has low carbon conversion rates (about 94%); however, because of this, leftover char can be sold as a byproduct. This attribute has led to small-scale gasifiers of this type being used to power vehicles since World War II. It is difficult to scale up, because of the heat flow in the opposite direction of the gas flow, which creates eddies. Commercial applications are limited to date to biomass gasifiers smaller than 1 megawatt-thermal (MWth). Due to the low maximum temperature (highest in pyrolysis zone), most of the ash will tend to be fly ash.[25] A basic design of a down-draft gasifier is shown in Figure 2-7.

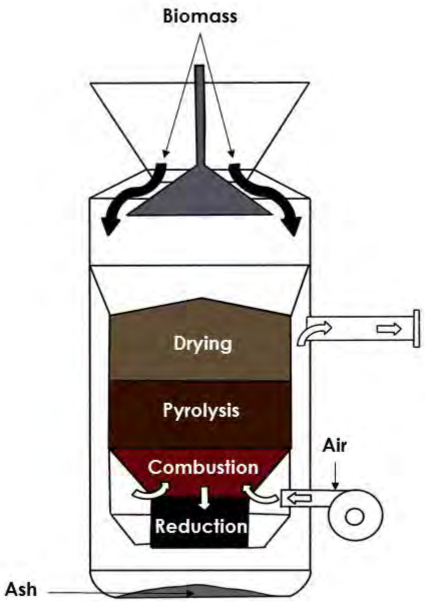


Figure 2-7. Typical down-draft gasifier[27]

The *up-draft gasifier*, conversely, has oxidant enter from the bottom of the gasifier and flow *counter* to the direction of the feedstock. This design creates a very intense combustion zone in the region that the two streams first meet, resulting in very high temperatures. These high temperatures can lead to damage to the gasifier components, such as the grate that prevents ungasified feedstock from falling to the bottom with the ash. Thus, it is often necessary to use steam or a gaseous coolant to protect sensitive areas of the gasifier.[24] However, this means that the high temperature gases act as a natural drying agent for the entering feedstock, which means that (1) fuels with higher moisture content can be used, including raw biomass and (2) up-draft gasifiers have higher carbon conversion rates than down-draft gasifiers. This allows the syngas to exit at a much lower temperature than in down-draft gasifiers, resulting in less waste heat. The devolatilization occurs at relatively low temperatures above the gasification zone, resulting in copious amounts of tar, phenols, and ammonia that must be separated from the syngas before downstream applications.[27] The Lurgi gasifier is an example of a commercial large-scale updraft gasifier. It has been successfully employed in several applications around the world, including Sasol's coal-to-liquids plant in South Africa since 1955 and the more recent Dakota Gasification Co. Great Plains Synfuels Plant (North Dakota, USA). A schematic of a typical updraft gasifier is shown in Figure 2-8.

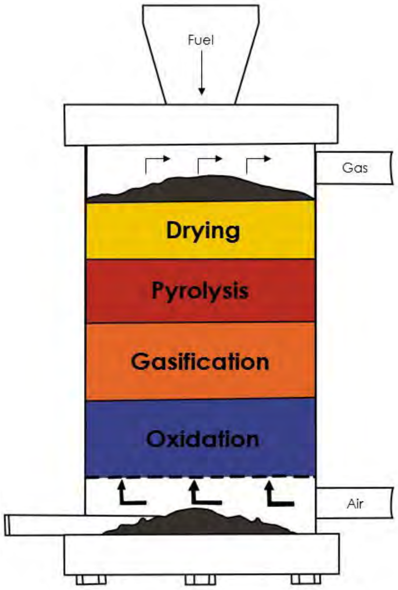
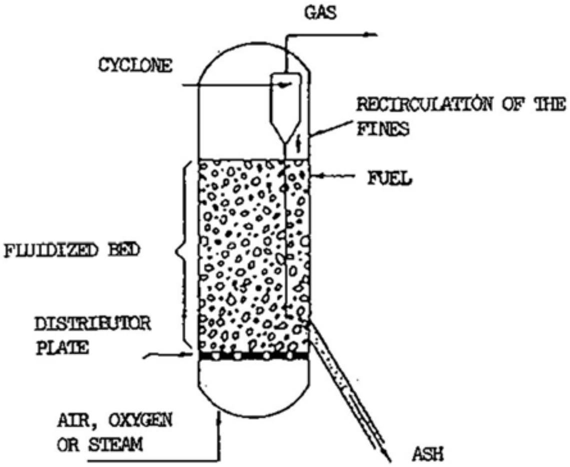


Figure 2-8. Typical up-draft gasifier[28]

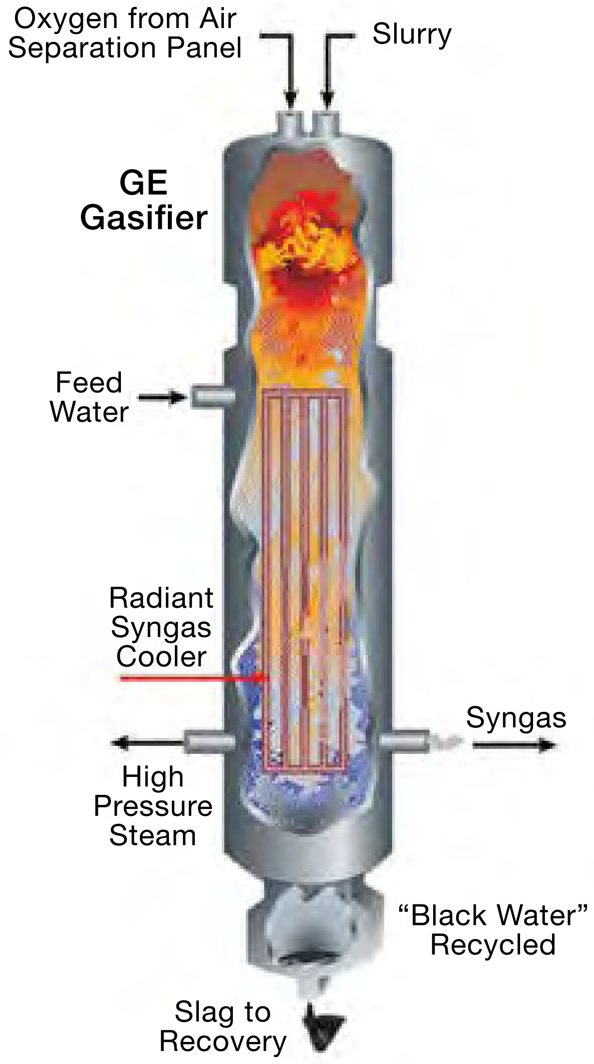
Fluidized bed gasifiers make use of fluidization to create a reactor bed. The reactor is initially filled with solid, dry feedstock particles and then air, oxygen, or steam is blown upwards through the bed at a steadily increasing rate. When the local velocity of the oxidizing agent reaches the average terminal settling velocity of the particles, they become fully suspended in the fluid stream, behaving very similarly to a liquid: hence, they have been "fluidized." They are more suitable for mid- to large-scale operations, and they are difficult to operate, because their use is dependent on sustaining a very delicate equilibrium state. They are also only suited for feedstocks that have very high ash-fusion temperatures (i.e., ash melting points), because slag production can cause the fluidized bed particles to stick together, which raises their terminal settling velocity and ruins the fluidization effect; and, they require dry feedstock that has been grinded down into smaller particles (less than 6 mm). They also operate at very high temperatures, making them suitable for the gasification of high rank coals, and have better throughputs than the fixed-bed types.[29] A schematic of a typical fluidized bed gasifier can be seen in Figure 2-9. Examples of this type of gasifier are the high-temperature Winkler (HTW) gasifier, the Kellogg-Rust Westinghouse (KRW, now defunct) gasifier, and the U-Gas® gasifier developed by the Gas Technology Institute (licensed to Synthesis Energy Systems [SES], but is in the process of being sold to Australia Future Energy [at the time of publication]).



Used with permission from FAO[29]

Figure 2-9. Typical (bubbling) fluidized bed gasifier

The fourth type of gasifier, the *entrained flow gasifier*, has been the most successful gasifier at large scales (greater than 250 MW). When a small particle is subjected to a fluid moving at a relatively high velocity, it will flow with the fluid, following along the streamlines of the fluid as it travels. At this point, the solid has become *entrained* or "trapped" by the moving fluid. The *entrainment* that results is a matrix of solid or liquid particles within a gaseous medium. This allows for a much more even temperature distribution and steady reaction rates. A schematic of a typical entrained flow gasifier is shown in Figure 2-10.



Source: U.S. Patent Office[30]

Figure 2-10. Typical entrained flow gasifier from GE

Entrained flow gasifiers are very common in large-scale power plants, because they can achieve very high syngas mass flow rates and high yields—higher than any other gasifier type. They operate at extremely high temperatures throughout the gasifier (with exit temperatures more than 2,000°F), resulting in large amount of waste heat, but with nearly zero tar in the syngas. They also have the distinction of being able to accept *virtually any kind of feedstock* and can be used in applications of nearly any scale, but they are particularly suited for very large (more than 250 MW) plants.[23] Finally, entrained flow gasifiers require the feedstock to be ground down into very small particles to create the entrainment (less than 100 microns) and are generally required to be oxygen-blown (details in Section 2.7.2). As the leading technology for large-scale plants, most models of entrained flow are manufactured by large power, oil, and gas companies, like General Electric (GE), Shell Corporation, and Mitsubishi Heavy Industries (MHI). Other models include the E-Gas™ gasifier (owned by McDermott, but is in the process of being sold to the Chatterjee Group in India [at the time of publication]) and the PRENFLO® gasifier (owned by Udhe after merging with Krupp Koppers in 1997 [at the time of publication]). Both the GE and the Shell gasifier technologies are now owned by Air Products Syngas Solutions™.

Finally, the transport gasifier was designed by KBR in 1996. It is very similar to the fluidized bed gasifier and can be used as both a gasifier and a combustion reactor. It is unique in that there is no true "bed" in the gasifier itself, as the feedstock, gasifying agent, and transport agent (sand) are constantly in motion throughout the system, much like an entrained flow gasifier, but with larger sand particles as the heat transfer agent. Ash and unconverted char particles are filtered out via a gravity-driven "disengager" (for larger particles) and a high-temperature cyclone filter (for smaller particles). Char particles separated in this fashion are sent back to the "mixing zone" (where the feedstock first enters the device) through a pipe called the J-leg, where they are re-gasified with a portion of the recycled syngas that came with them through the cyclone. A schematic of this type of gasifier can be seen in Figure 2-11.

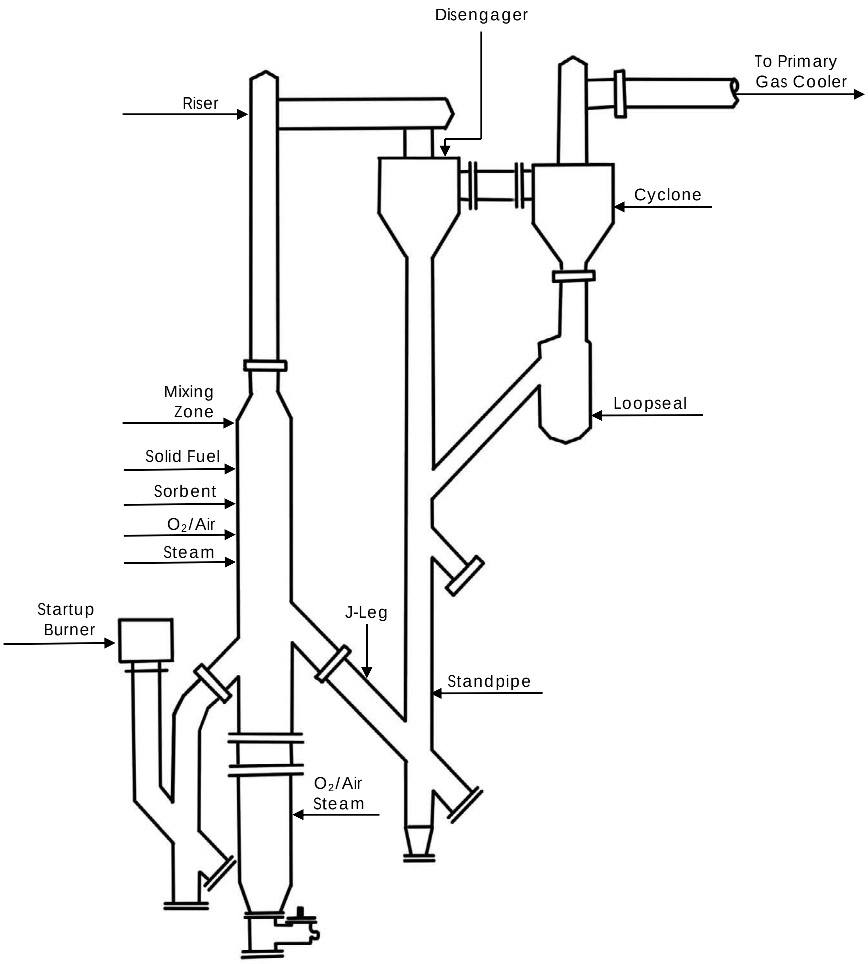


Figure 2-11. Transport gasifier from KBR[31]

The transport gasifier can operate at temperatures of up to 1,825°F and pressures up to 700 psig. Its operation is somewhere between a fluidized bed and an entrained flow gasifier, having high throughputs, but without needing high amounts of grinding to achieve such small particle sizes. For instance, the mixing zone area behaves similarly to a fluidized bed, but the riser and other such areas exhibit entrained flow behavior. While it has several advantages compared to entrained flow gasifiers, such as less waste heat and less-strict particle size requirements, there are no large-scale gasification plants using this gasifier (at the time of publication). A commercial plant constructed in Kemper County, Mississippi (USA) by Southern Company, but the plant only operated for a few months before being shut down. The plant is running using low-cost natural gas as a fuel (at the time of publication).

Another gasifier type that is worthy of mention is the *indirect gasifier*. Indirect gasifiers oxidize the fuel using only steam with no air or oxygen supply. This is possible because the combustion and gasification processes are carried out in different chambers, meaning that heat is transferred to the fuel indirectly through the walls. First developed by the Energy Research Centre of the Netherlands (ECN) and trademarked as “the MILENA process,” these gasifiers generally have low operating temperatures (typically approximately 800 to 900°C or approximately 1,470 to 1,652°F); as a result, long particle residence times are required to avoid high degrees of tar production. These gasifiers are well-suited to smaller, modular scales, because they can produce syngas with no added nitrogen without the need for an ASU (see Section 2.7.2). There are several examples of such gasifiers operating in Indonesia, Denmark, and the Netherlands.

2.7.2 AIR SUPPLY REQUIREMENTS

Gasification should ideally occur using pure oxygen (O_2) to support the reaction mechanism and avoid syngas dilution from inert nitrogen. The only current commercially viable way to produce high purity oxygen on the larger scale needed for most gasification processes is via a *cryogenic distillation process* using an ASU. Other, smaller-scale processes, such as vacuum swing adsorption (VSA), may be used for small oxygen sources, but the oxygen purity is limited to approximately 95% (balance nitrogen). While power plants can make do with an air-blown scheme (see Sections 3.2 and 3.3.3), chemical production plants require oxygen to maintain high syngas purity. A sample schematic of an ASU is shown in Figure 2-12. For smaller, modular gasification systems, as envisioned in this handbook, it is possible that alternatives to the cryogenic distillation process, such as membrane systems or adsorption-based processes, might be feasible (particularly for applications where enriched air or lower-quality oxygen supply would suffice); however, these technologies have not traditionally been applied at such scales. In addition, as discussed previously, indirect/steam-blown gasifiers would work well at modular scale since they can provide high-purity syngas without needing oxygen.

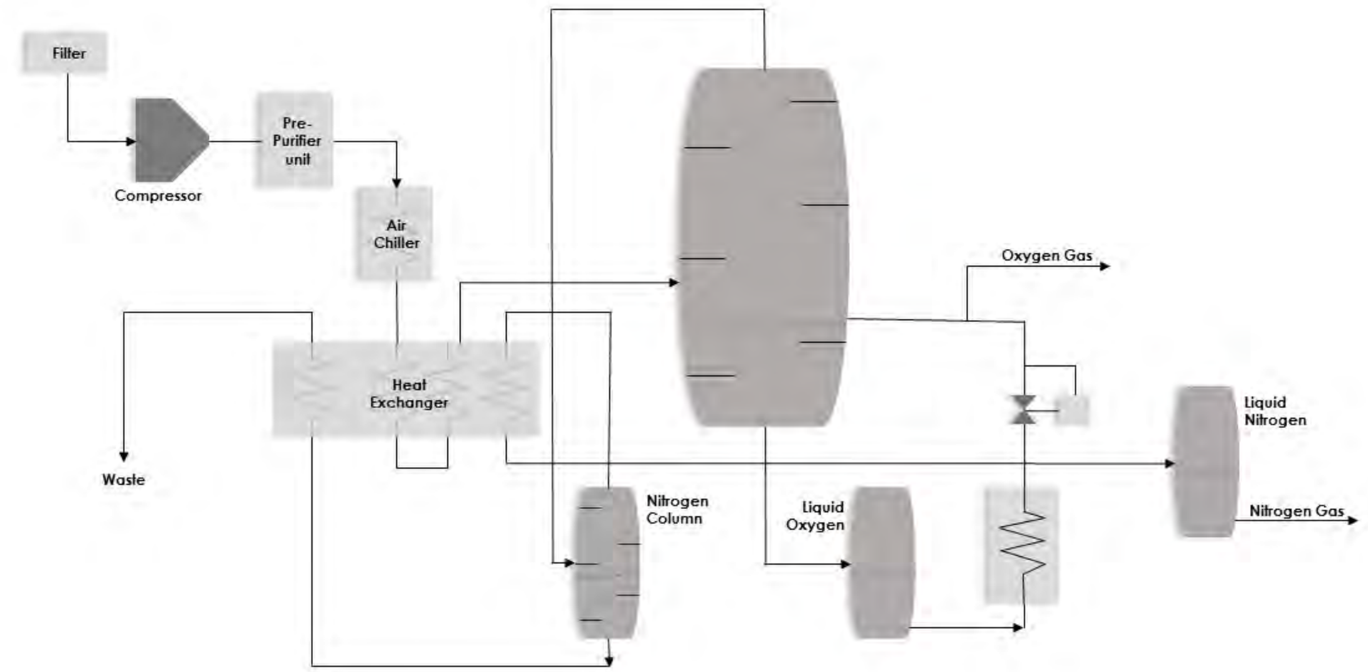


Figure 2-12. Typical ASU plant design[32]

ASUs work by repeatedly compressing and cooling air until the components turn into liquids. Nitrogen, oxygen, and argon have different condensation points, so they come out of solution at different points during the process, making it easy to separate them. In this manner, high-purity oxygen can be delivered to the gasifier, while nitrogen can be used in the gas turbine combustor (if one is present) to dilute the incoming syngas, reducing the flame temperature and, by extension, reducing thermal nitrogen oxide [NO_X] emissions).

The purity of oxygen needed depends on the gasifier itself and the application it is used in, but 98% purity is typically desired. Up to 99.5% oxygen purity is attainable, but with significantly increased costs. For example, most chemical production plants would prefer pure oxygen, as nitrogen is generally inert and would dilute potential product streams, which reduces throughputs and adds extra costs in removing the nitrogen from the product streams to increase the products' purity. Most entrained flow gasifiers require oxygen. At the time of publication, the MHI gasifier is the only large-scale commercial model that can accept air as an oxidant—but only for power applications (chemical production plants still require oxygen).[33] For power plants, an ASU is usually optional and comes with a tradeoff. "Oxygen-blown" gasification plants have higher gross power outputs due to better syngas quality and throughput (no nitrogen dilution). ASUs require very large amounts of electrical power and often make up more than 60 to 65% of the plant's total auxiliary power.[34] "Air-blown" systems have far fewer auxiliaries than oxygen-blown ones. "Steam-blown" gasifiers are similar to air-blown ones in this regard. The oxidant scheme that ends up being more efficient depends entirely on the specifics of the design. The designer should be aware of the specifications and demands of the gasifier and the application before recommending an ASU.

2.7.3 FEEDSTOCK PRETREATMENT REQUIREMENTS

As shown in Table 2-5, each gasifier type has certain requirements that the feedstock must have to be compatible. The most common requirement is *particle size*, which can be achieved by grinding. Entrained flow gasifiers have the strictest particle size requirements at 100 microns (0.1 mm), while fixed-/moving-bed types are the least strict at around 5 cm. Moisture and volatile content may be another requirement. Moisture content can be reduced through *drying*. Dryers come in a variety of types, including rotary dryers, fluid bed dryers, paddle dryers, and vibrating fluid dryers. The fuel can be dried using heat recovered from downstream operations (e.g., F-T synthesis), microwaves, hot gases generated by oil or gas combustion, steam, or electric heaters. A special mention goes to biomass, which, due to a number of feeding and availability issues, requires an extra layer of pretreatment before it can be used as a feedstock, such as torrefaction or flash pyrolysis.[34]

The *feeding system* of the gasifier also needs to be considered. Gasification requires special feeding techniques that is gasifier technology-specific; in general, it falls under two separate categories: slurry feeding or dry feeding.

- **Slurry Feeding**—The GE (formerly Texaco) and CB&I E-Gas™ gasifiers are examples of gasifiers that use a slurry feed system. The slurries are prepared by wet grinding the fuel in a rod mill. The feedstock is delivered by conveyor into the rod mill feed hopper. The fuel must be ground to the right particle size distribution to form a stable slurry with optimum solids concentration, which is typically 60 to 65% by weight. Recycled water is used when possible to minimize water usage. The prepared slurry is stored in an agitated tank and pumped into the gasifier at high pressure. Pumping the slurry is a proven process and can be designed proceed at relatively high pressure. The drawback is that preparing the slurry introduces a high amount of water to the gasifier. This water then absorbs a significant amount of the heat inside of the gasifier, lowering the gasifier's operating temperature. Higher oxygen consumption is required due to this lowered operating temperature, resulting in lower overall thermal efficiency.

- **Dry Feeding**—The Shell and Siemens gasifiers are examples of gasifiers that use a dry feed system. Dry feed systems use a lock hopper operating in batch mode, intermittently charging fines into the pressurized gasifier via staged opening and closing of valves on the top and bottom of the pressure vessel. During operation, the top valve is opened to receive fines into the lock hopper, while the bottom valve is maintained in a closed position. After the top valve is closed, the lock hopper is brought to or above the gasifier system pressure, typically with nitrogen from the ASU. Following pressurization, the bottom valve is opened, and the fines are discharged into the gasifier. After emptying the lock hopper, the bottom valve is closed, and the vessel is depressurized to allow another cycle. A parallel lock hopper system is normally employed to allow one unit to be online (i.e., discharging fuel at pressure to the gasifier), while the other is refilled and repressurized. Dry-fed gasifiers of both the entrained flow (Shell and Siemens) and fluidized-bed types (GTI U-Gas) usually require drying of the feedstock to reduce its surface moisture, which prevents small particles from sticking together during transport and feeding. The dry feed lock hopper system is reliable, but limited in achievable operating pressures (up to 600 or 700 psi). Therefore, it may be unsuitable for some higher-pressure operations. Also, unless the gasifier is oxygen-blown, and nitrogen (N_2) is readily available as a byproduct of the ASU, another source of nitrogen is required. If CO_2 is contemplated as carrier gas, additional design considerations are required to avoid the possibility of CO_2 condensing into a liquid or becoming supercritical before entering the gasifier.

Although the selection of dry feeding versus slurry feeding is mainly driven by the type of the gasifier (and as such may be inflexible), the nature of gasifier feeding (i.e., slurry or dry feed) will affect the ability to modularize the gasifier and gasification system. Each gasifier supplier may have different requirements or specifications to feed the feedstock, directly affecting efficiency and overall costs. For example, dry feed lock hoppers based on great vertical height configurations may not readily scale down. The different feeding options invoke different requirements for certain grinding and milling steps, likely making for different layouts.

For future advanced gasification systems, the U.S. Department of Energy (DOE) has engaged in collaborative work toward posimetric feeding, where solids are driven through a duct using positive displacement, usually via a rotating shaft element (e.g., Gas Technology Institute dry solids pump) to overcome the pressure limitations of lock hoppers. This type of feeding scheme may support modularization better than conventional lock hopper technology and serve as a future possibility for process intensification, because it can handle a variety of particle sizes and allow dry feeding at pressures more characteristic of slurry feeding. Also, the possibility of slurrying feedstock with liquid CO_2 has recently received ongoing attention in research and development (R&D), because it would address the excess water content of conventional slurry feeding schemes and the associated efficiency drawbacks.

Ultimately, feedstock pretreatment and feeding considerations warrant an iterative approach in determining the overall design layout of the plant. Different applications and different plant sizes will result in different requirements that affect what feedstock and gasifiers can be chosen. The designer should consider the right combination of feedstock, feeding scheme, gasifier, and oxidant supply scheme to create the most optimal syngas mixture that is best suited to the desired end-products. The following questions should be considered from a design point of view:

- How flexible is the plant idea in terms of overall size? How does this limit gasifier choices? Are multiple gasifier trains necessary to achieve the required throughput?
- For power plants, will air-blown or oxygen-blown gasification result in higher overall efficiency?
- How much grinding and/or drying is needed? Can this cost be reduced by going back and selecting a different fuel for the design? Could doing so also affect plant performance (efficiency/end-product throughput)?

2.8 SYNGAS CLEANUP TECHNOLOGIES

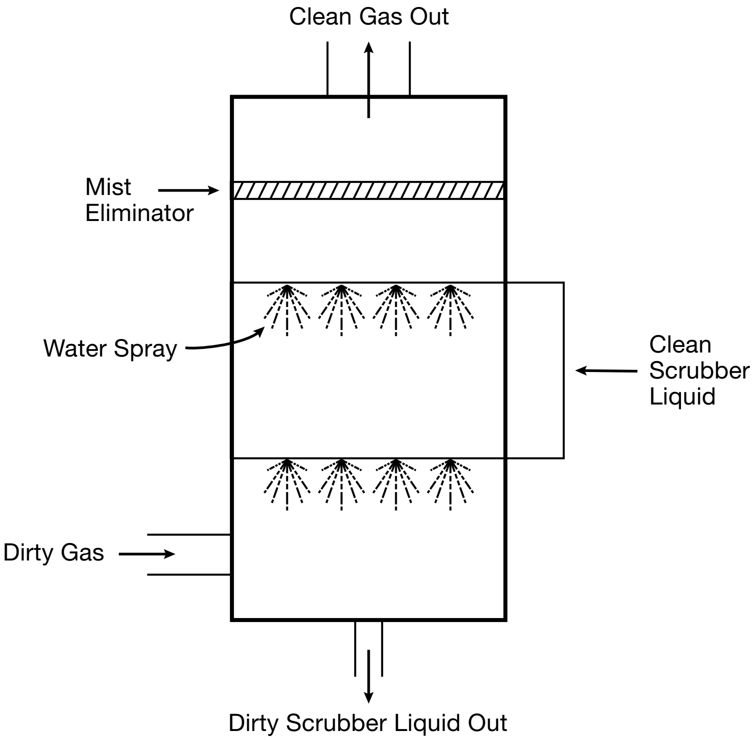
With the basic gasification block and known end-goals defined, the next step is to design the gas cleanup system that will purify the syngas before it can be utilized for the desired end product(s). Cleaning the syngas also helps curb emissions, as most of the impurities in the syngas are also common air pollutants (like particulate matter or carbon monoxide) or will become pollutants after processing (e.g., $H_2S \rightarrow SO_X$, $HCN \rightarrow NO_X$). This section will cover several technologies that the designer may consider when designing the gas cleanup system. The key contaminants that need to be controlled at this stage are particulates, sulfur compounds, and mercury (CO_2 is covered in its own section). Further discussion of these technologies is covered in Chapters 6 and 11. The designer should be aware that each of these technologies (depending on the type and model) requires a specific temperature range and the exit temperature of most gasifiers will be much higher than most of these technologies can handle. Thus, nearly every cleanup system will require some degree of cooling along the way. Gasifiers, like the GE Gasifier, often come with a radiant syngas cooler that can pre-cool the syngas, but it is up to the designer if further cooling during cleanup is needed. Cooling to at least 900°F is required to condense potentially damaging alkali metal compounds. A quench system can also be installed instead of or alongside the built-in gasifier cooling process, if needed.

2.8.1 PARTICULATE REMOVAL

Particulate matter, such as fly ash, fine slag, and stray carbon fines (soot), is important to remove from the syngas as early as possible, before any chemical or power conversion process has begun, as most of these processes were not designed for solids to be present. For example, gas turbines are very sensitive to particulates, which can damage the blades and reduce efficiency and raise maintenance costs. There are three primary pre-combustion technologies that can remove particulate matter: cyclones, wet scrubbers, and candle filters. Generally, a combination consisting of cyclones followed by either wet scrubber or candle filter can be employed to remove 99.9+% particulates.

Cyclones use the principle of centrifugal force to remove particulate matter from the surrounding gases. Since the solids are denser than the surrounding fluid, centrifugal force pushes them against the sides of the conical-shaped cyclone chamber where gravity pulls them down and out of the bottom. Meanwhile, the gases in the center of the tube spiral upwards and exit from the top.[35] A cyclone by itself often cannot meet the particulate removal targets needed to protect downstream equipment or meet environmental regulations. In this case, a filter, often a *candle filter* or *wet scrubber*, is used after the cyclone for further particulate removal.

Wet scrubbers use a liquid medium, usually water, to “scrub” the particulates from the gas stream. There are many wet scrubber designs. In the simplest design, water is sprayed into an enlarged chamber as the syngas flows through the scrubber from the sides near the bottom. The water droplets intercept the solid ash and soot particles, carrying them down to the bottom where they can be drained out. The cleaned gases can then freely exit through the top of the scrubber. An illustration of a wet scrubber is shown in Figure 2-13.



Source: EPA[36]

Figure 2-13. Spray tower wet scrubber

Wet scrubbers have an advantage over cyclones in that they can remove other compounds like ammonia and chloride compounds in addition to particulates. They are more expensive than cyclone filters, but they tend to be the preferred method of cleaning, because they can remove chlorides and ammonia.[36] The designer should consider this when deciding whether to use one of these or another technology to remove particulates. Note that this unit operation will generate a liquid waste stream that will require treatment or disposal.

Candle filters are the preferred technology if hot, particulate-free syngas is needed for downstream operations. Candle filters are composed of a pressure vessel, inside of which lie numerous small, perforated, ceramic tubes (the “candles”). These tubes are usually coated with a thin layer of woven or sintered material to increase filtration efficiency.[37] Fluid media passes over the surface of the “candles” and pass through the pores, while solids deposit on the surface of the candles, forming a “cake” layer. As the cake layer gets thicker, the pressure differential across the filter increases, and, once the differential reaches a predetermined value, the filter is back pulsed, causing the cake to break and the pieces to fall to the bottom of the filter.

Candle filters are most effective in the temperature range of 550 to 900°F. Above 900°F, alkali metal compounds tend to slip through and bypass the filter, as the vapor pressures of these compounds may be very high at that temperature range. Below 550°F, however, ammonium chloride compounds may “blind” the filter more quickly, reducing its overall lifespan.[38, 39]

2.8.2 SULFUR REMOVAL

Sulfur is often one of the limiting factors in terms of temperature when it comes to syngas cleaning. Current generation sulfur removal technologies that are used on syngas require temperatures below 120°F. For IGCC applications, the most common method of sulfur removal is by physical absorption using a regenerable commercial solvent, such as Selexol™, that operates at about 100°F using a dimethyl ether of polyethylene glycol, or Rectisol® that uses methanol at -40°F. Rectisol is commonly used when the syngas is used in a downstream cryogenic operation. For warm applications, Rectisol may not be competitive. Other sulfur removal processes that could be considered include alkanolamines and liquid redox systems. In many cases, the H_2S and other sulfur species cannot be vented as-is and must be converted to solid sulfur. The sulfur removal technology selected may depend upon the amount of sulfur removal required. The most prominent sulfur-containing compounds that result from gasification are hydrogen sulfide and COS. COS is very difficult to remove using these traditional AGR methods, so it is often shifted into hydrogen sulfide via a catalytic hydrolysis reactor by reacting it with steam/water (i.e., $COS + H_2O \rightarrow CO_2 + H_2S$). Carbonyl sulfide hydrolysis is most efficiently performed from 350 to 400°F and can achieve 99.5% conversion with proper catalysts.[38] Carbonyl sulfide hydrolysis, due to the similar reaction mechanism and required catalysts, can be performed simultaneously with water-gas shift if a sour-shift process is used. For chemical/fuel applications, this is an optimal way to control the hydrogen/carbon monoxide ratio using the water-gas shift reaction. The water-gas shift reaction also produces more CO_2 alongside hydrogen, making a potential CO_2 capture system easier to operate downstream.

A diagram of a typical absorption process is shown in Figure 2-14. The raw syngas enters the absorber column on the left, where it encounters a commercial solvent. The solvent absorbs the hydrogen sulfide (and some carbon dioxide) from the syngas and exits from the bottom. The cleaned syngas leaves the absorber through a vent at the top of the column. The rich solvent is then regenerated in a stripper column, which is heated by a steam reboiler (usually integrated with a steam loop to provide the heat). The increased temperature causes the captured gases to come out of solution. The gases then exit through the top of the stripper, and moisture is removed from the vented gases via a knockout drum. A sulfur-recovery step, such as the Claus process or liquid redox sulfur recovery (LRSR), may be required to convert the H_2S-laden gas to a material that may be safety disposed (see Section 2.10). Alternatively, H_2S can be separated out and sold as a useful byproduct. The regenerated "lean" solvent is then routed back to the absorber to remove more hydrogen sulfide from the syngas. Some acid gas removal processes may be able to selectively remove H_2S and CO_2 in the same unit. For example, the Coffeyville Resources ammonia plant (Kansas, USA) utilizes Selexol to remove H_2S and CO_2 from their syngas product selectively. In this instance, the H_2S flows to a sulfur recovery unit, some CO_2 is used internally to produce urea, and the remainder is either vented or sold to a third party for use in EOR operations.[40]

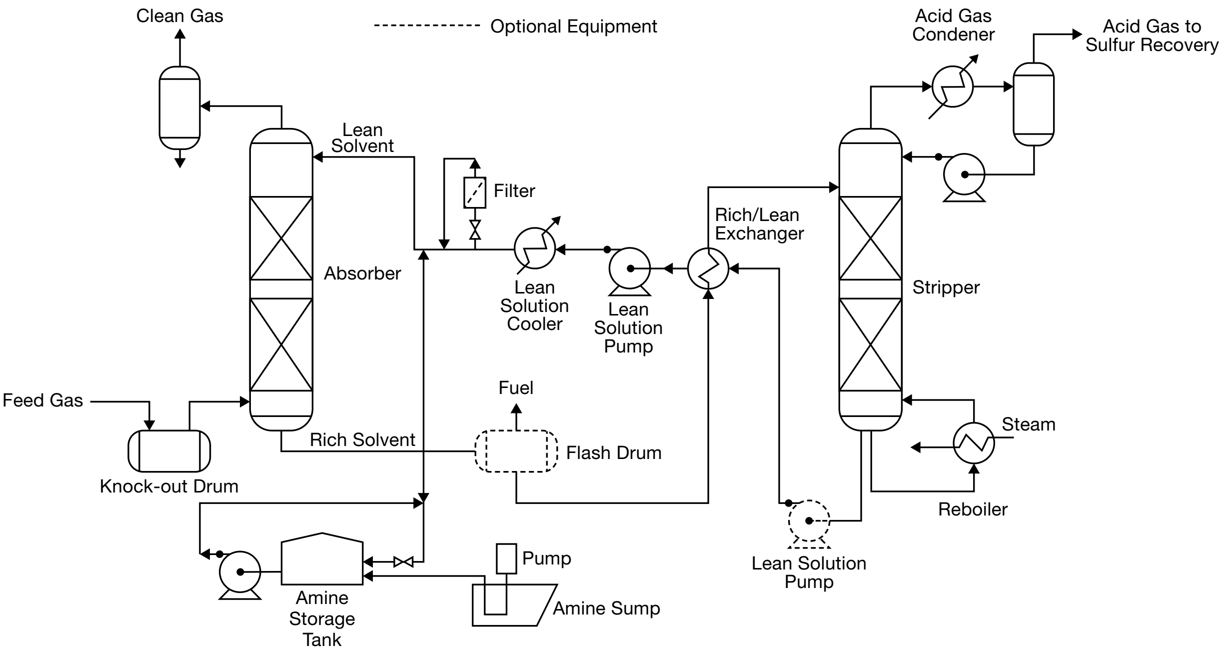


Figure 2-14. Typical absorption acid gas removal system based on methyl diethanolamine (MDEA)[41]

For higher temperature sulfur removal, there is a *Warm Gas Cleanup* option that is available from Research Triangle Institute (RTI) called the *Warm Gas Desulfurization Process (WDP)*, which uses a zinc-oxide reagent to capture H_2S and COS and convert them into a high-purity stream of SO_2. This process is more extensively detailed in Chapter 3 and Chapter 6. The designer should, in general, be aware of the allowable sulfur content for any downstream process for the syngas. For example, catalytic processes, such as methanol and F-T synthesis (FTS), require sulfur removal down to less than 100 parts per billion (ppb) or, more preferably, less than 30 ppb.

2.8.3 MERCURY REMOVAL

Unlike most ash components, elemental mercury is volatile and remains in the syngas after particulate/ash removal. As it is highly toxic to humans, mercury oftentimes must be removed through a separate system specifically designed to remove mercury. It is possible to perform this removal step downstream after the syngas utilization stage, but it is usually more effective and efficient to remove any mercury while the gas pressures are still high and gas volume flow is low. Mercury can be removed using a molecular sieve or an activated carbon filter via adsorption. The previously mentioned Rectisol® technology is also able to remove trace amounts of mercury alongside ammonia and hydrogen cyanide. Both technologies have extensive commercial success and can achieve high (greater than 90%) removal efficiencies. In SNG plants, extensive mercury removal is required to protect downstream components.[42]

A well-designed gas cleanup system is vital to the success of any gasification-based plant, regardless of application or syngas end-use. The designer should be able to answer the following questions before moving on to the next phase of the design:

- What is the expected range in volumetric composition of each component of the raw syngas (at gasifier exit), and how does this match up with the requirements of desired downstream devices/processes?
- Is any additional water-gas shift necessary to further augment the syngas H_2/CO ratio for, say, F-T or SNG synthesis?
- What are the purity requirements for the final syngas mixture, and what technologies work best to achieve these requirements for the current plant design and budget (e.g., lower capital and/or energy cost, high purity)?

2.9 CO_2 CAPTURE, UTILIZATION, AND STORAGE

With the increased demand for net-zero or net-negative carbon technologies, carbon capture and storage (CCS) has become necessary for gasification plants. For gasification-based plants, CO_2 capture is easier and cheaper to implement compared to other industries due to the high pressure of the syngas that is produced (resulting in greatly reduced gas volume and higher CO_2 concentration levels).[34] One important note for potential plant designers and owners: CO_2 capture modules (especially sorption processes) involve high levels of integration with other components in the plant.

To take advantage of the high pressure from the gasifier, the CO_2 capture system needs to be installed *before* any syngas conversion process (such as a gas turbine or F-T reactor) occurs. In other words, the plant must be designed with carbon capture, utilization, and storage (CCUS) already in mind. For applications other than power generation, most will require a CO_2 capture system for syngas cleanup, to increase product purity and improve reaction kinetics.

Pre-combustion CO_2 capture can be beneficial because of the ability to be performed simultaneously with AGR. In this case, the WGS reaction, as stated earlier, can be performed alongside carbonyl sulfide hydrolysis ("sour-shift" CCS) to increase the amount of CO_2 available for capture. One caveat to this approach is that the syngas will end up being very rich in hydrogen (see Section 2.6.1 for the impact on turbine design).

For Rankine cycle power plants, *oxy-fuel combustion* is an attractive CO_2 capture method. Since the fuel is syngas, there will not be any fuel-bound nitrogen and the sulfur and soot will have already been cleaned up by pre-combustion syngas cleaning (see Section 2.8). If the syngas was created using oxygen from an ASU, further oxygen from the ASU can be used to combust the syngas, resulting in a stream of very high purity CO_2 that (upon having the moisture removed) can be utilized or stored as-is with little or no further treatment.[43]

For gasification plants, almost all the technologies available for AGR are also useful for CO_2 capture. In general, they can be divided into three major categories: sorption processes, membrane processes, and cryogenic processes. Sorption processes include absorption and adsorption, membrane processes include organic and inorganic types, and cryogenic processes involve the sequential condensation of individual gas components at specific temperatures. Though it is the commonly accepted industrial term, the term "cryogenic" may be misleading if the CO_2 is simply being liquefied, because the temperature is only slightly below 0°F in most cases. Cryogenic processes are utilized in some natural gas treatment facilities. More information is provided in Chapters 11 and 12.

On the utilization and storage side of CCUS, enhanced oil recovery (EOR), where CO_2 is pumped into an oil well to raise the well pressure and increase the amount of oil that can be harvested, is the only current large-scale user of carbon dioxide. Other potential small-scale end uses of CO_2 include carbonated beverages, as a refrigerant, as a flame retardant (liquid CO_2 is commonly found in most fire extinguishers), as a pneumatic gas in pressure-based tools and mechanisms, and many other niche applications. If the captured CO_2 cannot be used, the plant owner can opt to store the CO_2 underground inside a compatible sedimentary rock (like sandstone and limestone)[44] or volcanic basalt[45] formation. The decision to store CO_2 underground must be made early in the project design to allow for the lengthy permitting process associated with CO_2 injection wells (e.g., in the United States). Carbon dioxide injection wells are classified as Class VI injection wells and few have been successfully permitted (at the time of publication). Some U.S. states, in particular North Dakota, have been or are pursuing primacy from the federal government, which may reduce the overall time for permit granting. In general, the plant owner and the designer should consider the following questions when designing a plant with CCUS:

- What percentage of the CO_2 needs to be captured when combined with the use of carbon-neutral fuels to achieve a maximum of zero carbon emissions?
- What available capture technology is most suited to the type of plant being designed (see Chapter 12)?
- Are there any necessary steps that need to be taken to enhance the CO_2 purity?
- Is the local geology suitable for CO_2 injection?
- What utilization options are available for the captured carbon dioxide? Can additional profits be secured from the sale of the captured CO_2?
- What federal (and/or state) mandates exist to take advantage of CO_2 storage? U.S. federal tax credits may make the capture and storage of the CO_2 attractive. If the gasification facility is developing fuel that can be sold into this market and the CO_2 is stored, the resulting tax credits will help justify the capital investment of the CCS system.

2.10 END-PRODUCT RECOVERY, CLEANUP, AND PURIFICATION

The final step of the preliminary design process is to finalize the deliverable state of the end products. For power plants, this step is not needed unless the power plant also produces some byproduct, such as char, or exhaust stream that requires treatment either for selling on the market or as part of meeting federal emissions regulations. One example of a technology that may be needed for this purpose is a *Claus Unit*: a well-known and highly utilized commercial system[46] that is designed to convert hydrogen sulfide from the AGR process into salable elemental sulfur via a series of chemical reactors and condensers. A basic overview of a Claus plant is shown in Figure 2-15.

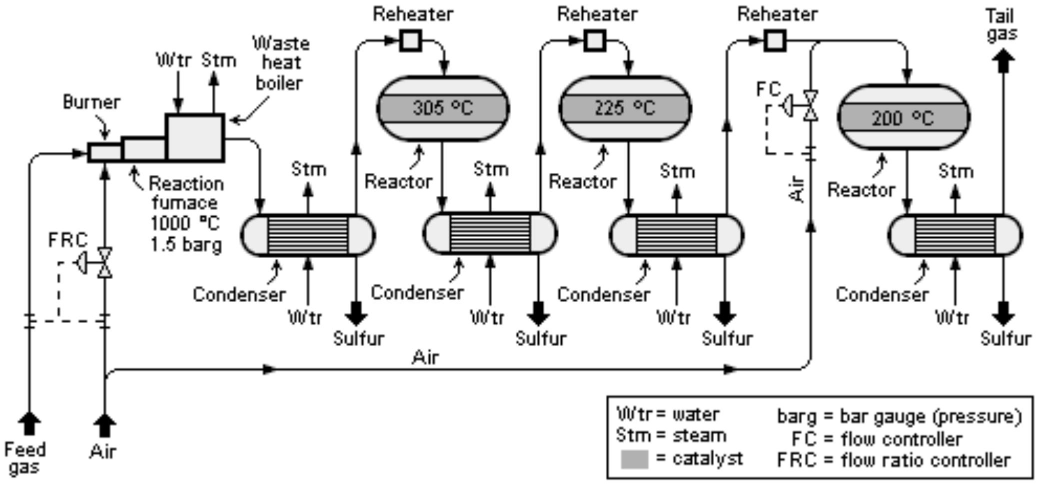


Figure 2-15. Claus process overview

The Claus unit operates by first burning the hydrogen sulfide to produce sulfur dioxide and water. Following this, the sulfur dioxide recombines with hydrogen sulfide to produce elemental sulfur and water. The first stages of the process perform these reactions thermally, while later stages augment the process with catalysts. Condenser sections use cooling water to remove any elemental sulfur produced as a liquid. Up to 97% sulfur recovery can be attained with only three catalytic stages.[46] Any leftover gases are burned, creating sulfur dioxide, and then either vented or sent to a tail-gas treatment unit. In IGCC plants, the leftover gases can also be recycled back to the AGR unit to further enhance sulfur recovery efficiency.[37, 47] The H_2S can also be converted to sulfuric acid as a saleable byproduct.

For CO_2 utilization designs, the captured CO_2 will likely need to have some of its impurities removed before utilization. These impurities include water, sulfur compounds (especially hydrogen sulfide for pre-combustion capture), and nitrogen, among others. These impurities can be removed via a cryogenic cooling/compression system not unlike that of an ASU, catalytic oxidation reactions (oxygen-only), a sorption process (like activated carbon or alumina), or some combination thereof.[48, 49]

For SNG, alcohol synthesis, and F-T plants, the purification is usually handled by the pre-process syngas cleanup system. However, the F-T process deserves special mention, as the post-processing/cleanup occurs at the same time that the end products of FTS are finalized (e.g., the generation of hydrogen to lightly hydrocrack the heavy waxes into diesel fuels and/or naphtha or the separation of the solid naphtha fuels from liquid diesel and gasoline). More detailed discussion of end-product purification is included in Chapter 10.

2.11 INTEGRATION WITH EXISTING INFRASTRUCTURE AND CONNECTING TO THE GRID

In the case of a brownfield project, some infrastructure may be incorporated into the overall plant design. This can help reduce the cost of new plant construction, but it comes with additional design constraints that need to be considered, as described in Section 2.4. In general, when adapting older technologies for a retrofit operation, some form of experimentation should be performed to determine the current operational performance of each available unit for the purpose of creating an informed plant computerized model before the construction phase.

In addition, power plants will generally have to be adapted to an *electrical grid*: either a national grid or local grid to supply power to consumers. All pieces of turbomachinery will be outfitted with a generator that will create the electricity, which, in most cases, will need to be converted to a three-phase alternating current (in the United States, the generator will almost always have this feature built-in). From there, the electric current will need to be sent to a step-up transformer to deliver the generated electricity to a substation that will, in turn, transmit the power to the distribution grid. Power plant owners should ensure that a power substation is within a reasonable distance of the plant to mitigate transmission losses and that the generated electricity is quickly and efficiently delivered to the power grid. The way that the plant will interact with the grid will be determined by whether the plant is a *baseload plant*, a *load* *following* plant, or a *backup* plant. A baseload plant will generally operate at full load with as little variability as possible. A backup plant will only operate for brief periods of time to meet periods of increased demand. As such, it will require very fast startup and shutdown times (not feasible for IGCC). Finally, a load following plant is somewhere between the two, operating at part-load (or full load with a backup power supply of some kind) with the ability to "ramp" up or down to keep up with periods of high demand variability.

At this stage, the designer(s) and operator(s) should work to integrate any new components of the proposed design with existing infrastructure and the grid. Owners and contractors should consider the following questions before proceeding with construction:

- What and how many of each power-producing device are used in the existing plant design? How many main generators are there, and how much power are they projected to produce under normal, steady operation conditions?
- If the plant produces power, is it a base load, load following, or backup plant? How does this affect the way the plant is to be integrated with the grid?
- What existing infrastructure is present onsite, and is it still usable with the current version of the design?

2.12 RELIABILITY, AVAILABILITY, AND MAINTAINABILITY REQUIREMENTS

A few of the benefits of a modular design are the effects on reliability, availability, and plant maintenance. With a standardized reference plant coupled with a high degree of modularization, improvements can be made in each of the aforementioned areas.[50] Modular designs are more reliable than non-modular designs due to the concept of *fault tolerance*, in which a broken or damaged module can be quickly swapped out with a replacement one while repairs are made on the former.[51] This reduces unnecessary income loss from added plant down time and increases availability as well. Modular designs are easier to retrofit due to this simple swap-in-swap-out capability, in addition to the fact that the ability to add, subtract, or modify modules makes it easier and faster to make incremental design changes to the system as a whole.[51]

No industrial plant has perfect availability. Most power plants are rated according to a *capacity factor*, which is the number of hours each year that the plant operates at full load divided by the number of actual hours in a year (8,760). Alternatively, it can refer to the amount of energy delivered to the grid during a given year divided by the total energy it would deliver if run at full load all year round (i.e., plant power rating x 8,760 hours). For example, normal IGCC plants have an average capacity factor of 85%.[37, 47] Modular systems will likely be able to achieve higher. If reliability, availability, and maintainability are a concern then the modular approach will greatly benefit plant owners and investors. Plant owners, managers, and designers should consider each of these three aspects for each plant module selected for the overall design. Some questions to consider related to reliability, availability, and maintainability requirements include the following:

- How many different design modules will the plant have and how are they to be connected to one another? Will each module be able to interact with the other modules in the way the design diagrams predict?
- Is there room in the budget going forward for enough backup modules in the event of emergency maintenance of one or more of the current ones?
- Is there enough onsite labor to assemble all the modules when they arrive from the manufacturer(s)? How much work is required onsite for complete assembly?
- What is the overall reliability of each module, and how will this affect the plant's throughput (chemical/fuel plants) or capacity factor (power plants)?

2.13 PROCESS SENSORS, SYSTEM ANALYTICS, AND CONTROL STRATEGIES

Once the base design of the plant is finalized, the next step is the ability to control and manage the operation of each section and their integration. Having a proper *control system* in place helps to maximize plant availability and reliability. Control systems also play an important role in plant safety (see Section 2.15). To implement a proper *control strategy*, however, an *analytical system* that measures the various parameters that need to be controlled is also necessary. For example, a load following plant will need some combination of devices (e.g., voltmeters, ammeters) to keep an accurate measurement of its current power output. These two digital systems should work in tandem as a sort of digital ecosystem that will help operators maintain steady plant operation. Such a system should be able to manage large amounts of data from each component and provide quick access to said data to all required personnel, utilize an asset model that can integrate all available data into easily understandable graphical data, and allow users to use advanced analytic services to rapidly respond to changes in plant operation by monitoring, improving, and updating analytics as needed.[52]

For gasification plants, some (non-exhaustive) examples of some key analytical parameters that may need to be monitored and controlled include the following:

- Gasifier operating conditions (e.g., pressure, temperature).
- Raw syngas composition (at gasifier exit).
- Syngas composition after cleaning (special attention to hydrogen/carbon monoxide ratio).
- Power output (if applicable).
- Total product(s) throughput (if applicable).

- Emissions analytics (e.g., sulfur oxide [SO_X], NO_X).
- Vibrational analysis of mechanical systems and turbomachinery (where applicable).
- Mass flows and temperature profiles in heat exchangers.
- Cooling tower water consumption (if applicable).
- Concentrations of contaminants in wastewater (before and after treatment).
- Fuel feed rate and particle size distribution (PSD).

There are many analytical devices that can be used to measure each of these parameters. An exhaustive list is not included in this handbook. The remainder of this section will focus on the implementation and integration of the analytical system with a reliable set of controls.

In general, there are three established control system types: a programmable logic controller (PLC), a distributed control system (DCS), and a supervisory control and data acquisition (SCADA) system. A PLC is a central computerized system that has been adapted for use in harsh industrial environments for the purposes of monitoring and controlling an industrial activity, such as an assembly line, ore processing facility, or wastewater treatment plant. PLCs have the advantage of being able to be reprogrammed at will, having similar architecture to most home computers (e.g., CPU, I/O modules, memory), and being able to automatically respond in real-time to analytics changes.[53]

A DCS, by contrast is a system in which multiple controllers are distributed throughout the plant linked to a central, integrated, supervisory control center. In this way, each individual controller monitors the performance parameters of an individual section of the plant and report its findings to a remote, centralized computer controller. Each module independently operates, which makes for an easier, less-computationally-intensive decision-making process. A DCS can be used to enhance production efficiency by improving response times. The decentralized nature of the DCS means that plant reliability is also improved: whereas a failure in a single central digital controller would result in a complete shutdown of all operations, an individual controller in a DCS only has dominion over a single component, so the rest of the plant may still be able to function while the one control unit is repaired. Due to the distributed nature of the system, they are most suited to larger, more complex tasks that require many multiple avenues of control.[54]

In a gasification facility, the gasifier is, in a lot of ways, a fired heater and must be controlled in a similar manner. This is most critical in oxygen-blown plants, where the nearly pure oxygen stream supports the combustion of almost any material. As a result, the control system, sensors, and output devices associated with the gasifier itself will likely need to be part of a Safety Instrumented System (SIS), which is somewhat similar to a burner management system on a regular fired heater. An SIS is an independent control system that has increased reliability associated with the main processor. Multiple input sensors (e.g., temperature, pressure, and flow transmitters) are usually required for an SIS, in addition to multiple output devices (e.g., valves). Output devices are regularly tested to ensure that they will perform their safety function on demand during an actual emergency. There are many brands of SIS available for the plant designer and they can usually be integrated into the facility's regular PLC or DCS network.

Finally, a SCADA system is very similar to a DCS in that it uses discretized control units to monitor individual plant components. Unlike DCSs, SCADA systems (which can also include PLCs) are mainly used for data gathering and analysis, whereas DCSs are more geared toward real-time control over the whole plant at once.[55] Many industrial plants utilize some combination of these systems as part of an overall *control strategy*. It is up to the plant managers to choose and implement an overall monitoring and control structure that best suits the plants' needs. The staff should consider the following questions in designing such a system:

- What and how many sections of the plant require computerized analytics monitoring? What data is needed to properly inform plant operators (and possibly the general public) about the plant's functions?
- What data measurement systems are best suited for measuring each parameter?
- How can the data best be digitized so that it can be communicated to the plant's control system?
- What overall control system structure is best for processing all the plant data and adjusting when necessary?
- What is the protocol for adjusting individual unit performance? Should the control system automatically adjust each parameter? Should operators have manual control? How much?
- What is the overall *control strategy* that can most effectively achieve the desired result of maximizing the value of the plant's output in the quickest, most reliable manner at the lowest costs?
- Can the chosen control system be used to improve overall plant safety (Section 2.15)?

2.14 ECONOMIC ANALYSIS

A complete economic analysis should be performed on the final design to confirm the economic viability of the project. Economic analysis at this stage always carries some degree of uncertainty, so potential investors should beware and assume some degree of variability in any cost estimates given. The first major parameter to determine is *capital cost*. This accounts for all funds needed at the beginning of the project. It includes land and permitting costs, inventory costs, construction costs, and so on. Capital costs can be split into several categories/classifications based on the specific aspects of plant construction or operation. For the purposes of this section, "Total Capital Cost" will be used in lieu of "Total As-Spent Cost," as described in the U.S. Department of Energy Volume 1 report on fossil plants.[47] A rough estimate of total capital cost should be obtained using historical, comparative, and/or cost estimating tools early on in the project life cycle to assess project viability: if the total cost is more than what the investor has on hand and/or could reasonably obtain through a loan, then the project cannot proceed. However, another useful way to represent capital costs is in terms of plant output. For example, power plant projects will usually have the total capital cost expressed per unit power output, as in \$/megawatt (\$/MW) or \$/kilowatt (\$/kW). This has the benefit of eliminating the scaling effect (i.e., larger, more productive plants will be more expensive), allowing the project manager or a potential investor to judge a particular project's costs more relative to another.

Another set of costs to examine are *operating and maintenance (O&M) costs*. O&M costs are, as the name implies, related to those charges associated with operating and maintaining the plant over the course of its expected life. These include labor, fuel, administration, waste disposal, and so forth. O&M costs are usually split into two categories. "Fixed" O&M costs are those that are independent of product delivered, while "variable" O&M costs are directly proportional to product delivered. Predictable fixed O&M costs can be added to the total capital costs to determine the total fixed cost (TFC) of the plant and include items like workers' salaries, administrative costs, and so on. Variable O&M costs essentially represent the total variable cost (TVC) of the plant and include items like feedstock consumption, parasitic electricity loads, and waste disposal costs.[47]

Finally, an economic statistic unique to power plants is the *cost of electricity (COE)*, expressed in \$/kilowatt-hour (\$/kWh) or, less commonly, \$/megawatt-hour (\$/MWh). COE can be understood in several different ways. Firstly, COE is calculated using every single known expenditure projected across the plant's lifespan converted into a single, hourly annuity that begins once the plant starts operating and ends once the plant's lifespan has elapsed. In other words, COE can be thought of as an hourly payment on a "loan" that is scheduled to be completely paid off once the plant has been decommissioned or remodeled. Another way of understanding COE is as the "breakeven" point cost. Since it represents the total cost of the plant normalized by energy output, COE represents the total amount that the plant owner must charge its customers for electricity for the entire project to have a *net present value (NPV)* of \$0.00. If the owners are to make any profit, the *price* of electricity must be higher than the COE. In a regulated market, profit margins are fixed, so plant designers have an incentive to keep COE low to minimize *payback period* (the amount of time needed for the initial investment amount to be returned to the investor). In a deregulated market, companies will be in competition, so keeping COE low will result in greater profit margins for plant owners and investors, as well as give more flexibility in setting electricity prices. Similarly, for gasification plants whose end product is a chemical or fuel, the projected overall cost to produce the product should be calculated and compared against the projected sales price of the expected market. An analysis should be made to ensure that adequate profit can be achieved to obtain the desired rate of return for the project.

Analysts, managers, and investors should consider the following questions before choosing to back a particular plant proposal:

- What is the total capital cost and is it affordable? Is a loan necessary to cover the initial capital investment?
- How does the capital cost compare to the plant's size/throughput? How much capital is necessary to have one unit of production of the desired end product (e.g., \$/kilowatt, \$/[pound/hour])?
- What is the TVC of the plant? How do these costs plus the capital and fixed O&M costs affect the plant's breakeven cost (COE for power plants)?
- Given the answers to each of the previous questions, which potential project is the safest/most lucrative investment available?

2.15 SAFETY REQUIREMENTS

A final consideration that should always be considered during the design, planning, construction, and operating phases is the safety of all personnel and equipment involved in the project. This topic is extensively covered in Chapter 13. As such, this section will provide an introduction and offer general safety/risk assessment considerations that are unique to gasification.

Whether the application is power generation or chemical/fuel production, gasification safety requirements generally fall under the same fields as chemical processing plants (due to the nature of the gasification process itself). The chemical processing industry uses several safety and risk evaluation techniques, which include hazard and operability (HAZOP), Layers of Protection Analysis (LOPA), and SIS studies.[56, 57, 58]

A *HAZOP* study is a structured and systematic technique for identifying potential hazards in a system and potential operability issues that may lead to a product that does not meet current standards. An initial HAZOP study should generally be performed at the design package stage before undertaking a detailed HAZOP study during the full front-end engineering design (FEED). A HAZOP study is performed based on the idea that risk events are caused by elements that deviate from intended design or operational specifications. A set of "guide words" are usually used to identify such deviations by stimulating imaginative thinking during brainstorming (that is, words and phrases like "more," "less," "other than," or "as well as"). A HAZOP study is good for identifying a system's ability to meet user specifications and safety standards, identifying weaknesses in the physical design, confronting potential hazards that are difficult to quantify, assessing weaknesses in automated and manual control systems, and assessing hazards associated with non-standard modes of operation (such as shut down and startup). A HAZOP study does not have a means to rank different identified hazards in terms of severity or importance (though HAZOP teams may integrate elements of another method to do so), cannot be used to assess the effectiveness of current safety precautions, and has no means of assessing hazards that are associated with the interaction of multiple parts of a given system.[58]

A *Layers of Protection Analysis (LOPA)* is often performed alongside or in lieu of a HAZOP study. A LOPA's purpose is to calculate a target *Safety Integrity Level (SIL)* for a later *Safety Instrumented Systems (SIS)* study. A LOPA is considered "semi-quantitative" in that they determine risk factors and the results are expressed as numbers and raw data. The probabilistic values determined are usually rough estimates rather than something measured directly. The first phase of a LOPA is to identify any potential hazards that could occur (these can be taken from a HAZOP study). Second, a *risk factor* is assigned to each potential hazard (how likely it is to cause harm). Finally, an estimate of how much *harm* will be done is assigned, either to individuals present at the time of the (theoretical) incident or to the plant's operational capacity afterwards. Upon identifying the risk and harm factors associated with each hazard, several preventative measures are instituted to ensure that the potential incidents that could cause the identified harm to occur will not happen. The "Layers" of protection refers to the fact that each of these measures is redundant: if the first "layer" fails, a second "layer" is still present that could prevent an incident, and so forth.[57, 59] The result of a LOPA is the required SIL, or the desired target risk level that is deemed acceptable for each identified hazard. This value can then be input to a SIS, a piece of engineering control software (as mentioned above), and/or hardware controls (see Section 2.13) that can keep track of values that present a potential safety risk. The control system can then inform the user of an issue and even take automated steps, as programmed by the development team depending on the situation, to bring the overall system back into a "tolerable" state.[60]

In the case of gasification, the following are noteworthy risks:[56]

1. **Explosions:** Biomass gasification typically occurs in small plants at atmospheric pressure.[34] However, larger (more than 250 MW) plants can require high-pressure entrained flow gasifiers to achieve the necessary throughputs. This results in a stream of high-pressure combustible gases that, in the event of a leak, pose an explosion risk. Special care should be taken around pressure vessels to ensure that the design pressures are met to avoid possible gas leaks.
2. **Fire Hazards:** Gasification systems possess numerous areas where chemical reactions are permitted and encouraged and, ideally, will not occur anywhere else. Like the explosion hazards, care must be taken to prevent leaks in equipment to ensure that chemical reactions, especially highly exothermic ones, do not occur in non-designated areas. At pressures lower than the environment pressure (only likely to happen with low-pressure gasification), oxidant may seep into devices, causing auto-ignition, and a potential fire outbreak. Typical syngas mixtures can auto-ignite at temperatures as low as 600°C (1,112°F).
3. **Toxicity:** Raw syngas contains several compounds that are hazardous to human health if inhaled, such as carbon monoxide, hydrogen sulfide, and even trace amounts of NO_X. The gas tightness of the gasifier and all interconnected systems should be maintained to prevent leaks. In addition, properly ventilating such areas and installing carbon monoxide detectors to alert employees of a potential leak are advised.
4. **Feed material hazards:** The project manager should be aware of health risks associated with the chosen feedstock and take precautions to protect the lives and health of employees that will be exposed to any such hazardous materials.

There are numerous methodologies and considerations to consider when developing a plant safety protocol, including the following:

- What is the best methodology to use to determine potential safety hazards and risks for the current design?
- All industrial plants carry some sort of risk. What risk/severity combinations are acceptable, and which ones require the most improvement?
- Once all risk factors have been identified, what are the simplest, most effective means of reducing all risks to an acceptable level? How many layers of redundancy are needed for each safety solution?
- Ultimately, what can be done to continuously identify new hazards and implement necessary safety procedures to protect the health and safety of all employees and the continued operability of the plant as a whole?

2.16 SUMMARY

This chapter provides an overview of:

- Each of the considerations that designers, owners, and investors must consider when a gasification plant is slated to be built (whether the plant is for power generation or the production of chemicals or fuels).
- An effective methodology for the gasification plant design process.
- A cursory analysis of the various gasification and gasification-compatible power and chemical conversion technologies as well as basic information on how projects are managed and financed.

The remaining chapters cover the topics discussed in this chapter in greater detail and provide an explanation of the various features of gasification-based power and chemical/fuel production plants. Each chapter focuses on a specific module and the various design decisions. Chapter 13 focuses on the design of various ancillary systems and other key supporting plant elements, including plant safety.

It is important to note that it is critical that the scope of any gasification project (whether for an individual project or for design of a modular system that is intended to be replicated multiple times) be defined as completely and accurately as possible at the front end of the project and then locked down to the maximum extent possible for the remainder of the project completion. Scope changes, sometimes referred to as scope creep, in the later stages of a project are one of the major causes of project cost escalations and ultimately of project failure.

Endnotes

1 M. Godinho, N.R. Marcilio, A.C.F. Vilela, L. Masotti, C.B. Martins (2007) "Gasification and Combustion of the Footwear Leather Wastes", Journal of the American Leather Chemists Association 102(6):182-190.

2 P. Alvaerz-Alvarez, A. Pizarro, M. Barrio-Anta, A. Camara-Obregon, J.LM. Bueno, Gutierrez, D.F.R.P. Burslem (2018) "Evaluation of Tree Species for Biomass Energy Production in Northwest Spain", Forests 9(4): 160. *https://www.mdpi.com/1999-4907/9/4/160*.

3 M.A. Rasheed, P.L. Srinivasa Rao, A. Boruah, S.Z. Hasan, A. Patel, V. Velani, K. Patel (2015) "Geochemical Characterization of Coals Using Proximate and Ultimate Analysis of Tadkeshwar Coals, Gujarat", Geosciences 5(4):113-119. doi:10.5923/j.geo.20150504.01.

4 H.J. Gluskoter (1977) "Inorganic Sulfur in Coal", Energy Sources 3:125-131. doi:10.1080/00908317708945974.

5 T. Leustek and K. Saito (1999) "Sulfate Transport and Assimilation in Plants", Plant Physiology 120(3):637-644. doi:10.1104/pp.120.3.637.

6 EPRI (December 2016) Program on Technology Innovation: Alternative and Innovative Technologies for Coal Combustion Product Management. EPRI, Palo Alto, CA: DCN SE06968.

7 G. Zając, J. Szyszlak-Bargłowicz, W. Gołębiowski, M. Szczepanik, (2018) "Chemical Characteristics of Biomass Ashes", Energies 11(11):2885. *https://www.mdpi. com/1996-1073/11/11/2885/htm*..

8 M.I. Stewart, Jr. (2014) "Surface Production Operations, Volume 2: Design of Gas-Handling Systems and Facilities", Chapter 3: Heat Transfer Theory. 3rd Ed., Gulf Professional Publishing: Houston, TX, U.S. doi:10.1016/C2009-0-64501-3.

9 EPA (2019) "Clean Air Act Requirements and History", *https://www.epa.gov/clean-air-act-overview/clean-air-act-requirements-and-history*. Accessed December 10, 2019.

10 J. Ratafia-Brown, L. Manfredo, J. Hoffmann, M. Ramezan (2002) "Major Environmental Aspects of Gasification-Based Power Generation Technologies", Final Report. Science Applications International Corporation (SAIC) prepared for NETL. *https://www.netl.doe.gov/sites/default/files/netl-file/final-env.pdf*.

11 EPA (2019) "Summary of the Clean Water Act", *https://www.epa.gov/laws-regulations/summary-clean-water-act*. Accessed December 10, 2019.

12 EPA (2019) "Summary of the Resource Conservation and Recovery Act", *https://www.epa.gov/laws-regulations/summary-resource-conservation-and-recovery-act*. Accessed December 10, 2019.

13 Energy Watch (n.d.) "Regulated vs. Deregulated Markets", *https://energywatch-inc.com/regulated-vs-deregulated-electricity-markets/*. Accessed December 11, 2019.

14 NETL (2019) "Gasifipedia. 10.3. Syngas Conversion to Methanol." *https://www.netl.doe.gov/research/coal/energy-systems/gasification/gasifipedia/methanol*. Accessed December 11, 2019.

15 A. Dutta and S.D. Phillips (2009) "Thermochemical Ethanol via Direct Gasification and Mixed Alcohol Synthesis of Lignocellulosic Biomass", NREL: Golden, CO. Technical Report. NREL/TP-510-45913, *https://www.nrel.gov/docs/fy09osti/45913.pdf*.

16 NETL (2019) "Gasifipedia. 7.5. SNG from Coal: Process & Commercialization", *https://www.netl.doe.gov/research/coal/energy-systems/gasification/gasifipedia/coal-to-sng-process*. Accessed December 11, 2019.

17 R.D. Brasington, N.J. Kuehn, E.G. Lewis, L.L. Pinkerton, M. Turner, E. Varghese, M. Woods (July 2011) "Cost and Performance Baseline for Fossil Energy Plants Volume 2: Coal to Synthetic Natural Gas and Ammonia", DOE/NETL. DOE/NETL-2010/1402, *https://www.netl.doe.gov/sites/default/files/netl-file/SNGAmmonia_FR_20110706.pdf*.

18 E.H.G. Morris (1935) "Liquid Fuel: Its Application, Uses, and Advantages." Proceedings of the Victorian Institute of Engineers 1933-1934. 17:14-34, 60, *https://digitised-collections.unimelb.edu.au/handle/11343/24777*.

19 R. Jones, J. Goldmeer, B. Monetti (June 2011) "Addressing Gas Turbine Fuel Flexibility", GE Energy. Report No. GER4601. Revision B. *https://www.ge.com/content/dam/gepower-new/global/en_US/downloads/gas-new-site/resources/reference/ger-4601b-addressing-gas-turbine-fuel-flexibility-version-b.pdf*.

20 NETL (2019) "Gasifipedia. 10.2. Fischer-Tropsch Synthesis", *https://www.netl.doe.gov/research/coal/energy-* systems/gasification/gasifipedia/ftsynthesis. Accessed December 11, 2019.

21 V. Shah, N.J. Kuehn, M.J. Turner, S.J. Kramer (October 2014) "Cost and Performance Baseline for Fossil Energy Plants Volume 4: Coal-to-Liquids via Fischer-Tropsch Synthesis", DOE/NETL, DOE/NETL-2011/1477, *https://netl.doe.gov/projects/files/CostandPerformanceBaselineforFossilEnergyPlantsVolume4CoaltoLiquidsviaFTSynthesis_101514.pdf*.

22 Hydrogenics (2019) "Fuel Cells", *https://www.hydrogenics.com/technology-resources/hydrogen-technology/fuel-cells/*, Accessed December 12, 2019.

23 J. Rezaiyan and N.P. Cheremisinoff (2005) Gasification Technologies: A Primer for Engineers and Scientists. Taylor and Francis: New York, NY.

24 T.B. Reed and S. Gaur (2001) "A Survey of Biomass Gasification 2001", 2nd Ed., NREL and The Biomass Energy Foundation: Golden, CO.

25 C. Turare (1997) "Biomass Gasification – Technology and Utilisation", ARTES Institute, University of Flensburg: Flensburg, Germany, *http://cturare.tripod.com/bio.htm*.

26 J. Loganbach, P.V. Smith, R. Leonard, L. Rogers, P. Vimalchand, G. Liu (2001) "Development Status of the Transport Gasifier at the PSDF", Gasification Technologies 2001, NETL, Southern Company Services, and Kellogg Brown & Root, Inc.: San Francisco, CA.

27 Y. Richardson, M. Drobek, A. Julbe, J. Blin, F. Pinta (2015) "Chapter 8: Biomass Gasification to Produce Syngas", in Recent Advances in Thermo-Chemical Conversion of Biomass. Elsevier: Amsterdam, the Netherlands. pp. 213-250.

28 C. Mandl, I. Obernberger, F. Biedermann (2009) "Updraft Fixed-Bed Gasification of Softwood Pellets: Mathematical Modeling and Comparison with Experimental Data", Proceedings of the 17th European Biomass Conference and Exhibition. Hamburg, Germany.

29 Food and Agriculture Organization of the United Nations (1986) "Wood Gas as Engine Fuel", Forestry Department. Paper 72. Rome, Italy. *http://www.fao.org/docrep/t0512e/T0512e00.htm*.

30 United States Patent and Trademark Office (February 2017) CPC Definition - Subclass C10J, *https://www.uspto.gov/web/patents/classification/cpc/html/defC10J.html*.

31 F. Wallace, X. Guan, R. Leonard, M. Nelson, P. Vimalchand, W. Peng, P.V. Smith, R.W. Breault (2006) "Operation of the Transport Gasifier at the PSDF", Proceedings of the 31st International Technical Conference on Coal Utilization and Fuel Systems. Clearwater, FL, USA.

32 Thermo Electron Corporation (2005) "Safe Operation of Cryogenic Air Separation Unit Using a Process FT-IR Gas Analyzer", Public Company Papers. PI.6009.1105.

33 H. Ishii (2010) "A Technical Update of the MHI Air Blown and Oxygen Blown Gasifier", Mitsubishi Heavy Industries. Proceedings of the 2010 Gasification Technology Conference. Washington, DC.

34 H. A. Long, III and T. Wang (2011) "Analysis of Biomass/Coal Co-Gasification for Integrated Gasification Combined Cycle (IGCC) Systems with Carbon Capture," ECCC Report 2011-01, Energy Conversion and Conservation Center, University of New Orleans. New Orleans, LA.

35 J. Perritano (2013) "How Cyclonic Separation Works." How Stuff Works. *http://electronics.howstuffworks.com/everyday-tech/cyclonic-separation1.htm#*, Accessed October 27, 2016.

36 EPA (2002) "EPA Air Pollution Control Cost Manual." 6th Ed. Report No. EPA/452/B-02-001. *https://www3.epa.gov/ttncatc1/dir1/c_allchs.pdf*.

37 DOE (September 2013) "Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 2a", DOE/NETL, DOE/NETL-2010/1397.

38 NETL (2016) "Gasifipedia. 6.2.3. Particulate Removal", *https://www.netl.doe.gov/research/coal/energy-systems/gasification/gasifipedia/particulate-removal*, Accessed January 10, 2016.

39 H.A. Long, III and T. Wang (2016) "Parametric Techno-Economic Studies of Coal/Biomass Co-Gasification for IGCC Plants with Carbon Capture Using Various Coal Ranks, Fuel-Feeding Schemes, and Syngas Cooling Methods", International Journal of Energy Research 40(4):473-496, *https://onlinelibrary.wiley.com/doi/pdf/10.1002/er.3452*.

40 K. Kubek (2009) "Large CO_2 Sources and Capture Systems." Workshop on Future Large CO_2 Compression Systems. Sponsored by EPRI, DoE, & NETL. Gaithersburg, MD, U.S.A. *https://www.nist.gov/system/files/documents/pml/high_megawatt/2_3-Kubek-Approved.pdf*.

41 NETL (2019) "Gasifipedia. 6.2.1. Acid Gas Removal (AGR): MDEA", *https://www.netl.doe.gov/research/coal/energy-systems/gasification/gasifipedia/mdea*.

42 NETL (2019) "Gasifipedia. 6.2.4. Mercury/Toxics Emissions", *https://www.netl.doe.gov/research/Coal/energy-systems/gasification/gasifipedia/mercury-toxics-emissions*, Accessed December 13, 2019.

43 T. Wall (2005) "Fundamentals of Oxy-Fuel Combustion", IEA Greenhouse Gas R&D Programme. Oxy-fuel Combustion Research Network, Cottbuss, Germany. *http://www.ieaghg.org/docs/oxyfuel/w1/04W1Wall.pdf*.

44 D. Daly (2018) "Energy and CO_2 Management: Carbon Capture and Storage", 2018 Lignite Education Seminar. Plains CO_2 Reduction Partnership: Bismarck, ND. *https://cdn.lignite.com/wp-content/uploads/2018/06/12-CO_2-Daly-FINAL-DA.pdf*.

45 D. Nield (November 23, 2016) "Scientists are Trapping CO_2 and Turning It into Stone", Science Alert. *https://www.sciencealert.com/scientists-are-again-turning-co2-into-solid-rock-to-fight-climate-change*.

46 M. Stewart, Jr. (2014) "Chapter 9: Gas Sweetening", in Surface Production Operations 3rd Ed. Volume 2, pp. 433-539. Gulf Professional Publishing: Houston, TX. doi:10.1016/C2009-0-64501-3.

47 R. James, A. Zoelle, D. Keairns, M. Turner, M. Woods, N. Kuehn (September 2019) "Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 4", NETL-PUB-22638.

48 The Linde Group (2018) "CO_2 Purification and Liquefaction", *https://www.linde-engineering.com/en/process-plants/co2-plants/co2-purification-and-liquefaction/index.html*.

49 Z. Abbas, T. Mezher, M.R.M. Abu-Zahra (2012) "Evaluation of CO_2 Purification Requirements and the Selection Processes for Impurities Deep Removal from the CO_2 Product Stream", Proceedings of the 11th International Conference on Greenhouse Gas Control Technologies. Kyoto, Japan. Energy Procedia 37:2389-2396 doi:10.1016/j.egypro.2013.06.120.

50 S. Mills (2019) "Modularisation for Clean Coal", IEA Clean Coal Center. Report CCC/299. London, England, U.K.

51 N. Rasmussen and S. Niles (2011) "Modular Systems: The Evolution of Reliability", Schneider Electric – Data Center Science Center. White Paper 76, Revision 1. *https://download.schneider-electric.com/files?p_Doc_Ref=SPD_SNIS-66ZTJB_EN*.

52 W.J. Howard, W. Satin, R. Farr, and J. Flenge (March 9, 2017) "Using Advanced Analytics and Controls", Power Engineering 121(3). *https://www.power-eng.com/2017/03/09/using-advanced-analytics-and-controls/*.

53 V. Muthukrishnan (October 22, 2019) "Programmable Logic Controllers (PLCs): Basics, Types, and Applications", Electrical 4 U. *https://www.electrical4u.com/programmable-logic-controllers/*.

54 S. Melendez (November 2, 2018) "Advantages & Disadvantages of Distributed Systems", Techwalla. *https://www.techwalla.com/articles/advantages-disadvantages-of-distributed-systems*.

55 J. Papiewski (June 30, 2019) "What is the Difference Between SCADA and DCS Systems?", Bizfluent. Reviewed by Jayne Thompson. *https://bizfluent.com/facts-7588416-difference-between-scada-dcs-systems.html*.

56 Waste Management World (2012) "Gasification: Refining Safety", https://waste-management- *world.com/a/gasification-refining-safety*, Accessed December 20, 2019.

57 M. Broadribb and M.R. Currie (2010) "HAZOP/LOPA/SIL – Be Careful What You Ask For!", AIChE Spring Meeting and Global Congress on Process Safety. Paper 22a. March 21-25, 2010. San Antonio, TX.

58 Product Quality Research Institute (n.d.) "HAZOP Training Guide", Manufacturing Technology Committee – Risk Management Working Group, *http://pqri.org/wp-content/uploads/2015/08/pdf/HAZOP_Training_Guide.pdf*.

59 J. Merriman (April 8, 2013) "LOPA – Layer of Protection Analysis Technique", Industrial Plant Safety: Free Industrial Safety Information/News. *https://industrialplantsafety.com/lopa.html*.

60 D. Yoset (2017) "Safety Instrumented Systems vs. Process Control Systems", Cross Company. Process Control Systems Integration Blog. *https://www.crossco.com/blog/safety-instrumented-systems-vs-process-control-systems/*.

3.0 GUIDELINES FOR THE DESIGN OF A GASIFICATION MODULE

3.1 INTRODUCTION

This chapter focuses on the design of the gasifier module in accordance with the objectives of the overall gasification plant to convert raw solid fuel into power (i.e., electricity) and/or liquid or gaseous fuels/chemicals. The emphasis is on three types of medium-to-large scale plants—an integrated gasification combined cycle (IGCC) power plant, an F-T plant, and a substitute natural gas (SNG) plant. Smaller, modular power or liquid fuel plants whose designs can be standardized for shop fabrication are also considered.

The goal of the chapter is to provide users with a set of guidelines to assist in developing a design package that can be used by an Architecture and Engineering (A&E) firm and/or gasifier vendor for detailed design/engineering of the gasifier. The guidelines enable the user to select/narrow the options for a suitable gasification technology that aligns with the anticipated plant size (i.e., capital investment), location, and feedstock availability to enable the production of a suitable end product (i.e., power and/or liquid fuels/chemicals/substitute natural gas/hydrogen) in a cost-effective manner. Alternatively, if the user only knows what and how much of the end product(s) is(are) needed and the current cost to obtain them (if they can be sourced at all), the guidelines help to determine a suitable gasification technology, location, and feedstock.

A simple methodology to carry out the selection of a gasifier type is presented alongside a mass balance and input/output requirements for the gasifier, including feedstock pretreatment requirements for the selected gasifier (if any). Recommendations are included for the required analytical, control, and operator safety requirements for gasifier startup, normal operation, and shutdown. Water requirements, treatment, and discharge regulatory requirements are discussed, in addition to suggestions for disposal methods for solid waste/slag in an environmentally safe manner.

3.2 GASIFICATION FUNDAMENTALS

Gasification consists of a complex set of reactions of a carbon-based fuel with steam and air (or oxygen). Gasification has been the subject of several reviews, including those by Johnson,[1] Reed,[2] Probstein and Hicks,[3] Higman and van der Burgt,[4] Knoef,[5] Rezaiyan and Cheremisinoff,[6] Bell et al.,[7, 8] and the National Energy Technology Laboratory.[9] In contrast to a boiler where essentially complete combustion of carbonaceous material to carbon dioxide (CO_2) and water (H_2O) occurs, the reactions that occur during gasification are moderated to produce a syngas containing as much carbon monoxide (CO) and hydrogen (H_2) as possible, while preserving as much of the initial energy content (heating value) of the fuel as possible. Syngas is a valuable and flexible building block that can be used to produce a variety of products, including heat, power, liquid transportation fuels, SNG, hydrogen, and/or chemicals. Thus, gasification is a flexible technology not only with respect to the choice of feed, but also with respect to the end products. The steps of fuel particle transformation that occur during gasification include drying, devolatilization, pyrolysis, oxidation, and char gasification. The vaporization of water associated with the initial drying step can significantly influence gasification thermodynamics when high-moisture feedstock is gasified or when the feedstock is fed to the reactor as a slurry.[7]

The onset of devolatilization is nearly simultaneous with pyrolysis and occurs at temperatures above 325°C (617°F). The organic carbon-carbon, carbon-sulfur, and carbon-nitrogen bonds break to form unstable molecules that further react or pyrolyze to primarily produce a mixture of gases (H_2, CO, CO_2, etc.), phenols, aromatics, and lighter compounds like methane, as well as tar, soot, and char. The sulfur and nitrogen in fuel can result in the formation of gaseous contaminants, such as hydrogen sulfide (H_2S), carbonyl sulfide (COS), and ammonia (NH_3). Oxygen is fed to the gasifier, either as a nearly pure stream of oxygen, or as air. Depending on the proximity of the fuel and oxygen (or air) feed points, the volatiles may crack and/or oxidize to form combustion and partial combustion products and heat. Oxygen also reacts with char to produce CO, CO_2, and heat. The heat increases the reaction environment temperature to 850 to 1,300°C (1,562 to 2,372°F) (depending on the type of gasifier) and drives the endothermic char-steam (hydrogasification), char-CO_2 (Boudouard reaction), and steam reforming reactions to produce CO and H_2.

$$\text{C (char)} + H_2O \rightarrow CO + H_2 \qquad \text{(R3.1)}$$

$$\text{C (char)} + CO_2 \rightarrow 2C \qquad \text{(R3.2)}$$

$$CxHy + xH_2O \rightarrow xCO + (1/2\ y + x)\ H_2 \qquad \text{(R3.3)}$$

The water-gas shift (WGS) reaction ($CO + H_2O \rightarrow CO_2 + H_2$) is another important reaction that occurs during gasification that influences the exit syngas H_2/CO ratio. The extent of WGS depends on the temperature and the degree of contact between the syngas and char, with the char acting as a catalyst for the reaction. However, below a certain temperature, WGS becomes very slow, so the ratio no longer significantly changes. This temperature, where the WGS reaction essentially “freezes,” is generally different for different gasifiers depending on their syngas-char contacting pattern. Combined with differences in steam inputs and different feedstock species and compositions, this results in different H_2/CO ratios in the exiting syngas from different gasifiers.

Appropriate amounts of fuel steam, and air (or pure oxygen) need to be fed to the gasifier to achieve a steady state temperature profile and a self-sustaining operation that does not require indirect heating. The optimum oxygen/fuel and oxygen/steam ratios are different for different gasifiers. As discussed in Chapter 2, gasifiers can have various flow configurations, including downdraft, updraft, fluidized bed, entrained flow, and transport. The choice of a gasifier technology (discussed in Section 3.4.1) and its flow configuration significantly influences both the content of tar and methane in the syngas and the relative amounts of oxygen and steam that are used. For example, in an updraft moving-bed gasifier, pyrolysis occurs at relatively low temperatures, typically 600 to 700°C (1,112 to 1,292°F). Fuel and oxygen flow through the reactor in a countercurrent fashion, so nearly all the oxygen is consumed before the pyrolysis section of the gasifier. Consequently, this type of gasifier produces substantial quantities of aromatics, tar, and methane. On the other hand, entrained flow gasifiers, which typically have much higher O_2/H_2O ratios than updraft moving-bed gasifiers, operate at temperatures greater than 1,300°C (2,372°F) and produce no tar and very little methane.[7]

The simplest and least expensive source of oxygen for gasification is compressed air. However, this introduces a large quantity of nitrogen that dilutes the resulting syngas. Consequently, air-blown gasifiers produce a low British thermal unit (Btu) gas (150 to 175 Btu/standard cubic foot [scf]), whereas oxygen-blown and steam-blown gasifiers produce a medium Btu gas (300 to 350 Btu/scf) containing much less nitrogen. Most of the large gasifier plants include a cryogenic distillation-based air separation unit (ASU) to provide oxygen for gasification. The ASU is a substantial fraction of the overall gasification plant capital cost, yet, at large scales, its unit costs are tractable and enable operation of efficient, large, slurry-fed entrained flow gasifiers that can efficiently produce large quantities of medium-Btu syngas. However, these advantages do not apply at small scales. In addition, the ASU often accounts for a large portion (60 to 70%) of a given plant's total auxiliaries. Therefore, the simplicity of air-blown or steam-blown gasifiers is attractive for small-to-medium-sized gasification plants, particularly if the syngas is to be burned for heat or power generation. Another option is to split the difference between the oxygen-blown and air-blown options by using a lower-cost air separation unit, such as vacuum swing adsorption (VSA), that generates oxygen in the range of 90 to 93% purity (if the facility size is small enough for a VSA to provide oxygen [see Chapter 5]).

3.3 OVERALL PLANT CONSIDERATIONS FOR DESIGN OF GASIFIER MODULE

To design a gasification module, an evaluation of the entire plant should be undertaken first to narrow the choices for the gasification technology/vendors and to determine a value proposition. Although plants can be classified as mini, small, medium, or large, the classification is somewhat arbitrary. A suggested classification is provided in Table 3-1, based on the possibility that gasifiers for mini to medium plants can be modular, standardized, and factory built to reduce the impact of economies of scale by allowing the economic use of several gasifiers running in parallel, whereas large plants have to be custom designed and built from scratch.

Table 3-1. Plant size classification and corresponding feed, equipment, and products

SIZE CLASSIFICATION		MINI	SMALL	MEDIUM	LARGE
Approximate fuel feed rate required, tons/day (tpd)[1]		3 to 14	60 to 315	400 to 700	1,100 to 4,500
PRIMARY END PRODUCT	POTENTIAL EQUIPMENT REQUIRED; BYPRODUCTS				
Power[2]	IGCC (gas and steam turbines), Gas cleanup; sulfur, CO_2				~150 to 550 MWe
	IGCC (gas and steam turbines); Gas cleanup, sulfur, CO_2			~50 to 80 MWe	
	Gas turbine, Gas cleanup; process heat, steam, and/or hot water)		~5 to 25 MWe		
	Gasifier-IC engine, Gas cleanup; process heat	0.2 to 1 MWe			
	Fuel Cells[10]	Up to 0.4 MWe			
Process heat, steam, or hot water	Syngas burner, heat exchanger (cold gas efficiency—~80%; HX efficiency— ~98%)	1-5 MBtu/h	21-113 MBtu/h	NP	NP
Liquid fuel, methanol, SNG[3]	Shift reactor, CO_2 separator, catalytic converter, product upgrader/separator	NP	120-630 bpd[4]	800-1,400 bpd[4]	2,200-9,000 bpd[4]

NP-Not practical
1 Assumes an average fuel heating value of 11,000 Btu/lb, as fed
2 Electrical efficiency; Mini~27 %; Small~30 %; IGCC~45% (without CO_2 capture)
3 ~Based on ~2 barrels gasoline equivalent/ton of fuel
4. bpd—barrels/day gasoline equivalent

Essential plant characteristics/parameters should be considered and determined even at this initial process design stage. These include plant size (i.e., investment), plant location, feedstock, end products (i.e., power, heat, fuels, and/or chemicals), CO-shift requirements, syngas cleanup requirements, CO_2 capture requirements, and power producing equipment size compatibility (e.g., gas turbine) or syngas to liquids/chemicals/SNG reactor and product upgrading/separation system. These evaluations enable a preliminary determination of a suitable air- or oxygen-blown gasifier technology module. Also, at this preliminary stage, the investors should at least be aware of the economies of scale, gas cleanup requirements, byproducts, environmental and other impacts associated with the land where the plant would be located, and its proximity to the fuel source and market for end products. Each of the above characteristics are discussed and evaluated in the following subsections (Table 3-1 is used as the reference point). Based on these evaluations, a procedure is suggested to determine a preliminary overall value proposition to enable a decision whether to proceed with a more detailed gasifier module design.

3.3.1 PLANT SIZE AND LOCATION

The most important factors the investor(s) needs to initially consider are the plant's size and location, which are interrelated to an extent. The land needs to be of sufficient area and accessibility while possessing the necessary qualities to allow for the construction of the plant and its infrastructure without creating environmental issues and hardships to the associated community and other nearby industry.

3.3.1.1 PLANT SIZE

Table 3-1 provides a suggested size classification (i.e., mini, small, medium, and large) based on power produced. The amount of power produced is then used to determine the suitable type of power producing equipment (i.e., IGCC, fuel cell bank, gas turbine, or internal combustion [IC] engine). The table then assumes a fuel heating value of 11,000 Btu/lb and electrical efficiencies depending on the power producing device to determine the amount of fuel required. Then, based on the fuel input, plants that could be suitable for producing end products other than power, such as heat or fuels/chemicals/SNG products, are designated. Note that mini to small plants can possibly be suitable for heat (or electricity using a steam Rankine cycle) as the primary product, whereas small to large plants could be suitable for producing liquids. For a liquid fuel/SNG producing plant, a total yield of approximately 2 barrels of gasoline equivalent/ton of fuel is assumed to determine the amounts that can be produced.

To take advantage of economies of scale, the largest plant should be selected that can be located on the available (or potential) land and built using the available cash and other investment. However, a smaller modular plant could also be considered to reduce financial risk. Cash from the investor(s) combined with potential grants and loans from financial institutions; any other incentives (e.g., CO_2 capture credits); and/or the fuel cost, quality, and quantity available can be used to estimate a preliminary plant size (see Section 3.3.4).

3.3.1.2 LOCATION/LAND

Many factors dictate the choice of the land with associated cost considerations. The most important factors are provided in Table 3-2.

Table 3-2. Important factors in selection of plant location

IMPORTANT FACTORS IN SELECTION OF PLANT LOCATION	Land ownership
	Land size suitable for contemplated plant size
	Vicinity to feedstock under consideration; feedstock ownership
	Transportability of fuel—rail and road access
	Fuel cost and quality, pretreatment requirements
	Greenfield or brownfield
	Suitability of available infrastructure if brownfield
	Access to power grid and water supply
	Products selling price and demand in vicinity
	Access to supply chain for main product and byproducts
	Buy in from nearby township and industry
	Treatment and disposal capability for wastewater, solid waste, and other waste streams
	Environmental considerations and permitting requirements

If the investor(s) owns suitable land, it would generally be economically favorable to use that land for the plant provided it has sufficient space to house the plant and is conveniently accessed by road and/or rail car (for large plants). Feedstock ownership by the investor(s) is preferable and the selected land should preferably be in the vicinity (i.e., no more than 100 miles away) of the source to reduce transportation costs. The feedstock cost and quality are also important and will be discussed in the following section.

Some factors in Table 3-2 are self-explanatory and do not require further discussion. However, product demand, selling price, and access to the supply chain should be noted as very important considerations for plant profitability. For example, even a mini to small power plant could be profitable in remote locations that either have access to sources of cheaper fuels and/or where power is expensive (e.g., Hawaii and Alaska [USA]). Similarly, a small to medium CTL plant could possibly be profitable in certain foreign countries or near a U.S. military site that can provide the land and is paying higher prices for conventionally sourced liquid fuel, or a small to large SNG plant could be profitable in certain foreign locales like China, Japan, Taiwan, or South Africa where natural gas is not significantly available and import prices are high (more than $10 USD/MMBtu).

3.3.2 FEEDSTOCK

The most important properties of gasifier feedstocks that dictate their selection are the “as received” heating value (which decreases with increases in its mineral matter, oxygen, and moisture content), sulfur content, and ash fusion temperature. All these properties can be measured using standard American Society for Testing Materials (ASTM) methods. Biomass is noteworthy due to its low density and high elasticity. These properties can be overcome through the use of appropriate pretreatment methods like torrefaction, as will be discussed in Chapter 4.

Simple pretreatment methods are generally desirable for all types of gasifiers to reduce the impurities. These methods not only increase the feedstock heating value; they can also somewhat reduce the production of sulfur containing gases, such as H_2S and COS, during gasification that need to be reduced to low parts per million (ppm) to parts per billion (ppb) values via syngas cleanup to protect downstream equipment. However, any pretreatment adds to the cost of the feedstock so there is a tradeoff. At the gasifier site, conveying equipment, comminution/sizing equipment, and a dryer are the essentials needed to deliver the feedstock in the required particle size range and moisture content to the gasifier.

Beyond the previous factors discussed in this section, feedstock selection is also dictated by gasifier choice, and, in turn, gasifier choice is dictated by the plant size. Gasifier choices and corresponding designs will be discussed in Section 3.4.1. Also, as discussed in Chapter 2, gasifiers can have various flow configurations including downdraft, updraft, fluidized bed, entrained flow, and transport. A downdraft (i.e., co-current downward flow of feedstock and oxidant) moving-bed gasifier is preferable for mini plants, whereas an updraft (i.e., countercurrent) moving-bed or fluidized-bed gasifier (i.e., bubbling or circulating) is preferable for small to medium-sized plant.

For fuel ash content, melting it to form a slag requires the feedstock to have an ash fusion temperature that is below the operating temperature of the gasifier.[1] Ash begins to soften before fully melting, because it is a complex and variable mixture of alumino-silicate minerals with numerous other constituents that melt over a range of temperatures. Fluidized bed gasifiers are typically compatible with fuels that have ash softening temperatures above 1,100°C (2,012°F) and operate around 950 to 1,000°C (1,742 to 1,832°F) to achieve high conversion while avoiding temperatures that would make the ash soften and stick to the side walls, resulting in poor fluidization. On the other hand, slag forming gasifiers, such as the Texaco/General Electric (GE) and Shell entrained flow gasifiers (both now owned by Air Products), the DOW/Destec/E-GasTM entrained flow gasifier (in the process of being sold to The Chatterjee Group by current owner, McDermott [at the time of publication]), and the British Gas/Lurgi (BGL) moving-bed gasifier, operate at temperatures above 1,300°C (2,372°F) so that the slag will be sufficiently fluid.

3.3.3 SYNGAS UPGRADING

Syngas upgrading (discussed in Chapter 6) refers to a train of various unit operations and processes that are carried out to treat and clean the raw syngas to make it suitable for production of the desired end products. The train depends not only on the plant size, end product, gasifier, and equipment/operation producing the end product (Table 3-1), but also on the need to shift the syngas using sour WGS to allow higher CO_2 capture and/or achieve the desired H_2/CO ratio for chemical/fuel production (Table 3-3).

Table 3-3. Syngas upgrading steps for various plant configurations

PLANT SIZE	END PRODUCTS	SUGGESTED GASIFIER	CO_2 CAPTURE	END-PRODUCT EQUIPMENT	SUGGESTED SYNGAS UPGRADING/CLEANUP TRAIN
Mini	Power, heat	Downdraft sub-atmospheric, air-blown	No	IC engine	Cyclone, quench, desulfurizer (if necessary), trace tar and siloxane removal
Mini	Heat	Downdraft sub-atmospheric, air-blown	No	Close-coupled syngas burner	Cyclone
Small	Heat	Updraft atmospheric, air-blown	No	Close-coupled syngas burner	Cyclone
Small	Power, heat	Updraft, fixed or fluidized, pressurized, air-blown	No	Gas turbine	Cyclone, quench, tar scrubber, desulfurizer
Small	Liquid fuel, methanol, SNG	Updraft fixed or fluidized, pressurized, air-blown	No	Catalytic converter, liquid upgrading	Cyclone, quench, tar scrubber, sour-shift (if necessary), desulfurizer, sulfur guard bed
Medium	Power	Updraft fixed or fluidized, pressurized, air-blown	Yes	IGCC (Gas and steam turbine)	Cyclone, quench, tar scrubber, sour-shift (as necessary), desulfurizer and CO_2 capture, Claus sulfur recovery
Medium	Liquid fuel, methanol, SNG	Updraft fixed or fluidized, pressurized, air-blown	Yes	Catalytic converter, liquid upgrading	Cyclone, quench, tar scrubber, sour-shift (as necessary), desulfurizer, CO_2 capture, Claus sulfur recovery, sulfur guard bed
Large	Power	Entrained, pressurized, oxygen-blown	Yes	IGCC	Cyclone (as necessary), quench (or coolers and candle filter), sour-shift, desulfurizer, CO_2 capture, Claus sulfur recovery
Large	Liquid Fuel, methanol, SNG	Entrained, pressurized, oxygen-blown	Yes	Catalytic converter, liquid upgrading	Cyclone (as necessary), quench (coolers and candle filter), sour-shift, desulfurizer, CO_2 capture, Claus sulfur recovery, guard bed

Air-blown gasifiers are suggested for plants ranging from mini to medium, whereas oxygen-blown gasifiers are suggested for large plants, because an ASU using cryogenic distillation does not scale down economically, making them unsuitable for smaller-scale plants. It is also noteworthy that mini plants (less than 1 megawatt-electric [MWe]) are sub-atmospheric, and a downdraft fixed-bed gasifier is recommended to allow bulk tar destruction within the gasifier itself. Also, small plants that only produce heat or steam for a Rankine cycle are suggested to run at atmospheric pressure, since pressurized gasifiers are significantly more complex to design and operate. On the other hand, a pressurized updraft moving-bed or bubbling fluidized-bed gasifier (even though they make significant tar) are recommended for power-producing small to medium plants, not only to reduce equipment size and overall plant cost, but also because of the difficulty of scaling up downdraft gasifiers above 1 MWe. Pressurized entrained flow gasifiers are best suited for large plants due to their high throughput for a given size/cost. However, certain dry entrained flow gasifier designs (such as Mitsubishi Heavy Industries [MHI]) could also be applied to smaller plants. Entrained flow gasifiers run at very high temperatures (more than 1,300°C/2,372°F) and essentially eliminate all tar in syngas.

Furthermore, as shown in Table 3-3, the number of steps in the syngas upgrading train increase as a function of end products (i.e., liquids > power > heat) and plant size (i.e., large ≈ medium > small > mini). It is envisioned that mini to small power plants will require some desulfurization even with low-sulfur fuels to protect the downstream engine or gas turbine and will not require Claus sulfur recovery. Sour WGS is included in the syngas upgrading train for small plants that produce liquids/SNG and all medium to large plants. Shift is needed to (1) adjust the H_2/CO mole ratio to make it suitable for the desired end product and (2) increase the CO_2 concentration in syngas to reduce the cost of CO_2 capture (and increase the capture rate) before the converter or gas turbine.

3.3.4 VALUE PROPOSITION

It is recommended that investors considering building a gasification-based power or liquid fuel/methanol/SNG plant first implement a preliminary techno-economic analysis (TEA) using a qualified engineer on their staff or with the help of a qualified consulting engineer. Once they have ensured that there is a good probability of a reasonable return on investment, they can make a more informed decision to proceed with a preliminary process design package that is a suitable starting point for an A&E firm to develop a more detailed design, including higher level TEAs and an environmental impact assessment.

The information provided in previous sections is intended to assist investors in carrying out a preliminary TEA for their intended gasification plant and narrowing the choices of the plant size, gasification technology, feedstock, and location. The profitability of the plant depends not only on its size and type, but also on several location specific factors. The preliminary TEA may help identify a location to maximize the potential advantages of these factors.

TEA studies for gasification plants have been published by many research groups for several scenarios. Recent reports and research publications include those by the National Energy Technology Laboratory (NETL),[9, 11] Porcu et al.,[12] Ozonoh et al.,[13] Mantripragada and Rubin,[14] Yu and Chiei, et al.,[15] Martelli et al.,[16] Jiang and Bhattacharyya,[17] Man et al.,[18] Lu et al.,[19] Chen et al.,[20] Sciazko and Chmielniak,[21] International Energy Agency Energy Technology Systems Analysis Program,[22] Cormos and Cormos,[23] and Trippe et al.[24] These references cover plants ranging from mini biomass gasification power plants to large-scale coal/biomass to liquids (CBTL) plants and provide a starting point for the investors to evaluate the potential for profit through the preliminary TEA.

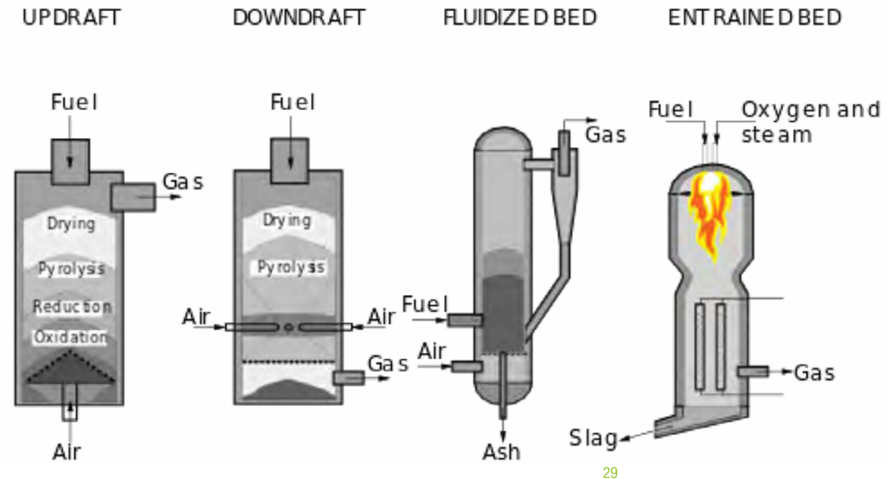
The investors should also be aware of the essential elements that should be included in a more rigorous calculation of profitability and payback period. The economic evaluation of a process proceeds in several steps,[25] including (1) preparing a process flow diagram, (2) conducting mass and energy balances, (3) sizing major equipment, (4) estimating the capital cost, (5) estimating the operations and maintenance (O&M) or production costs, (6) forecasting the product sales price, and (7) estimating the return on investment. For more details on the overall process, the investor can also refer to a guide for gas to liquids plants published by the U.S. Energy Information Administration (EIA).[26] Typically, the cost estimates to guide decision making at the preliminary TEA stage are accurate to within -30 to +50%, not including contingency, and a typical contingency range is 15 to 40%.[25] The estimates are carried out using engineering judgment, Lang factors, and capacity factoring (e.g., the six-tenths rule for capacity adjustment).[25]

Smaller distributed plants could be attractive to smaller investors, because they present a lower overall financial risk. Since syngas production represents a major part of the overall cost of any gasification-based industrial plant, cost reductions to syngas production using advanced modular technology for small plants have the potential to significantly reduce the cost of the plant. For example, clean syngas production for a large CTL plant typically requires 65% of the overall plant investment. Thus, if syngas production costs can be reduced for small plants, the risk to investors will be further reduced. For this reason, the U.S. Department of Energy (DOE)/NETL and other research groups are pursuing R&D on the development of small distributed-scale, modular plants. Through such R&D, if the smaller, modular gasifier and other modular components could be standardized, they could be shop fabricated at a significantly reduced cost, thereby avoiding the penalty typically associated with scaling down. Furthermore, larger plants of the future will require CO_2 capture and either storage or sale, although tax credits may offset some of this cost.

Another promising possibility of capital and operating cost reduction is process intensification (PI). Although there have been several definitions of PI over the past 30+ years,[27] the one that is most relevant in this context is *"Technologies that replace large, expensive, energy-intensive equipment or process with ones that are smaller, less costly, more efficient, or that combine multiple operations into fewer devices (or a single apparatus)"* by Touris and Porcelli.[28] Some examples relevant to gasification technologies include: (1) using a single catalytic reactor downstream of the gasifier for syngas tar reforming, methane reforming, ammonia decomposition, and adjustment of the H_2/CO ratio to 2.0 for liquids production; (2) combining the Fischer-Tropsch process and wax upgrading into a single liquid production reactor to primarily produce desired diesel and jet fuel hydrocarbons from syngas; (3) char micronization to enhance the surface-area-to-volume ratio, which enhances heat transfer rates and reaction site availability, reducing particle residence time and reducing required gasifier size; and (4) indirect gasification with steam (discussed in Chapter 2), which can eliminate the need for oxygen from an ASU.

3.4 GASIFIER SELECTION

Once the project developers have carried out a preliminary TEA and decided whether to proceed, the next steps are to select the fuel and the gasifier technology and develop a preliminary process design package (factors for determining fuel selection are provided in Section 3.3.2). Gasifiers have been in use since the 1800s, first for town gas, then for electricity and SNG, and, more recently, for liquids since the 1920s.[29] Although many developments in gasifiers have occurred over the years, all gasifiers to date can be classified according to the method of contacting solid fuels with the gasification reagents (generally steam and air [or oxygen]): moving-bed (updraft or downdraft), fluidized-bed (bubbling or circulating) and entrained flow. Conceptual diagrams depicting these methods of contacting are shown in Figure 3-1.[29] Several gasifier technologies were described in Chapter 2 and some relevant descriptions were also provided in Section 3.2. Commercial technologies currently in use and/or significantly proposed in future plants will be discussed below to assist project developers in gasifier selection and design.



Used with permission from Kopiersperre[29]

Figure 3-1. Examples of solid fuel and air (oxygen) contacting patterns in gasifiers

3.4.1 CURRENT COMMERCIAL GASIFIER TECHNOLOGIES

Several gasification technologies are employed in operational gasification plants and many advanced technologies are proposed for future plants. Three databases of operational and potential future gasification plants are available.[30, 31] Two of these—the world (non-U.S.) database and the U.S. database—were compiled and updated in 2016,[30] whereas the third—the Chinese database—was compiled in 2014.[31] The first two databases only list proposed gasification plants and not ones that are operational. The Chinese database lists both existing and planned. Since all these databases are at least four years old, some of the data could be obsolete, as many of these projects may not have moved forward due to a lack of funding or other reason. The investor should exercise caution and carry out their own due diligence if they are interested in following up on any of the projects mentioned in these databases. Nonetheless, the databases are a valuable resource and show the breadth of the applications of gasification for producing power, liquid, and SNG, and list the variety of gasifiers employed.

3.4.2 BIOMASS GASIFICATION

Biomass gasification plants for power, such as Rankine cycles and IGCC, are mature technologies. However, biomass gasification plants designed to produce electricity using an IC engine and gasification plants for biomass-to-liquids (BTL) applications are still in their infancy, with generally only mini-scale systems demonstrated. Before moving on to currently operational and proposed gasifiers, biomass gasification is worth discussing, with an emphasis on producing electricity using an engine or turbine and BTL. Further details of biomass gasification are available in three overviews.[32, 33, 34] These overviews provide interested investors with detailed information on various past and future projects, gasifiers used, new developments, composition of syngas produced, and problems associated with biomass gasification. As biomass waste is CO_2 neutral, there is significant worldwide interest in developing biomass gasification-based systems. However, most of these systems are small (<500 kilowatt-thermal [kWth]) and mainly produce heat by combusting the tar-laden syngas in a burner coupled to the gasifier.

Technical barriers to the development of larger biomass gasification-based plants are the collection/availability of large quantities of appropriate biomass within a 50-mile radius; biomass variability, seasonality, feeding problems due to low mass density and grindability; low energy density due to the high oxygen content, high moisture content, high cost associated with pretreatment (e.g., drying, forming, sizing, torrefaction); and high levels of tar formation (1 to 10 wt%) in updraft moving-bed and fluidized-bed gasifiers. Tar can cause equipment blockages, increased maintenance, and operational problems. Heavier tar can condense and plug downstream equipment. Lighter tar can form aerosols and render the syngas unusable for producing electricity using an engine or turbine or a catalytic reactor in BTL applications.

Tar mitigation methods external to the gasifier include catalytic cracking[34] and thermal plasma.[35] These methods are unattractive, as they are both costly and prone to catalyst poisoning and have scaleup issues, respectively. A well-designed downdraft moving-bed gasifier (Figure 3-1) can reduce the tars produced to less than approximately 500 to 1,000 parts per million by weight (ppmw), because they have to travel down through the combustion zone where they would burn, and also through the char bed that may act as a tar destruction catalyst. Unfortunately, scaleup of downdraft gasifiers is difficult, because of the heat rise in the opposite direction to the gas flow. Thus, the largest commercial designs that utilize these gasifiers are generally smaller than 500 kWth. Another design that has undergone significant research for tar mitigation consists of a conveying pyrolyzer close coupled to the downdraft gasifier.[36]

Small commercial downdraft gasification systems are available from various vendors including Syntech Bioenergy and Ankur Scientific. The lower tar content from these gasifiers (even though at a level of 500 to 1,000 ppm) creates problems and requires removal downstream using filters before the engine. Long-term, reliable operation of a downdraft gasifier coupled to a larger than 150-kWe, IC engine generator set has yet to be demonstrated, mainly due to problems created by tar. Advanced modifications to the downdraft design to maximize the temperatures in the gasification zone could result in further reduction of the tar through reforming/cracking of the tar on the char surface.

On the other hand, updraft moving-bed and bubbling fluidized-bed (BFB) gasifiers have been scaled up to larger sizes, but generally produce significant quantities of tar (more than 0.2%, as much as 5%) and ammonia (more than 3,000 ppm). Following a water quench that leads to wastewater containing heavy tar, oxygenates, ammonia, and other contaminants, the remaining tar needs to be dealt with downstream using a solvent scrubbing scheme, such as the Olga Process.[37] Multiple cleaning steps significantly add to the capital cost. Entrained flow gasifiers operate at temperatures on the order of more than 1,300°C (2,372°F) and can thermally destroy the tars within the gasifier. Thus, these gasifiers are the ones that are the most desirable for commercial application for biomass gasification, but they have yet to be developed at a commercial scale.

Slurry-fed entrained flow gasifiers cannot be used for low energy density fuels, such as biomass, because of the significant amount of water that needs to be added to prepare the slurry. Consequently, dry-fed entrained flow gasifiers are the preferred choice for biomass gasification. The biomass feed needs to be sufficiently dried for feeding to these gasifiers. No commercial entrained gasifiers are currently available for biomass; however, there has been recent research in laboratory- to pilot-scale, non-slurry entrained flow gasifiers for biomass.[38, 39, 40, 41, 42, 43, 44] The ash in the biomass is preferably removed as a slag in these gasifiers. One study involved pilot-scale tests in a downdraft 1-megawatt-thermal (MWth) entrained flow biomass gasifier and found that the oxygen stoichiometric ratio was the most important parameter for influencing cold gas efficiency and soot formation.[39] The study concluded that slag removal needs to be improved, possibly by including additives like calcium oxide with the biomass.

An advantageous and potential scenario for biomass to make headway toward medium- to large-scale would be to gasify a coal-biomass mixture in entrained flow gasifiers. This would be a stop-gap measure while gasification and biomass pretreatment technology matures. In a carbon-constrained world, this combination may provide the synergy needed to achieve this economy of scale. Biomass, being CO_2 neutral (excluding CO_2 associated transportation and harvesting, if wastes are not used), can help to reduce the carbon footprint. By also pursuing CO_2 capture, the plant can potentially be carbon negative.[19, 45] Biomass feedstocks are also compatible with steam-blown gasifiers, such as the MILENA gasifier from the Energy Research Centre of the Netherlands (ECN).

3.4.3 COMMERCIAL GASIFICATION TECHNOLOGIES

Gasification is a mature technology with many gasification plants in operation and proposed in future projects around the world, but mostly in China. A description of many of these gasifiers is available from various resources.[7, 46, 47] These pressurized gasifiers, their contacting method, and where the fuel and oxidant are fed are listed in Table 3-4. In addition, these references provide operating parameters, such as feedstock size/type, pressure, and steam and oxidant requirements for various gasifier types, along with the syngas composition generated. The pros and cons of dry- vs slurry-fed entrained flow gasifiers, the use of air versus oxygen as oxidant for IGCC, quench versus heat recovery for entrained flow gasifiers, and the syngas cleanup requirements are also provided. The NETL Gasifipedia website provides descriptions of a number of these gasifiers.[9] To narrow the choice of gasifier selection for a plant, investors are encouraged to thoroughly review these gasifier descriptions.

Table 3-4. Gasifiers and their type[7, 46, 47]

GASIFIER		GASIFIER	
NAME	TYPE	NAME	TYPE
Lurgi Mark IV	Updraft, dry bottom moving-bed, fuel fed from top	British Gas Lurgi (BGL)	Updraft, moving-bed, slagging version of Lurgi
High Temperature Winkler	Bubbling fluidized-bed	U-Gas (Synthesis Energy Systems)	Bubbling fluidized-bed, non-slagging
Foster Wheeler	Circulating fluidized-bed (air or O_2 blown)	KBR (TRIG)	Circulating fluidized-bed (transport), non-slagging
GE-Radiant Cooling	Slurry, entrained-flow, top fed, slagging	GE- full quench	Slurry, entrained flow, top fed, slagging
E-Gas	Two stage, slurry feed, slagging	SCGP (Shell)	Dry, entrained flow, middle fed, water wall, slagging
Siemens (Noell)	Dry, top fed, entrained flow, slagging	MHI	Two-stage, dry, entrained flow, slagging
Multi-Purpose Gasifier (Lurgi)	Entrained flow, top fed, multi-fuel, slagging	Prenflo (Uhde)	Dry, entrained flow, membrane wall, slagging
MILENA Gasifier (ECN)	Indirect, steam-blown, bottom fed, non-slagging		

Also, as mentioned in Section 3.4.1, there are three databases of operational and potential future gasification plants available.[30, 31] These databases show numerous ongoing and proposed gasification-based plants to make power, SNG, liquid fuels, and chemicals in China. Although the Lurgi and GE gasifiers have been the most common, newer gasifiers have been used or proposed. These newer commercial gasifiers (most from China, with one from Japan) are listed in Table 3-5.

Table 3-5. Recent commercial gasifier developments in China and Japan[9, 48]

GASIFIER NAME	BRIEF DESCRIPTION	REFERENCE OR DEVELOPER
ECUST	Variant of GE gasifier; O_2 and coal slurry enter gasifier near top at four locations, four burners, cooling to raise steam, downdraft entrained flow, slagging	9, 47
TPRI	Entrained flow, two-stage, updraft, dry, water cooled, membrane wall, slagging	9, 47
HT-L	Entrained flow pulverized pressurized coal gasifier; details proprietary	Changzheng Eng. Co., China Aerospace Sci. & Tech Corp.
Tsinghua	Variant of GE-quench gasifier, primary and secondary O_2 feed from top to shift hot zone down, membrane wall, slagging	9, 47
ICC-CAS	Agglomerating bubbling fluidized-bed gasifier, limited to small units, limited development	47
MCSG	Slurry gasification of coal with oil, petroleum, asphalt; entrained flow; variant of GE-quench gasifier, proprietary nozzle technology, slagging	9, Northwest Research Institute of Chemical Industry
SEDIN	Variant of Lurgi moving-bed updraft gasifier	47
EAGLE	Two-stage updraft dry entrained flow gasifier, variant of MHI and TPRI gasifiers, tangential injection causes swirling flow and increases residence time, slagging	9, Electric Power Dev. Corp and NEDO, Japan

These lists and descriptions are intended to guide investors to make an informed decision about selecting a gasifier for the plant under consideration. Most of the proposed large-scale plants are either based on Lurgi updraft moving-bed technology or entrained flow gasifiers. Less than a handful of projects based on BFB, circulating fluidized-bed (CFB), and transport gasifiers have been selected. The trend for large-scale gasifier plants of the future appears to be clearly slated toward entrained flow gasifiers.

In contrast to China, only a few coal gasification-based plants have been constructed and operated in the United States over the past 40 years or so, as listed in Table 3-6. The first three plants in this table are coal- or petcoke-to-chemicals or SNG projects, whereas the next three are IGCC projects. The three coal-to-chemicals plants have been generally successful, as seen from the fact that they have been in operation for 20 to 37 years. In particular, the Eastman project is a showpiece for the world as a successful coal-to-chemicals complex funded entirely by private investment.

Table 3-6. U.S. coal gasification plants for power, SNG, and chemicals[9, 49]

NAME AND YEAR OPERATING	PRODUCT/TYPE	SIZE	GASIFIER AND FUEL	PROCESS STEPS OR COMMENTS
Eastman Chemical Company Kingsport Plant 1983 to present	Acetyl chemicals/ Coal to chemicals	0.5 to 1 billion lb/year	GE-quench; Kentucky and Virginia bituminous	Coal gasified to syngas, Syngas partial shift/ cleanup, CO separation, methanol synthesis and carbonylation to acetic acid
Dakota Gasification Co. Great Plains Synfuels Plant 1984 to present	SNG, ammonia, phenol,	153 MMscfd	Lurgi Mark IV; North Dakota Lignite	Lignite gasified to syngas, tar and oil removal, sour-shift, Rectisol®, methanation, CO_2 to EOR
Coffeyville Resources Fertilizer Plant, Kansas 2000 to present	Fertilizers; petcoke to chemicals	1,250 tpd NH_3; 3,000 tpd UAN	GE-quench; petcoke from next door refinery	Petcoke gasified to syngas, sour-shift, two-stage Selexol™, H_2 separation by pressure swing, NH_3 synthesis, CO_2 and NH_3 to UAN synthesis, CO_2 to EOR
Wabash River Energy, Indiana 1995 to 2016	IGCC	250 MWe	EGas (CBI), Bituminous coal, petcoke	Details available in References 9,48, 49
Tampa Electric IGCC Project, Polk County, Florida 1996 to present	IGCC	260 MWe	GE Radiant Cool, Bituminous coal	Details available in References 9, 50
Duke Energy, Edwardsport, Indiana 2013 to present	IGCC	618 MWe	GE-Radiant Cool, Bituminous coal	Details available in References 51, 52
Southern Company, Kemper County, Mississippi (never completed)	IGCC	582 MWe	Transport	Lignite gasified to syngas, full syngas cleanup w/ CO_2 capture, converted to NGCC in 2017

Although the IGCC projects in Table 3-6 have demonstrated their respective gasifier technologies, they have not been highly successful due to cost overruns and/or outages leading to low availability and the inability to demonstrate the required long-term operation. It appears that there is a lot of initial learning involved in operating these plants due to their highly integrated nature. Any issues upstream can have a domino effect on the whole plant, leading to catastrophic failures. There must be an understanding and a long-term plan to absorb the higher cost of the first-of-a-kind plant knowing that the cost of subsequent plants will be lower.

Construction cost overruns (e.g., Duke Energy, Kemper County IGCC) speak to the risk that investors face with large plants and encourages the development of small, modular plants. Additional cost saving can possibly be realized by eliminating the ASU for smaller, modular plants for IGCC by using an air-blown, dry-fed entrained flow gasifier or indirect, steam-blown gasifier.

3.5 PRELIMINARY PROCESS DESIGN, MASS BALANCE, AND PROCESS CONTROL

If the project developers and investors have determined—guided by the value proposition assessment (see Section 3.3.4)—an approximate size (based on investment) and are ready to move forward, their next task is narrowing the choices of fuel, land (i.e., location), and commercial gasifier. It is prudent at this stage to investigate two or three choices of each before making a final choice of the combination. Water availability/requirements for the plant, wastewater treatment/disposal, and slag/ash handling are additional prominent factors that guide this decision. These factors, along with the information in previous sections and discussions and performance data from the gasifier vendors, should be used as a guide for the final choice. Once the final choice of the gasifier is made, a preliminary process design including a process flow diagram (PFD) and mass balance around the plant should be developed for presentation to an engineering design firm. The following discussion is limited to the gasifier.

3.5.1 HEAT AND MASS BALANCE AROUND GASIFIER

Knowing the proximate/ultimate analysis of the selected feedstock and the syngas composition available from the gasifier vendor, the mass flow rate of syngas produced per unit mass flow rate of feedstock fed can be determined through a carbon balance and carbon conversion. Approximate syngas compositions from various gasifiers are available. Typical carbon conversion rates for various gasifiers are shown in Table 3-7. Gasifier vendors will also generally provide the syngas composition and carbon conversion rate under a nondisclosure agreement (NDA).

Table 3-7. Typical carbon conversion rates for various gasifiers

DOWNDRAFT MOVING-BED	UPDRAFT MOVING-BED	FLUIDIZED BED	ENTRAINED FLOW
90 to 95	>98	90	>99

In addition to carbon balance, hydrogen and oxygen balances around the gasifier are also useful to validate the carbon-based mass balance and help to estimate the steam concentration in the wet syngas. It is cumbersome to measure the steam concentration in the wet syngas using standard methods like a gas chromatograph (GC) while the plant is online, so generally the syngas constituent concentrations are only measured after condensing the steam. In a quench gasifier, like GE (in quench mode), it is not practical to measure the steam concentration before the quenching stage.

Vendor specifications of the steam-to-carbon and oxygen-to-carbon ratios can be used to estimate the outlet steam concentration by attempting to close the hydrogen and oxygen mass balances. Once the steam and oxygen input to the gasifier is known, the total oxygen and hydrogen inputs to the gasifier can be calculated by adding the contributions from the feedstock moisture and the hydrogen and oxygen contents. Accurate proximate and ultimate analyses are important, since oxygen in the ultimate analysis is calculated by difference. Estimating the carbon, hydrogen, and oxygen balances provides a more accurate picture of the overall mass balance around the gasifier. Measurement of sulfur containing gases in the syngas then allows for the estimation of sulfur captured within the solid waste matrix (e.g., slag or bottom ash, char, fly ash).

3.5.2 OPERATION AND PROCESS CONTROL FOR GASIFIER

Detailed startup, operating, and shutdown procedures, and safety protocols will be provided by the selected gasifier vendor. Even so, conducting a process hazard analysis (PHA) is critical for the gasifier and should be carried out for a detailed piping and instrumentation diagram (P&ID) around the gasifier provided by the A&E firm with input from the gasifier vendor. Not only does the PHA evaluate each phase of the operation (i.e., startup, normal operation, and shutdown) to ensure safe operation, it also considers the effects on process and safety aspects of failure of any of the components (e.g., flow stoppage due to a valve stuck closed or over-pressurization due to a valve stuck open). To further emphasize the importance of the PHA, a recent incident of a stuck valve that failed to fully close caused an explosion at Eastman's Chemical Company Kingsport Plant (Tennessee, USA) during a gasifier switch.[50] The PHA is intended to assist with any recommended modifications to the P&ID and the operating procedure to the A&E firm and the gasifier vendor who will then finalize the P&ID and startup, operating, and shutdown procedures. The final P&ID should be thoroughly reviewed by a professional engineer (PE) to ensure that the recommendations from the PHA have been included. During construction, similar reviews should be conducted regularly to ensure that the plant is being built strictly according to the final P&ID.

A gasifier is normally started with an inert gas flush followed by heat up using a natural gas or propane burner, which is then followed by injecting the feedstock, oxidant, and steam. Rotation from one gasifier to another is done by starting up the second gasifier with syngas directed toward a flare and bringing it online while shutting down the first gasifier. Trained and skilled operators are needed due to the challenges associated with maintaining the desired syngas composition during this rotation.

A few examples of actions for safe operation and maintaining the performance of existing gasifiers include (1) directing the raw syngas during startup, upsets, and shutdown to a flare; (2) continuously analyzing the oxygen in the syngas to ensure that the oxygen level is close to zero for proper operation and to avoid an explosive mixture (preheating the gasifier can help prevent this scenario); (3) monitoring temperatures in several selected locations within the gasifier; (4) using the difference between two selected temperature measurement locations in a feedback control loop[51] to adjust the gasifier inputs of feedstock and oxidant; (5) measuring the syngas composition using continuous analyzers and GC to adjust the feedstock and oxidant flow; and (6) periodically measuring the composition of the feedstock to ensure that the heating value of the fuel is within the required window and that the sulfur and mineral matter levels have not exceeded the limit.

3.6 ENVIRONMENTAL CONSIDERATIONS

Important environmental considerations in a gasification plant include water requirements, wastewater treatment and recycle/disposal, solid waste (i.e., slag, bottom ash, char, and fly ash) disposal/recycle/sale, criteria pollutants air emissions, and blowdown effluent control. The level of discharges allowed is governed by federal, state, and local regulations. A permit is required to operate a plant; a comprehensive environmental impact statement needs to be prepared and submitted to the proper authorities to obtain the permit. Detailed information on these topics is available in the literature.[9, 51, 52, 53, 54, 55]

3.6.1 WATER REQUIREMENTS

Pulverized coal (PC) power plants are a huge net consumer of water. Due to the use of a gas turbine in an IGCC plant to produce a major portion of the power, reliance on steam is reduced, resulting in significantly lower water consumption compared to a PC plant. All plants of the future will require CO_2 capture, leading to increased water usage. However, since CO_2 capture integrates better with IGCC, the increase is expected to be less compared to PC plants. In contrast to IGCC, F-T plants that are water-cooled have more than twice as much water consumption (more than 5 gallons/gallon product) compared to an oil refinery (2.5 gallons/gallon product). On the other hand, zero-discharge, air-cooled F-T plants have the potential to reduce the water consumption to 1 to 1.5 gallon/gallon product.

3.6.2 WASTEWATER TREATMENT

Wastewater from gasification processes includes collected rainwater, all wastewater streams generated or captured during normal operations, and equipment purges/wash-downs during maintenance activities. It contains small amounts of dissolved solids, organic compounds, and gases and needs to be treated to remove the contaminants before being recycled back to the process or being discharged to a body of water. Depending on the feedstock used, the wastewater streams may also contain trace toxic compounds such as arsenic that will affect how the streams are handled and treated. Since the reuse of the water within the gasification plant minimizes water consumption and water discharge, it is desirable to recycle most of the treated process water so that only a relatively small amount is discharged. Wastewater treatment technologies, including those based on mechanical, chemical, and/or biological methods, are all commercially available and used frequently by other heavy industries. The complexity of the technology, however, depends on the gasifier. Tar and oil producing gasifiers, such as Lurgi, require more complex schemes compared to entrained flow gasifiers.

3.6.3 SOLID WASTE

The amount of solid waste and potential solid products, such as slag, bottom ash, char, and fly ash, depends on the amount and composition of the mineral matter content of the feedstock being gasified. Compared to conventional combustion, gasification produces much less fly ash. Gasifiers operated at temperatures below the ash fusion temperature, such as Lurgi, produce mostly bottom ash and char whereas those operated at temperatures higher than the melting point of ash, such as entrained flow gasifiers, produce a slag that flows to the bottom of the gasifier before being cooled to solidify.

Slag is an inert, glass-like material composed mostly of non-volatile metals and silicon. The mineral compounds are bound together in molten form until the slag is cooled in a water bath at the bottom of the gasifier. Volatile metals, such as arsenic, selenium, and mercury, are not captured in the slag, leave the gasifier as gaseous species, and are typically captured in the gas cleanup train. Slag can encapsulate toxic heavy metals to form a non-leachable material (which is highly desirable from an environmental standpoint) that can potentially be marketed. Char is composed of unreacted carbon and ash. It can be recycled back into the gasifier to increase carbon usage. Alternatively, it can potentially be used to make activated carbon, which can be sold as an adsorbent.

3.7 SUMMARY

The entire gasification plant should be subjected to a preliminary TEA before embarking on a detailed design of the gasifier module. The factors to be considered include investment, which dictates plant size; plant location; land availability; salability of the end products; environmental and permitting constraints; solid fuel availability; quality and cost; air- versus oxygen-blown designs to potentially eliminate the ASU; and syngas quality and cleanup requirements for converting it to the desired end product. A value proposition should then be determined using simple factoring methods. Smaller, modular plants that can potentially overcome the economy of scale issue should be considered to reduce risk. Once a decision has been made to move ahead, the next most important task is gasifier selection, which is strongly dictated by feedstock properties, plant size, and the desired end products. Once the gasifier choices are narrowed, vendors should be contacted to provide their fuel requirements and performance specifications. Based on this information, the final gasifier selection should be made, and startup, normal operation, shutdown procedures and process control strategies provided by the vendor should be evaluated. The disposal/treatment of solid waste products and treatment/recyclability of the wastewater should also be evaluated. A PFD and mass balance should be prepared around the gasifier and submitted to the vendor and an engineering firm for review and preparation of a detailed P&ID. A PHA should be conducted to determine any additional process control and safety requirements and finalize the P&ID before beginning procurement and construction.

Endnotes

1 J.L. Johnson (1979) Kinetics of Coal Gasification, New York, NY: John Wiley and Sons, Inc., ISBN 0-471-05575-1.

2 T.B. Reed (1981) Biomass Gasification: Principles and Technology (ed. T. Reed), Park Ridge, NJ: Noyes Data Corporation, ISBN 0-8185-0852-2.

3 R.F. Probstein and R.E. Hicks (1982) Synthetic Fuels, New York, NY: McGraw-Hill, ISBN 0-07-050908-5.

4 C. Higman and M. van der Burgt (2003) Gasification, New York, NY: Elsevier Science, ISBN 0-7506-7707-4.

5 H.A.M. Knoef, ed. (2005) Handbook Biomass Gasification, Enschede, Netherlands: BTG Biomass Technology Group BV, ISBN 90-810068-1-9.

6 J. Rezaiyan and N.P. Cheremisinoff (2005) Gasification Technologies: A Primer for Engineers and Scientists. New York, NY: Taylor and Francis.

7 D.A. Bell, B.F. Towler, and M. Fan (2009) "Gasifiers", in Coal Gasification and its Applications, Chapter 4, pages 35-69.

8 D.A. Bell, B.F. Towler, and M. Fan (2009) "Gasification Fundamentals", in Coal Gasification and its Applications, Chapter 3, pages 73-98.

9 NETL (2019) "Gasifipedia", *https://netl.doe.gov/research/Coal/energy-systems/gasification/gasifipedia*, Accessed March 29, 2020.

10 ChemViews (July 2, 2013) "Fuel Cell Capacity and Cost Trends", Wiley-VCH Verlag GmbH & Co. KGaA, Weinheim, DOI: 10.1002/chemv.201300069, *https://www.chemistryviews.org/details/ezine/4817371/Fuel_Cell_Capacity_and_Cost_Trends.html*.

11 NETL (2019) "Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity", NETL-PVB-22638, Final Report, DOE-NETL Contract No. DE-FE0025912.

12 A. Porcu, S. Sollai, D. Morotto, M. Mauro, F. Ferrara, and A. Pettinau (2019) "Techno-Economic Analysis of a Small-Scale Biomass to Energy BFB Gasification-Based System", Energies 12:494.

13 M. Ozonoh, T.C. Aniokate, B.O. Oboirien, and M.A. Daramola (2018) "Techno-Economic Analysis of Electricity and Heat Production by Cogeneration of Coal, Biomass, and Waste Tyre in South Africa", J. Clean. Prod. 201:192-206.

14 H.C. Mantriprogada and E.S. Rubin (2011) "Technoeconomic Evaluation of Coal to Liquids (CTL) Plants with Carbon Capture and Sequestration", Energy Policy 2803-2816.

15 B. Yu, and I. Chien (2015) "Design and Economic Evaluation of a Coal-to-Synthetic Natural Gas Process", I & EC Research 54:2339-2352.

16 E. Martelli, T. Kreutz, and S. Consonni (2009) "Comparison of Coal IGCC with and without CO_2 Capture and Storage: Shell Gasification with Standard vs Partial Water Quench", Energy Procedia 1:607-614.

17 Y. Jiang and D. Bhattacharyya (2016) "Techno-Economic Analysis of a Novel Indirect Coal–Biomass to Liquids Plant Integrated with a Combined Cycle Plant and CO_2 Capture and Storage", I & EC Research 55:1677-1689.

18 Y. Man, S. Yang, D. Xiang, X Li, and Y. Qian (2014) "Environmental Impact and Techno-Economic Analysis of the Coal Gasification Process with/without CO_2 capture", J. Clean. Prod. 71:59-66.

19 X. Lu, L. Cao, D. Wang, W. Xing, S. Cai, B. Shen, Q. Yang, C.P. Nielson, and M.B. McElroy (2019) "Gasification of Coal and Biomass as a Net Carbon Negative Power Source for Environment-friendly Electricity Generation in China", PNAS 116(17):8206-8213.

20 Q. Chen, M. Lv, Y. Gu, X. Yang, Z. Tang, Y. Sun, and M. Jiang (2018) "Hybrid Energy System for a Coal-Based Chemical Industry", Joule 2:607-620.

21 M. Sciazko and T. Chmielniak (2012) "Cost Estimates of Coal Gasification for Chemicals and Motor Fuels", in Gasification for Practical Applications, Chapter 10, 247-276.

22 IEA, Energy Technology Systems Analysis Programme (May 2010) "Syngas Production from Coal", Technology Brief S01, *www.etsap.org*.

23 A. Cormos and C. Cormos (2019) "Techno-Economic Assessment of Combined Hydrogen and Power Generation with Carbon Capture: The Case for Coal Gasification", Appl. Thermal Eng. 147:29-39.

24 F. Trippe and E. Henrich (2011) "Techno-Economic Assessment of Gasification as a Process Step within Biomass to Liquid (BTL) Fuel and Chemicals Production", Fuel Processing Technology 92:2169-2184.

25 R.M. Sari and K. Kolmetz, eds. (2014) "General Process Plant Cost Estimating (Engineering Design Guideline), Rev 01", KLM Technology Group, *www.klmtechgroup.com*.

26 EIA (December 2015) "Engineering Economic Analysis Guide: Liquid Fuel Technologies", *https://www.eia.gov/outlooks/documentation/workshops/pdf/eeaguide_12082015.pdf*.

27 F.J. Keil (2018) "Process Intensification", Reviews in Chemical Engineering 34(2):135-200.

28 C. Touris and J.V. Porcelli (2003) "Process intensification – has its time finally come?", Chem Eng Progr 99:50–55.

29 Wikipedia (December 3, 2020) "Gasification", *https://en.wikipedia.org/wiki/Gasification*.

30 NETL (n.d.) Gasification Plant Databases. *https://www.netl.doe.gov/research/coal/energy-systems/gasification/gasification-plant-databases*.

31 NETL (n.d.) China Gasification Database. *https://www.netl.doe.gov/research/coal/energy-systems/gasification/gasification-plant-databases/china-gasification-database*.

32 J.P. Ciferno and J.L. Malano (2002) "Benchmarking Biomass Gasification Technologies for Fuels, Chemicals and Hydrogen Production", Report to NETL. *https://www.netl.doe.gov/sites/default/files/netl-file/BMassGasFinal_0.pdf*.

33 C.J. Roos (2010) "Clean Heat and Power Using Biomass Gasification for Industrial and Agricultural Projects", WSU Extension Energy Program, WSUEEP08-033 Rev. 5 *https://www.researchgate.net/publication/228449066_Clean_Heat_and_Power_Using_Biomass_Gasification_for_Industrial_and_Agricultural_Projects*.

34 V.S. Sikarwar, M. Zhao, P. Clough, J. Yao, X. Zhong, M.Z. Memon, N. Shah, E.J. Anthony, and P.S. Fennell (2010) "An overview of Advances in Biomass Gasification", Energy Environ. Sci. 9.

35 F. Fabry, C. Rehmet, V. Rohani, and L. Fulchari (2013) "Waste gasification by Thermal Plasma: A Review", Waste and Biomass Valorization 4:421-439.

36 A.S. Mednikov (2018) "A Review of Technologies for Multistage Wood Biomass Gasification", Thermal Engineering 65:531-546.

37 R.W.R. Zwart, A. Van der Drift, A. Bos, H.J.M. Visser, M.K. Cieplik, and H.W.J. Konnemann (2009) "Oil-based gas washing - flexible tar removal for high-efficient production of clean heat and power as well as sustainable fuels and chemicals", Energy Research Center of the Netherlands, ECN-V-09-025.

38 K. Qin, P.A. Jensen, W. Lin, and A.D. Jensen (2012) "Biomass gasification behavior in an entrained flow reactor: gas product distribution and soot formation", Energy Fuels 26(9):5992-6002.

39 F. Weiland, H. Hedmana, H. Wiinikkaa, M. Marklund (2016) "Pressurized Entrained Flow Gasification of Pulverized Biomass – Experiences from Pilot Scale Operation", Chemical Engineering Transactions 50:325-330.

40 K. Qin, W. Lin, P.A. Jensen, and A.D. Jensen (2012) "High-temperature entrained flow gasification of biomass", Fuel 93:589-600.

41 J. Zhao, Q. Chen, H. Zhao, X. Cao, Q. Mei, Z. Luo, and K. Cen (2009) "Biomass-oxygen gasification in a high- temperature entrained-flow gasifier", Biotechnol. Adv. 27(5):606-611.

42 J.J. Hernandez, G. Aranda, J. Barba, and J.M. Mendoza (2012) "Effect of steam content in the air-steam flow on biomass entrained flow gasification", Fuel Process. Technol. 99:43-55.

43 X. Ku, J. Wang, H. Jin, and J. Lin (2019) "Effects of operating conditions and reactor structure on biomass entrained-flow gasification", Renewable Energy 139:781-795.

44 P.H. Jafari, A. Wingren, G. Hellstrom, and B.R. Gebart (2020) "Effect of Process Parameters on the Performance of an Air-Blown Entrained Flow Cyclone Gasifier", Int. J. Sustain. Energy 39(1):21-40.

45 T. Wang and H.W. Long (2011) "Performance and Economic Analysis of Biomass/Coal Co-Gasification IGCC Systems with Supercritical Steam Bottom Cycle, Part 1—Post–Combustion Carbon Capture", Proc. ASME 2012 Int. Mech. eng. Congr. & Expo., Houston, TX, November 9-15.

46 R.W. Breault (2010) "Gasification Processes Old and New: A Basic Review of major Technologies", Energies 3:216-240.

47 J. Phillips (2006) "Different Types of Gasifiers and their Integration with Gas Turbines", The Gas Turbine Handbook.

48 A. Minchener (2013) "Challenges and Opportunities for Coal Gasification in Developing Countries", IEA Clean Coal Center, CCC/225 ISBN 978-92-9029-545-7.

49 J. Phillips (March 2007) "Integrated Gasification Combined Cycle (IGCC) Design Considerations for High Availability, Vol 1: Lessons from Existing Operations", EPRI Report 101226.

50 H. Hayes (2018) "State discloses likely cause of Eastman explosion", Times News *https://www.timesnews.net/Business/2018/04/12/State-discloses-likely-cause-of-Eastman-explosion.html*.

51 W.P. Chan, A. Veksha, J. Lei, W.D. Oh, X. Dou, A. Giannis, G. Lisak, and T.T. Lim (2019) "A novel real-time monitoring and control system for waste-to-energy gasification process employing differential temperature profiling of a downdraft gasifier", J. Environ Manage. 15(234):65-74.

52 J. Colebroo, P. Shamblin, and P. Greywall (February 2010) "Integrated Gasification Combined Cycle (IGCC) Air Permitting: Overcoming Obstacles, Maximizing Flexibility", EUEC 2010, Phoenix, AZ.

53 J. Ratafia-Brown, L. Manfredo, J. Hoffmann, and M. Ramezan (2002) "Major Environmental Aspects of Gasification- Based Power Generation Technologies", Final Report. Science Applications International Corporation (SAIC) prepared for NETL. *https://www.netl.doe.gov/sites/default/files/netl-file/final-env.pdf*.

54 D. Zhao, W. Lun, and J. Wei (2017) "Discussion on Wastewater Treatment on Coal Chemical Industry", IOP Conf. Ser.: Earth Environ. Sci. 100:012067.

55 Q. Zhao and Y. Liu (2016) "State of the Art of Biological Processes for Coal Gasification Wastewater Treatment", Biotechnology Advances 34:1064-1072.

4.0 GUIDELINES FOR THE DESIGN OF A FEEDSTOCK HANDLING AND PRETREATMENT MODULE

4.1 INTRODUCTION

The required production of gasifier products and the conversion from feedstock to these products determines how much feedstock must be received. These quality parameters and product yields are key to understanding the required capacity of any feedstock handling system for its feedstock receiving and storage requirements. Most of the equipment and techniques discussed in this chapter are legacy technologies and infrastructure that have been historically used for coal, but, with potential future research and development (R&D), these items will be relevant for future gasification of clean feedstocks, such as biomass and municipal solid waste (MSW).

Site selection depends on many parameters, such as the proximity to the feedstock source and gasifier product markets. Project developers should consider the following questions:

- Will the fuel be delivered by river barge, railcars, trucks, or conveyor?
- Will gasifier products be shipped by rail, truck, or pipeline? Do the products need to be shipped at all (power production or other in-house use)?

These considerations require an economic analysis of the logistics and should be part of any site selection discussion. This chapter discusses the required capacity of a feedstock handling system, feedstock receiving options, sampling and analysis, feedstock storage and reclamation options, requirements for feedstock blending and feedstock flowability, pretreatment requirements, and health and safety concerns.

4.2 CAPACITY OF FEEDSTOCK HANDLING SYSTEMS

Feedstock properties vary in terms of inherent moisture, fixed carbon, total carbon, and heating value. Generally, for a given class of solid fuel, as moisture content increases, fixed carbon, volatile matters, and heating value decrease. Ash content dilutes fixed carbon, total carbon, and heating value analyses. The capacity of feedstock handling systems is a matter of matching feedstock source to a gasifier type to produce the desired products.

For example, if the project developer is evaluating a handling system for a feedstock with a heating value of 8,800 British thermal units (Btu)/pound (lb) versus a feedstock with a heating value of 11,000 Btu/lb. To ensure that the gasifier receives the same amount of fuel input energy, the handling facilities must move and store 25% more mass of the feedstock with the lower heat content.

4.3 RECEIVING

The type of receiving system is defined by the location of the gasification plant in relation to the source of the feedstock. Project developers need to consider the following questions:

- Is the plant to receive the feedstock directly by conveyor?
- Is the plant close enough to the feedstock sources so it can be delivered by truck?
- Is the plant located near a rail line or river?

These questions should be considered whether the site is a greenfield or brownfield site. If any feedstock receiving facilities are available at a brownfield site, they must be evaluated by competent engineers for structural integrity and to ensure the proper capacity.

In any case, the system will include scales to ensure that the correct amount of feedstock is delivered and a sampling system to ensure that the delivery of the feedstock meets specifications.

4.3.1 FEEDSTOCK RECEIPT

There are several methods for transporting feedstock to a gasification plant.

- **By Conveyor**—If the gasification plant is located directly at the source of the feedstock, it can receive the feedstock by conveyor belt. Conveyor systems are laid out by experienced engineering companies to account for the terrain, requirements for foundations and pillars, and the location of the feedstock storage facilities. Conveyor belt width and speed determine the capacity of the conveyor system. There may be transfer points along the conveyor route depending on the distance.
- **By Truck**—Road weight limits set the capacity for truck haulage. Trucks will usually dump their loads onto a stockpile area where dozers or loaders will move the feedstock into active storage and reclaim. Alternatively, a specially configured dump truck can put feedstock into a hopper and then onto a conveyor belt that will convey the feedstock into the storage and reclamation facilities.
- **By Rail**—Feedstock is transported in either bottom dump or rotary dump cars. The names essentially describe the required system for feedstock receipt. Bottom dump cars enter an unloading station and the hoppers at the bottoms of the cars would be opened. The feedstock would discharge into a larger hopper with a conveyor system at the bottom to convey the feedstock to the storage and reclamation facility. Rotary dump cars (Figure 4-1) enter a specially designed, more costly unloading station where they are turned on their side for feedstock discharge into a hopper. Again, a conveyor located at the bottom of the hopper transports the feedstock to the storage and reclamation facility.

Used with permission from Metso[1]

Figure 4-1. Rotary dump station for rail

- **By Barge**—If receiving feedstock by river transport, a barge unloading facility is required. Generally, there are two types of barge unloaders: clamshell (Figure 4-2) and bucket (Figure 4-3). A clamshell, or balance crane, unloader is swung over the barge to collect the feedstock. The clamshell is then swung over a feedstock receiving hopper with a conveyor system at the bottom. The conveyer transports the feedstock to the storage and reclaim facility. Bucket unloading systems (Figure 4-3) consist of a continuous bucket elevator that digs into the feedstock and feeds a conveyor system for transport to the storage and reclaim facility.

Used with permission from Metso[1]

Figure 4-2. A clamshell, or balance crane, unloader for barges

Used with permission from Metso[4]

Figure 4-3. Continuous bucket unloading system for barges

4.3.2 FEEDSTOCK WEIGHING

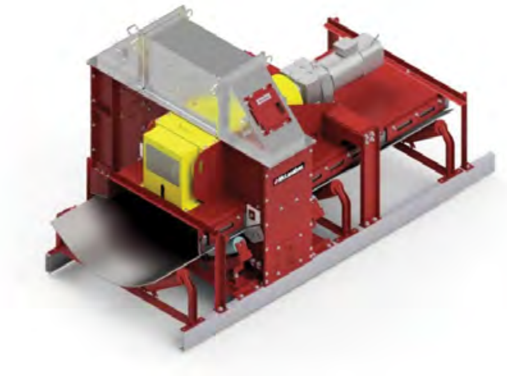
Feedstock is weighed at the origin as it is loaded into transportation vessels. It is also important to have a weighing system in the feedstock receiving area at the gasification plant. For conveyor deliveries, a belt scale is appropriate. For delivery by truck, a set of truck scales weigh full trucks when they enter the facility as well as empty trucks when they exit. Depending on the plant site, railcars may be loaded with weighed batches of feedstock. In the receiving station, railcars can be weighed again, or the discharge can be weighed in the hopper (fitted with load cells) or on a belt scale in the conveyor system. Barges are loaded at the feedstock origination site with batches of feedstock that are weighed based on the "draft" (or the depth in the water) of the barge. Following the unloading of the barge, the material may be weighed in the hopper (again, fitted with load cells) or using belt scales on the conveyor system. Scales should be distanced from the conveyor loading point to avoid disruptions at the loading point. Project developers should consult scale manufacturers for appropriate distances.

All scales and load cells need to be routinely calibrated for proper operation.

4.3.3 FEEDSTOCK SAMPLING AND ANALYSIS

Feedstock receipts need to be sampled and analyzed to confirm that the feedstock meets the specifications required for payment. Sampling and analysis typically follow methods established by the American Society for Testing and Materials (ASTM).* Feedstock samples for payment purposes should be collected using mechanical samplers. In some cases, grab samples are manually taken from the tops of transport vessels. Manual sampling is not recommended for feedstock receipts.

Mechanical Sampling Systems—Mechanical sampling systems are designed to collect a representative sample from a stream of feedstock. Sampler manufacturers rely on sampling statistics to properly design a sampling system while considering sampler location, the size of the sample cut required, the number of sample cuts required, cutter speeds, and requirements for sample size to conduct any required analysis. Swing-arm samplers often sit over a conveyor belt and collect a sample by swinging a collection arm through the feedstock on the conveyor belt. This sample is then typically ground to reduce the average particle size and then resampled to produce a sample for analysis. A swing-arm (or cross belt) sampler is shown below (Figure 4-4), followed by a swing-arm sampler as part of a complete sampling system (Figure 4-5).



Used with permission from McLanahan[2]

Figure 4-4. Swing-arm/cross belt sampler for conveyors

---

* These methods are available at *https://www.astm.org/Standards/coal-and-gas-standards.html.*

Used with permission from McLanahan[2]

Figure 4-5. A swing-arm/cross belt sampler as part of a sampling system

Sampling systems should be routinely tested for bias by comparing the analysis of the sample collected with the sampling system to a sample collected by a stopped-belt sampler. These tests should be conducted by a reputable laboratory.

If individual trucks, railcars, or barges are to be sampled, auger samplers (Figure 4-6) are available. The auger digs into the feedstock sample to recover material throughout the depth of the vessel. However, this is not a truly representative sample, because each particle in the truck does not have an equal chance of being collected.

Used with permission from McLanahan[2]

Figure 4-6. Auger sampler

Online Analyzers—While most contracts rely on a physical sample of feedstock to confirm the analysis for payment, online analyzers are often used in receiving stations, especially when feedstock is received from several sources. Per Woodward, analysis of a properly collected sample will accurately reveal a given fuel's quality, while online analysis provides real-time information to help control fuel quality.[3] Analyzers allow a quick analysis of a given fuel's properties and enable the receipts to either be blended in storage facilities or segregated for later blending. Online analyzers can measure moisture, ash, and sulfur content, or they can be more sophisticated and measure ash constituents. They would also prove useful in blending feedstocks of different quality from segregated storage areas to ensure that the analysis of the feed to the gasifier is relatively constant. Online analyzers must be routinely calibrated.

4.4 STORAGE AND RECLAIM

Power plants typically have an onsite solid fuel storage area holding three months of operation at full load. The area is not always filled, but the plant operators will strategically fill it to prepare for times when supply may be disrupted. Most notably, this could be during winter months when receipts may be delayed. Project developers should consider the capacity for a feedstock storage area for a gasifier. Fuel can be stored in piles with feeders located under the pile that feed conveyors to transport the feedstock to the gasifier for final preparation.

Some fuels can be stored in stacking tubes (Figure 4-7) located over underground feeders that feed a conveyor to the gasifier. If feedstocks of differing quality are received, they can be blended or segregated into different piles using multiple stacking tubes. With knowledge of the fuel quality in each stacking tube pile and an online analyzer situated over the gasifier feed conveyor, feedstock can be blended in different combinations to meet the gasifier specification by controlling the feed rate of each stacking tube reclaim feeder.

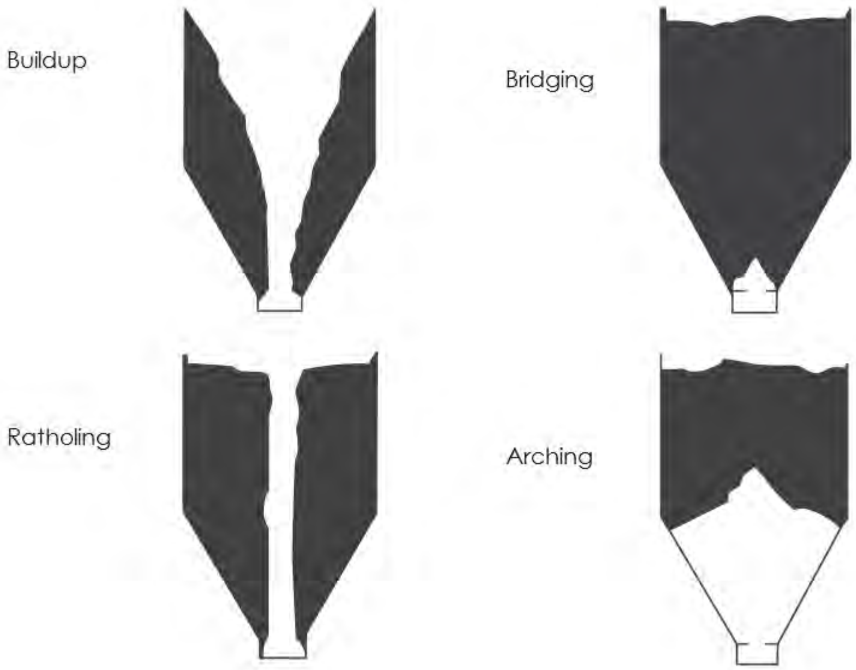
Used with permission from Marietta Silos[4]

Figure 4-7. Example of a stacking tube

Feedstock can also be stored in silos to protect it from the weather. Silos are equipped with hoppers at their bases. The discharge from these hoppers is controlled by feeders. Blending from silos is accomplished by using online analyzers to prepare a consistent feedstock quality.

4.5 FLOWABILITY

Feedstock properties like surface moisture and the amount of fines affect the ability of the fuel to flow readily in handling systems. This is an issue if it is stored in a bin or silo for an extended period under pressure, which compacts the feedstock at the bottom of the bin or silo. Bin bottom design (i.e., angle, materials of construction, size of opening) are important factors to ensure that feedstock does not arch/bridge, rathole, or buildup in a bin bottom (Figure 4-8).



Used with permission from EPRI[5]

Figure 4-8. Flow problems in bins

4.6 PRETREATMENT

The type of gasifier depends on the type of products (i.e., power, fuel, or chemicals) to be produced, which determines the general feedstock requirements and necessary pretreatment (see Chapter 2, Table 2-5). This pretreatment may include drying, grinding, and/or pulverization to achieve the gasifier feed moisture content and size requirements. There are also some gasifiers that require a slurry as the feed. In this case, the feedstock is first crushed to a finer top size and then fed with water into a rod, ball, or bead mill. The mix of feedstock and water is managed by the controlled addition of both the feedstock and water. The mill charge with rods, balls, or beads is controlled to produce the appropriate particle size distribution for the slurry—a bimodal distribution assists with forming a slurry that can remain dispersed in storage. In some cases, the slurry is up to 70% solids by weight. This high-solids content allows the slurry to remain relatively stable should it need to be stored prior to firing in the gasifier. Slurry storage tanks are usually fitted with agitators to help keep the slurry dispersed. Proprietary dispersants are added to ensure stability in some cases. Slurry viscosity is also an important consideration for the ease of pumping the slurry into the gasifier.

For biomass, an additional layer of pretreatment is necessary to ensure feeding consistency and allow for long-term storage. The most reliable, technologically mature method of pretreatment available for these purposes is torrefaction. With torrefaction, raw biomass is exposed to temperatures of about 500°F (260°C) for a period of 5-10 minutes. The process transforms the biomass into a reddish-brown, brittle, hydrophobic substance with physical and thermochemical properties that approach those of low-to-mid-grade coals (with lower sulfur content). Through this process, about 10% of the total chemical energy and 30% of the mass (mostly moisture) is lost, resulting in a net 28% increase in average heating value.[5,6]

In general, the feedstock will be fed via conveyor from the storage area to the pretreatment unit operations and then into the gasifier unit. If the pretreatment includes making a slurry, special positive displacement pumps are typically used to pump the slurry under pressure to the gasifier unit. Each system has unique requirements. Project developers should consider the following for each system:

- Dust generation in any drying or dry grinding/pulverization step.
    - Incorporate positive pressure enclosures to avoid dust emissions (as needed).
    - Ensure that housekeeping is performed, as dust can cause explosions.
    - Consider crusting agents for any long-term storage areas.
- Methods for oversize protection (e.g., cyclones or other classifiers) if the material that is too large affects gasifier performance or flow in the system.
- Instrumentation incorporation throughout the pretreatment system to measure water flow rates, tonnages, air flow, pressures, and other important variables; include these in any process control system that incorporates the entire handling system and the overall gasifier system.

4.7 HEALTH AND SAFETY REQUIREMENTS

The facility owners and operators should adhere to any local, state, or federal regulations for health and safety. A rigorous health and safety program should be implemented with training for hazard recognition, proper use of personal protective equipment, and task performance. The National Institute for Occupational Safety and Health (NIOSH) includes many health and safety guidelines on its website.[†] These guidelines include workplace safety and health topics related to hazards and exposures, chemicals, emergency preparedness and response, industries and occupations, diseases and injuries, and safety and prevention.

Some specific areas that should be addressed include spontaneous combustion by monitoring temperature and CO, and dust control (especially the control of fugitive dust and respirable dust) as these affect the requirements for explosion proofing any storage areas and also affect worker health. Good housekeeping is essential for reducing or eliminating these hazards. This includes removing any feedstock spills and dust accumulations. Feedstock handling personnel may use water hoses to clean floors and equipment in the crushing, grinding, and storage facilities.

4.8 SUMMARY

The design of feedstock handling systems for gasification plants requires knowledge of the plant location in relation to the feedstock source and product market and the required feedstock capacity for the facility. Handling facilities include feedstock weighing and sampling systems to ensure receipt of the required quantity and quality of feedstock.

Feedstock quality parameters should include requirements for moisture content and fines content, as these affect feedstock flowability and subsequent handling system design parameters, such as bin bottom designs. Other considerations include the requirements for compaction in storage to limit spontaneous combustion. Finally, good housekeeping is paramount in any handling system, as this mitigates the potential for dust explosions, fugitive dust, and respirable dust concerns.

[†] Available at *https://www.cdc.gov/niosh/.*

Endnotes

1 Metso (2020) *https://www.metso.com*, Accessed May 5, 2020.

2 McLanahan (2020) *https://www.mclanahan.com*, Accessed May 5, 2020.

3 D. Woodward (2012) Sampling and Analysis Workshop, CoalPrep 2012, Lexington, KY.

4 Marietta Silos (2020) Stacking Tubes for Coal Plants, Mines and Storage Facilities, *https://www.mariettasilos.com/stacking-tubes*.

5 B.J. Arnold (1995) A Guide to Coal Handling, EPRI TR-105110, EPRI, Palo Alto, CA.

6. Bergman, P.C.A., 2005, "Combined Torrefaction and Pelletization: the TOP Process." Renewable Energy in the Netherlands. ECN Biomass: ZG Petten, the Netherlands.

7 Bergman, P.C.A. & Kiel, J.H.A., 2005, "Torrefaction for Biomass Upgrading." Proceedings of the 14th European Biomass Conference and Exposition. Paris, France. ECN-RX-05-180. *https://www.researchgate.net/publication/228699171_Torrefaction_for_biomass_upgrading*.

5.0 GUIDELINES FOR THE DESIGN OF AN OXYGEN SUPPLY MODULE

5.1 INTRODUCTION

Chapter 5 focuses on the guidelines for the air separation/oxygen supply module. The air separation industry is competitive, with multiple equipment suppliers offering air separation unit (ASU) or oxygen production unit solutions for a wide range of plant sizes and applications. Substantial oxygen pipeline networks exist along the U.S. Gulf Coast and Europe, along with smaller pipelines around the globe. Depending on the facility location, oxygen pipeline access may be possible. The major air separation technologies include cryogenic separation, sorbent-based separation, and polymeric membrane-based separation units. With the objective of further reducing the cost of oxygen for particular applications and smaller scales, individual companies have been developing advanced oxygen production technologies, such as high-temperature ceramic membranes, chemisorption, and facilitated transport membranes, along with an increasing effort on process intensification. Many of the air separation suppliers are also experienced in plant operations. Beginning with supplies to steel producers in the 1950s and 1960s, these companies supplied customers with tonnage quantities of oxygen "over-the-fence" from supplier-owned air separation facilities.

The oxygen module with the best cost/performance for a specific application will result from the optimization of many variables and tradeoffs that go into its design. Therefore, the most efficient way for a project developer to explore oxygen supply module options for a project will be to engage with one or more potential ASU suppliers who can suggest potential options on a cost- and time-efficient basis. The key for the project developer in this engagement is the ability to thoroughly communicate the project's design basis premises and plant functional requirements, including reliability/availability, mechanical specifications, industry codes, preferences, and the capital/power trade-off parameter that will best guide potential suppliers toward the cost/performance objective for the oxygen module. The detail provided in these communications will evolve as the project progresses; however, it is important to be aware of all the elements that will guide a potential supplier's response.

Due to the critical need to optimize the cost/performance while maintaining process safety and reliability, the scale of individual process units that comprise the overall facility (based on current technology for most gasification applications) tends to be large to achieve necessary economies of scale and competitive economics. This generalization also applies to the air separation module that produces oxygen for the gasifier and other possible oxygen consumers in the facility. A project developer can explore with potential module suppliers the alternative possibility of (1) multiple smaller scale oxygen-producing modules comprising a single facility; (2) an oxygen facility in which some unit operations are large scale (e.g., the main air compressor), while other unit operations consist of multiples of smaller modules; or (3) individual smaller scale modules for smaller scale overall applications. Regarding this aspect, the U.S. Department of Energy's National Energy Technology Laboratory (DOE/NETL) is managing the Radically Engineered Modular Systems (REMS) Initiative, which is further discussed in the technology selection section (see Section 5.6).

The guidelines discussed throughout this chapter will enable a gasification project developer to efficiently engage with potential oxygen module suppliers to achieve objectives in the most productive manner regarding time, cost, and outcome.

5.2 LEVEL AND PURITY OF OXYGEN NEEDED TO SUPPLY THE GASIFICATION MODULE

The required purity of the oxygen entering the gasifier will be specified by the project developer and based on the requirements of the gasifier technology, the facility's product slate, and other oxygen uses internal to the facility (e.g., the sulfur recovery unit in an integrated gasification combined cycle (IGCC) plant, where hydrogen sulfide in the syngas is reacted with oxygen to enable sulfur recovery as elemental sulfur or sulfuric acid). The overall required purity is determined by an optimization study balancing the performance and economic feasibility of all oxygen consumers in the plant. For each product slate, the production of atmospheric gas products for export, such as liquid argon (LAR), liquid nitrogen (LIN), gaseous nitrogen (GAN), liquid oxygen (LOX), and rare noble gases (e.g., xenon, krypton, neon) are options if a cryogenic ASU is chosen. The nitrogen-enriched fraction must be vented if there are no export options available and the nitrogen stream is not needed in the gasification or conversion train. Rare noble gas production is limited to large ASUs and argon production will also be site specific. All these options affect the determination of the oxygen purity from the ASU unit that achieves the best economics for the overall project and dictates the oxygen purity of the oxidant to the gasifier. Examples of product slates include the following:

- Electric power only.
- Electric power plus chemicals or liquid fuels (e.g., methanol, ammonia, urea, Fischer-Tropsch [F-T] diesel).
- Electric power plus exported hydrogen.
- Electric power plus export syngas for chemicals production.
- Substitute natural gas (SNG) or chemicals only.
- Other combinations of these examples.

Typical outcomes for the above cases include (1) oxygen purity of 85 to 95 volume % for the production of electric power only via oxygen-blown IGCC and (2) oxygen purity of 99.5% for cases where liquid argon export is to be maximized and/or where certain chemical products comprise most or all of the product slate. However, in the case of ammonia production, where high-purity oxygen is needed for gasification, but nitrogen can be utilized for the ammonia synthesis reaction, more detailed study can be required. For example, Jiang and Feng concluded that the optimal oxygen purity for coal gasification-based ammonia production is 95% with cold purification of ammonia and 96% with hot purification.[1]

For gasification systems amenable to lower oxygen purities or to straight air feed, project specific optimizations can be performed to determine whether and to what extent an enriched air feed is recommended. The Nakoso IGCC plant in Japan has demonstrated the technical feasibility of a gasification application that requires just air as the oxidant to the gasifier.

Based on commercially available technologies, gasification-based projects have historically been large scale to achieve scale economies and enable competitive cost/performance in target applications. Oxidant supply for large-scale, oxygen-blown gasification systems currently (at the time of publication) favors cryogenic air separation. For substantially smaller project scales, other commercially available air separation technologies may offer better economics (e.g., sorbent-based methods, including pressure swing adsorption [PSA] or its adsorption variants, and polymeric membranes—all of which are typically associated with smaller scale and/or lower oxygen purities). Whether large or small, projects' cost/performance tradeoffs/optimization can be evaluated by the project developer with input from potential suppliers of the oxygen supply module and other major process units.

A quantitative comparison of the main options for oxygen production is provided in Table 5-1. Typical costs vary depending on scale and oxygen purity requirements, but in general, cryogenic units are most cost-effective at large scale (i.e., hundreds to thousands of tons per day capacity) for high-purity (i.e., approximately 99%) oxygen production. In theory, it is possible to have a small cryogenic air separation plant, but non-cryogenic units are usually more cost-effective when the desired production rate is less than approximately 20 tons per day, and they are often the most cost-effective oxygen production choice up to 60 tons per day and high-purity (greater than 95%) oxygen is not required.[2] Certain smaller-scale gasification applications may warrant consideration of adsorption and membrane-based air separation technologies if their oxygen production needs fall into the lower range, and/or lower oxygen purity is acceptable (e.g., for power production options). Process flexibility in start-up is generally better with non-cryogenic methods, so a smaller/modular system relying on these air separation methods would be better adapted for intermittent or fluctuating utilization.

Table 5-1. Comparison of air separation technologies[3]

TECHNOLOGY AND DEVELOPMENT STAGE	O_2 PURITY	CAPACITY (TONS PER DAY)	POSSIBLE BYPRODUCTS; QUALITY	ENERGY DEMAND (KWH/ TON O_2)	DRIVING FORCE	STARTUP TIME
Cryogenic, Matured	99+	4,000 (single train)	Nitrogen, Argon, Krypton, Xenon; very good	200	Electricity	hours/days
Adsorption, Matured	95+	up to 300	Nitrogen; bad (ca. 11% O_2)	500	Electricity, heat (70 to 90°C; 158 to 194°F)	minutes/hours
Membrane (Polymer), Matured	~40	up to 20	Nitrogen; bad	N/A (low purity)	Electricity	minutes
Membrane (ITM), R&D Phase	99+	pilot scale	Nitrogen; bad	400	Electricity, heat (800°C; 1,472°F)	hours

The available commercial and more advanced research options are displayed in Table 5-1, along with other emerging non-cryogenic air separation technologies (e.g., advanced oxygen sorbents and ceramic hollow fiber membranes) targeted initially at smaller gasification applications are being developed by NETL.

5.3 NEED FOR PURIFIED NITROGEN FOR OTHER DOWNSTREAM MODULES

The flow rate and purity requirements for nitrogen are dependent on the requirements of the other technologies employed in the facility, the intended product slate, and other elements of plant design (e.g., the degree of integration between the air separation module and the gas turbine in an IGCC plant), as well as for general plant operating and maintenance (O&M) uses. Examples of possible nitrogen requirements are as follows:

- High purity nitrogen carrier gas to inject dry pulverized feedstock into a dry feed gasifier. An example is the Shell dry feed gasifier used for the Nuon Buggenum IGCC plant (Buggenum, The Netherlands).
- Liquid nitrogen for speeding up cooldown of a cryogenic air separation plant. (This would be a very infrequent use of nitrogen.)
- High-purity liquid nitrogen (LIN) as an export for traditional industrial gas uses and/or for make up to a LIN storage tank.
- High-purity nitrogen for plant purge gas.

- Nitrogen for combustion turbine nitrogen oxide (NO_X) control via lowering the adiabatic flame temperature in the combustor and turbine expander mass flow augmentation. An example is nitrogen with less than 2% oxygen in an IGCC plant. (IGCC cycles without such integration between the gas turbine and the air separation plant typically accomplish NO_X control with fuel gas moisturization instead of nitrogen.)
- High-purity nitrogen for ammonia and/or urea production.
- High-purity nitrogen gas for use in heating up process modules during plant startup.
- Nitrogen for backup of plant process nitrogen flow to the combustion turbine in the event of an interruption of nitrogen gas production. (Note: Due to the typically very high flow rate of nitrogen to the combustor for IGCC plants [e.g., the Tampa Electric IGCC plant in Polk County, Florida (USA)] providing for backup from stored nitrogen would be extremely challenging. Steam for this backup application is the typical alternative, but special attention should be paid to the time lag between the cessation of process nitrogen gas and the onset of steam injection, as NO_X emissions will increase during this time. However, if a mixture of nitrogen and steam is used, there may not be a lag issue. An example of a nitrogen backup arrangement is shown in Figure 5-1, with backup strategy and supporting system choices subject to the noted considerations and issues.)

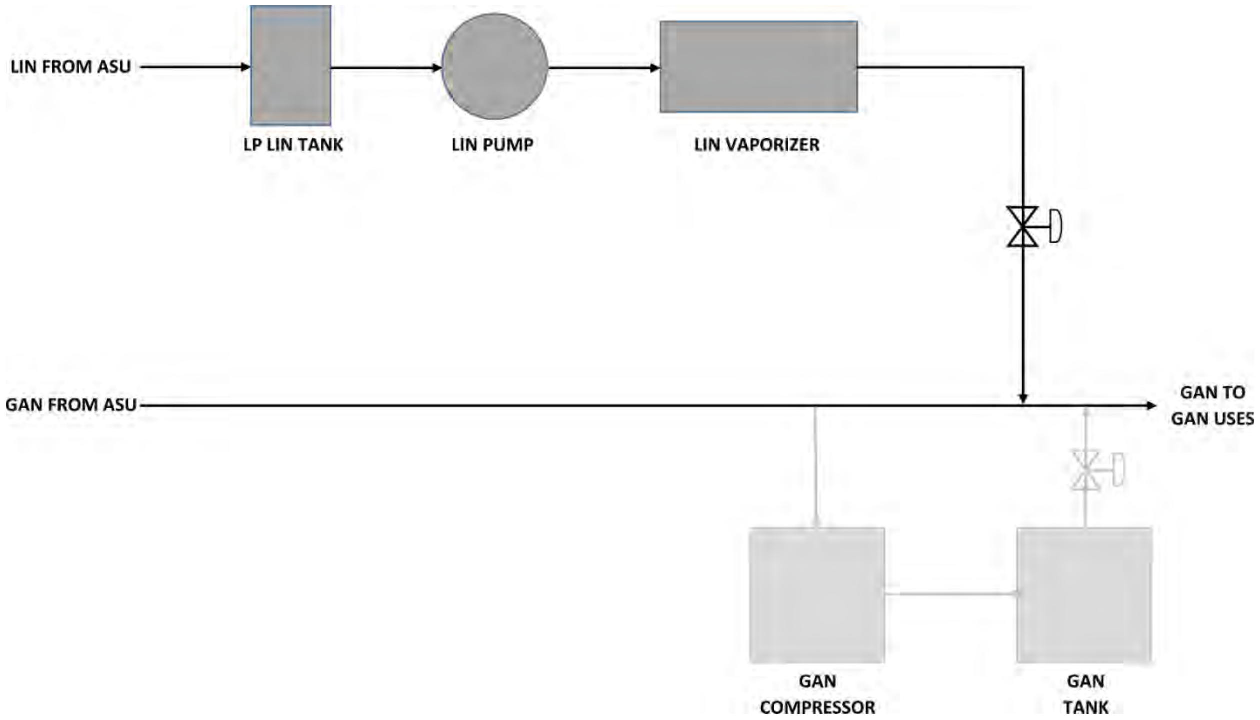


Figure 5-1. Liquid nitrogen (LIN) and gaseous nitrogen (GAN) backup arrangement

5.4 FEED PRESSURE FOR AIR/OXYGEN SUPPLY TO THE GASIFIER AND APPROPRIATE BLOWER, COMPRESSOR, OR PUMP

The project developer communicates the gasifier's requirements for oxygen (or air) and possibly nitrogen flow rate, purity, and pressure to be delivered to the gasifier, to potential oxygen or air module suppliers, along with similar requirements for other process units in the facility. In addition, the project developer provides module suppliers with project design basis premises, module functional requirements, the capital/power consumption tradeoff parameter, and other specification detail consistent with the development stage of the project. Given this information, potential suppliers should have an adequate basis to develop oxygen or air module designs to satisfy those requirements.

Providing the tradeoff parameter is more productive than simply instructing potential suppliers to emphasize low capital or low power consumption. This parameter is expressed in units of capital equivalent $/kW and can be used to quantitatively and uniformly guide module designers to minimize ([Module Capital Cost-installed] + [Tradeoff Parameter in $/kW] x [Module Power Consumption in kW]). Minimizing this sum does not account for costs associated with reliability/availability and maintenance. Nevertheless, it should be considered for guiding and ultimately selecting the optimum air separation plant design. A project developer's tradeoff parameter can vary materially from project to project depending on the objectives and financial criteria and affects module process cycle design, including the way the project's oxygen and possibly nitrogen delivery pressure requirements are optimally satisfied. For example, a relatively high $/kW tradeoff parameter places a high capital cost equivalent value on reducing power consumption in the oxygen module. When a module supplier optimizes a module process design and defines equipment specifications, such a high ratio would motivate the inclusion of more heat exchange surface area (e.g., in a cryogenic ASU) at a higher capital cost to achieve smaller pinch points in the heat exchange between process streams internal to the module for higher efficiency and lower module power consumption per amount of product. A high tradeoff parameter would also motivate designing for less pressure drop in the heat exchangers. In the case of a pumped LOX cycle, the LOX pump could achieve the required gasifier oxygen delivery pressure with a lower discharge pressure. The reverse is true for lower tradeoff parameters.

Furthermore, because various potential module suppliers may choose differing process cycle configurations to satisfy the project's required oxygen delivery pressure, the blower, compressor, or pump used to achieve that delivery pressure will be part of each potential supplier's individually optimized overall package and likely differ from one supplier to the other. This is another reason for machinery differences that potential suppliers will have in their designs. For example, the same oxygen flow rate and delivery pressure could be supplied by either an elevated pressure/pumped LOX cycle, a gaseous oxygen (GOX) compression cycle, or variations thereof (e.g., a partial pumped LOX cycle). Machinery selection could also be impacted by a project developer's specified preference (e.g., to exclude gaseous oxygen compression from consideration).

5.5 LIQUID OXYGEN STORAGE FOR SUPPLY INTERRUPTIONS

The project developer can specify backup LOX storage volume and send-out capacity based on industry experience with similar applications and should also solicit proposals from potential oxygen module suppliers for consideration of the following:

- Annual expected number and duration of each module's planned and unplanned outages. In general, the ASU should have high reliability and little or no extended outages compared to the gasifier.
- Potential supplier's track record of module reliability/availability without considering backup.
- Functionality of the proposed backup system, including speed of response to an interruption of gaseous oxygen production.
- Access to gaseous oxygen pipelines.

For example, suppose the expected annual on-stream availability for a cryogenic ASU was about 98%, resulting in approximately seven days per year of outages. If half of the outages were for planned maintenance, such oxygen module maintenance could be scheduled to coincide with planned maintenance for other process units in the facility. The remaining 1%, and the estimated number and duration of outages that comprise it, could then be used as the basis to estimate the amount of backup needed to extend the oxygen availability to near 100% of the plant's yearly full-load operating time. Such a breakdown of the outages comprising the 1% in this example is an extremely important consideration, because even a relatively short outage could result in a much longer outage of the overall facility. The combination of track record, backup technology, and the controls package is critical in the selection of the oxygen module supplier.

Backup oxygen storage would consist of stored LOX, or a combination of stored LOX and GOX, with high pressure gas storage based on the response time (following an oxygen module outage) for the LOX pumping and vaporization system to deliver oxygen at the pressure required by the gasifier. An example is LOX storage equivalent to eight total hours of process oxygen flow to the gasifier and any other process oxygen users, plus additional high-pressure GOX storage equivalent to five minutes of process oxygen flow, assuming five minutes are required to bring the LOX pumping and vaporization system up to the capacity needed to satisfy the gasifier minimum oxygen delivery pressure requirement.

If the LOX backup system can be designed and operated to quickly kick in enough to prevent the oxygen delivery pressure from dropping below the gasifier's requirement, or if the system can accept oxygen interruptions of the time needed for the LOX system to respond, the need for high-pressure oxygen gas storage would be removed. If practical for the particular project, eliminating high-pressure oxygen gas storage and delivery systems is advantageous, because of the cost and the site safety review ramifications of storing enough oxygen gas above the gasifier use pressure to cover even a relatively small time interval between any oxygen module outage and the LOX backup operating at the required rate(s).

Another consideration that could affect the oxygen (and possibly nitrogen) backup analysis is related to the number of individual modules that comprise the total oxygen (and possibly nitrogen) gas production. For example, if there were two modules, a backup approach could be to design the backup system capacity to cover the outage of one of those modules, with similar logic for larger numbers of modules in parallel with analysis and due consideration of any events that could lead to the simultaneous outage of all oxygen modules, but not the outage of any other process units in the facility.

For these reasons, a potential supplier's operating experience and the systems and controls technology for (1) minimizing any module outages and (2) minimizing or eliminating any time gap between oxygen gas production and design operation of the LOX backup system are very important criteria in selecting the ASU supplier. The project developer can attempt to deal with oxygen module availability and number and duration of module outages in the supply contract, but, in general, will find such an approach more practical if the module supplier would also be contracted for O&M of the module. An example of a LOX/GOX backup arrangement is shown in Figure 5-2.

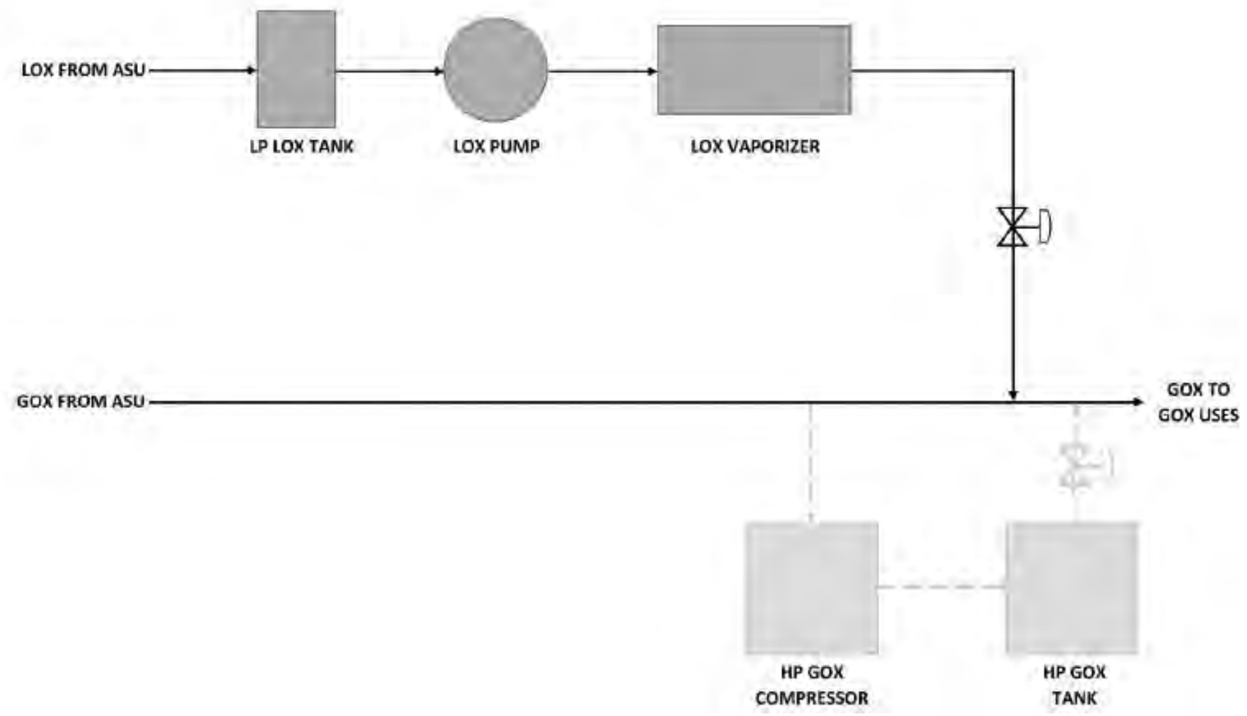


Figure 5-2. Liquid oxygen (LOX) and gaseous oxygen (GOX) backup arrangement

For those applications requiring both oxygen and nitrogen, similar logic and analysis could apply to process nitrogen backup, although for the applications involving large process nitrogen flows to the gas turbine combustor for NO_X control and expander flow augmentation (e.g., nitrogen-integrated IGCC) steam injection for backup fuel diluent is the likely alternative. One approach is to have a LIN pumping/vaporization system provide nitrogen to the combustor for the time interval between an interruption of nitrogen gas flow from the ASU and the time when the required steam flow has been achieved. If such a backup LIN vaporization system was not designed and/or operated to quickly kick in enough to prevent an interruption of diluent flow to the combustor, there would be a temporary increase in NO_X emissions. If this temporary increase was unacceptable, then GAN storage could be used to prevent the loss of diluent to the combustor. However, because of the very large nitrogen diluent flow to be replaced, the project developer should strive to avoid this need. Using a nitrogen-steam mixture can mitigate or eliminate the lag issue, as the steam flow rate can simply be ramped up to replace the missing nitrogen on the fly.

For a cryogenic ASU including LOX and possibly LIN process backup storage, replenishing those backup product(s)—possibly in combination with somewhat reduced production of primary products and/or when operating at more favorable ambient temperatures—becomes part of the module's functional requirement, along with the usual provision to enable LOX and LIN to be able to be hauled in from other locations and offloaded to plant storage. Regarding LIN storage, as covered elsewhere in this chapter, for most oxygen module technologies, there will be a need for LIN storage and vaporization (and replenishment) for traditional plant utility uses.

For applications where it is practical to back up the syngas supply from the facility with natural gas (e.g., some IGCC situations), the oxygen and nitrogen backup supply systems could be reduced in scope and cost.

5.6 AIR SEPARATION TECHNOLOGY

The project developer should select the technology for air separation by engaging with one or more potential air separation equipment suppliers and communicating the ASU's design basis premises (including expected use patterns for the facility's products); functional performance requirements (including reliability, availability, and maintainability); and other project requirements and developer preferences, such as mechanical specifications, process simplicity, tradeoff parameter for calculating the capital equivalent value of power consumption, and safety requirements (see Section 5.9).

Based on such requirements, potential suppliers are asked to provide estimated capital and operating costs for their proposed ASUs. Although different suppliers may propose different commercially available air separation technologies for the project developer to consider, the scale and oxygen purity requirements of most current gasification applications producing energy products or basic chemicals will likely be optimally satisfied with cryogenic air separation technology. The possibility for integration between the air separation module and the rest of the facility can also support the choice of cryogenic air separation. This is due to the many design parameters—both within the ASU itself and regarding its integration with the rest of the facility—that can be varied in optimizing to achieve a desired cost/performance outcome. For example, within the ASU, the number of distillation columns and the amount of heat exchanger surface area can be varied based on the project developer's capital cost/power consumption tradeoff ratio. And with regard to integration, for an IGCC application where a large flow of nitrogen gas supply to the gas turbine combustor can be used for NO_X control and expander flow augmentation, the ability to economically produce such a nitrogen stream at the required purity via an elevated pressure air separation cycle would further support the choice of cryogenic air separation.

However, the requirements for smaller capacity oxygen modules (e.g., for distributed power production) could bring non-cryogenic air separation technologies into consideration. The challenge to date has been the difficulty in achieving competitive cost/performance for small gasification-based systems. Toward this objective, the NETL REMS Initiative is underway. This initiative is comprised of several projects aimed at developing advanced non-cryogenic technologies and modular manufacturing capabilities to achieve acceptable cost/performance for smaller/modular gasification-based systems to produce power, fuels, and chemicals. Many advanced, non-cryogenic air separation projects are included as part of this initiative.

As other advanced non-cryogenic technologies and modular manufacturing approaches continue development toward competitive cost/performance in initial target markets and beyond, a project developer's approach to technology selection can follow the same logic and guidelines outlined herein for the selection of currently commercially available air separation technologies.

Currently, proven non-cryogenic technologies for oxygen production are primarily adsorption and polymeric membranes, which are discussed in the following paragraphs. Other developmental or less practiced non-cryogenic technologies include (1) chemical absorption in which, for example, a molten salt is oxidized by reacting with oxygen in the air feed stream, followed by desorbing that oxygen in a reducing step at a different set of conditions; (2) ion transport membranes, in which a ceramic membrane operates at high temperature and involves oxygen molecules ionizing on the feed surface of the membrane, passing through the membrane as ions, and recombining and shedding electrons at the product side of the membrane to form oxygen molecules, with the electrons passing in reverse through the membrane to the feed side where the process continues; and (3) facilitated transport membranes, where oxygen reacts with a substance on the feed side of the membrane, the reaction product passes through the membrane, and the oxygen is released on the product side. Examples of a chemical absorption-based air separation process and an ion transfer membrane-based air process are shown in Figures 5-3 and 5-4, respectively.

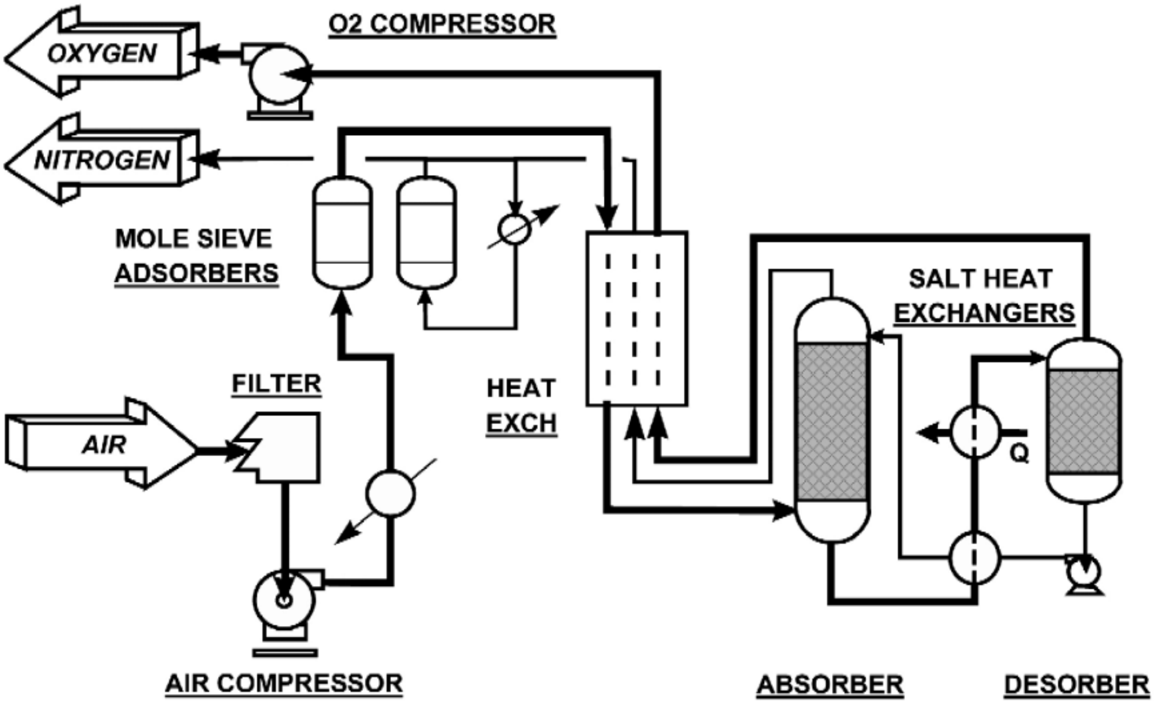


Figure 5-3. Chemical air separation process[4]

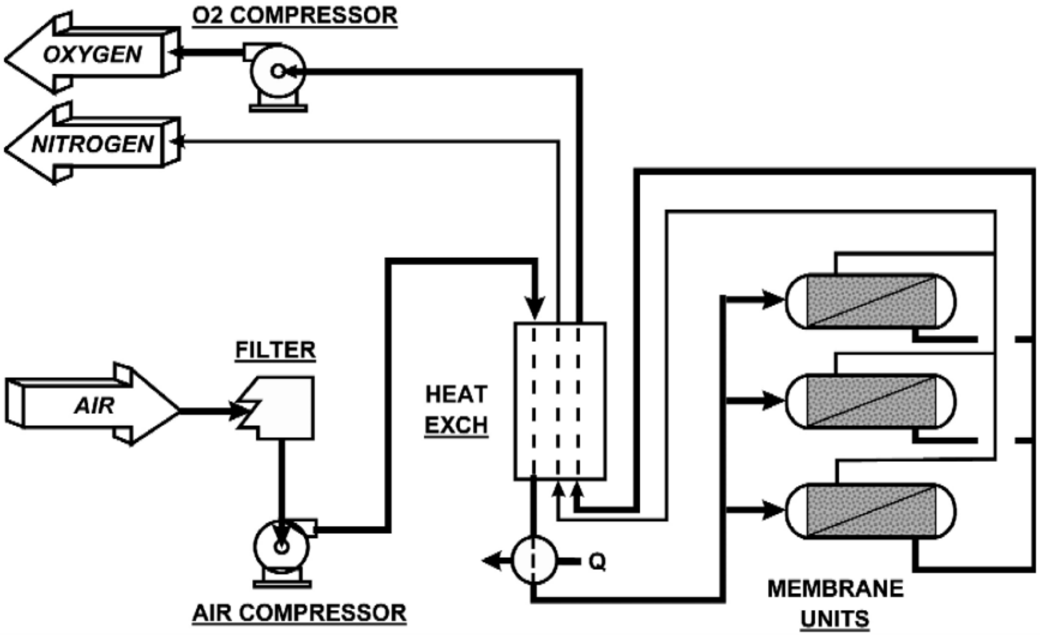


Figure 5-4. Ion transfer membrane air separation process[4]

5.6.1 ADSORPTION-BASED AIR SEPARATION

Adsorption-based air separation for oxygen is a proven technology and most useful for smaller applications that do not require high-purity oxygen, high-purity nitrogen gas co-product, or cryogenic liquid (e.g., LOX, LIN, LAR) co-products. Adsorption cycles for oxygen production are typically based on zeolite adsorbents that preferentially adsorb nitrogen and offer the benefit of simplicity and quick start up and shutdown. Pressurized air feed enters the adsorbent bed, with nitrogen being preferentially adsorbed on the zeolite and an oxygen-enriched stream passing through, typically for applications requiring purities up to approximately 95 volume % oxygen. When the adsorbent bed is saturated with nitrogen, the feed air is sent to a second regenerated bed in parallel with the first, while the first bed is regenerated using a purge gas. The process continues in like fashion by switching between the two beds whenever the bed currently in use becomes saturated. Saturated beds are regenerated either by heat up (temperature swing adsorption [TSA]), pressure reduction (pressure swing adsorption [PSA], vacuum swing adsorption [VSA], or vacuum-pressure swing adsorption [VPSA]) or both (pressure and temperature swing adsorption [PTSA]). Oxygen production rate is proportional to adsorbent bed volume, so adsorption processes do not scale up as cost effectively as cryogenic separation and they are typically applied to smaller applications as a result. An example of an adsorption-based air separation process is shown in Figure 5-5.

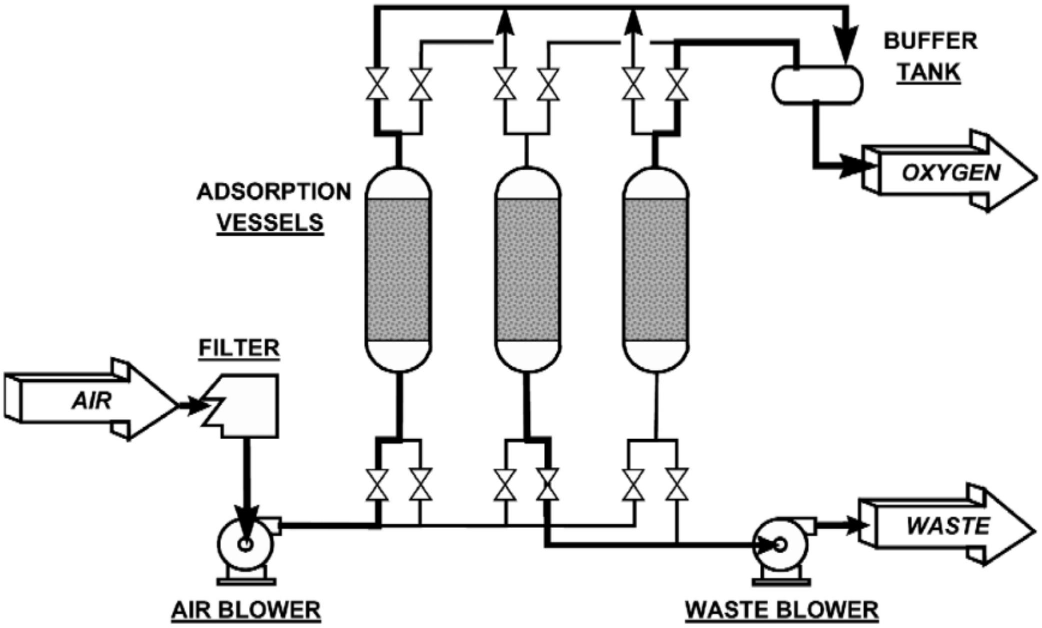


Figure 5-5. Adsorption-based air separation process[4]

5.6.2 POLYMERIC MEMBRANES

The separation of air with non-porous polymeric membranes for oxygen-enriched air production is based on oxygen molecules being physically somewhat smaller than nitrogen molecules and the ability to selectively permeate (i.e., dissolve and diffuse) through thin layers of some polymers. The oxygen permeability of a polymeric membrane is an experimentally determined property describing the ability of oxygen to permeate the membrane given an oxygen partial pressure driving force across the membrane. It can be used to calculate the flux (i.e., flow rate at standard temperature and pressure per unit of membrane area) at which oxygen can permeate the membrane for a given oxygen partial pressure difference and a given membrane thickness. The oxygen flux is proportional to the difference in oxygen partial pressure across the membrane and inversely proportional to the membrane thickness. The preferential permeability, or selectivity, of oxygen over nitrogen is relatively low, because oxygen and nitrogen molecules are relatively close in size, resulting in an oxygen-enriched air stream with oxygen purity typically up to 50 volume %. Applications are typically small due to the scaleup cost issue for polymeric membranes. An example of a polymeric membrane-based air separation process is shown in Figure 5-6.

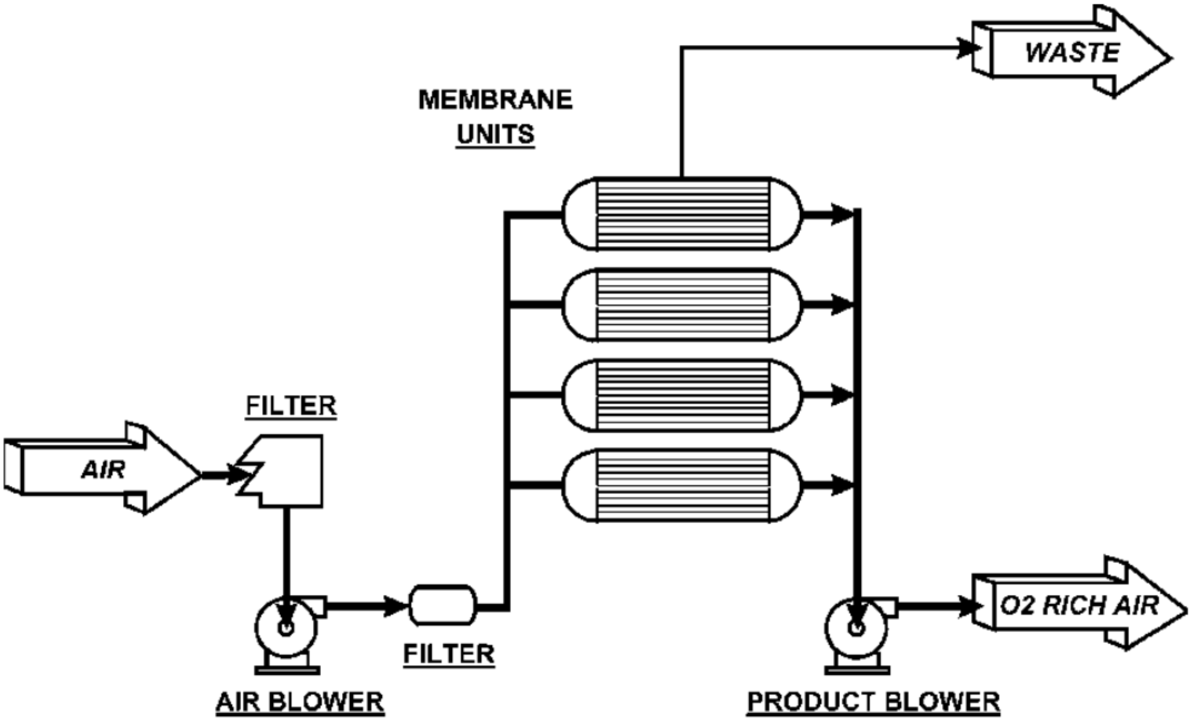


Figure 5-6. Polymeric membrane air separation process[4]

For any power generation application, it is particularly important for the project developer to analyze the relative cost/performance of options when selecting the air separation technology. These options may vary significantly in capital intensity, given the likelihood of increasing cycling and turndown of generating units to accommodate increasing amounts of renewables-based power entering the grid, which would reduce the number of operating hours in a year, with an accompanying penalty to the more capital-intensive options.

5.7 INTEGRATION WITH DOWNSTREAM MODULES

The project developer typically specifies any integration requirements after consulting with potential suppliers of the gas turbine (if any), the air or oxygen module, and any other relevant major technology suppliers. With such collaborative input, the project developer can perform cost/performance optimizations of potential integration modes, including studying the impact of integration on facility reliability/availability, turndown, cycling/ramping and other functional requirements of the overall facility. Process cycles involving integration of the ASU with the rest of the facility are a possible means of enhancing the cost/performance of the overall facility (e.g., reducing the $/net megawatt-house [MWh] cost of export power produced by an IGCC plant). Integration alternatives deserve careful analysis to ensure that added complexity does not compromise facility reliability/availability or otherwise negatively impact its required functional performance in a way that the potential benefit of integration is lost.

The following considerations affect the project developer's integration decision.

- For an IGCC facility, whether CO_2 is to be captured prior to the combustion of fuel gas in the gas turbine combustor. If such pre-combustion capture is expected, the resulting hydrogen (H_2) or H_2-rich fuel stream will have lower mass, because of the removal of CO_2. Hence, there would be lower mass in the combusted fuel stream sent to the gas turbine power generating expander, and less power produced by the gas turbine (assuming the selected gas turbine was not already chosen with this lower fuel mass input in mind). There would be a somewhat smaller motivation to take bleed air supply from the gas turbine air compressor to supply all or part of the feed air stream to the air separation module. Such an analysis, or any IGCC integration analysis, is impacted by the specific characteristics of available gas turbine equipment (i.e., an integration scheme that may look attractive with one gas turbine may not look attractive with another) as the overall plant design is developed to achieve optimal utilization of the gas turbine's design point specifications.

- Project developer preference, such as the balance between plant simplification (O&M costs and startup/shutdown times) and possible enhanced cost/performance enabled by a more integrated facility.
- For power and chemicals co-production, when nitrogen is required as a feedstock in the chemical section (e.g., for ammonia and/or urea production).
- NO_X control for an IGCC facility can be accomplished by adding steam and/or nitrogen to dilute the fuel stream and lower the adiabatic flame temperature in the combustor, reducing NO_X formation. Using nitrogen rather than steam as the diluent would favor a nitrogen-integrated ASU.

The production of a required product slate from a cryogenic ASU can be accomplished with many different process configurations and specific designs of the compression, pumping, distillation, and heat exchange equipment that comprise the ASU. In addition, a cryogenic ASU offers opportunities for efficiently utilizing energy in some of the process streams of the application it is supplying. Thus, applications requiring process oxygen and possibly process nitrogen are particularly amenable to project-specific optimization of the integration between the cryogenic ASU and the rest of the facility, based on the facility's specific product requirements, capital equivalent value of the ASU's power consumption, and facility and ASU functional requirements and design basis premises.

Regarding IGCC application, there are three primary integration modes between the gas turbine system and the ASU (shown in Figure 5-7).

- **Not integrated**, except for the supply of oxygen to the gasifier and other possible oxygen requirements (e.g., a thermal/catalytic sulfur recovery unit; overall $2\ H_2S + O_2 = 2\ H_2O + 2\ S$) and nitrogen requirements (e.g., utility nitrogen and LIN make up to storage). In this example, the feed air to the air separation module is supplied by a standalone air compressor and nitrogen for NO_X control is not supplied to the gas turbine combustor. An example is the original Wabash River IGCC repowering project in West Terre Haute, Indiana (USA), in which NO_X control was accomplished with steam dilution.
- **Nitrogen-integrated**, in which the ASU is supplied by a standalone air compressor and pressurized nitrogen from the ASU is returned to the gas turbine combustor where it lowers the adiabatic flame temperature in the combustor and reduces NO_X, while increasing mass flow to the gas turbine expander. A cryogenic air separation cycle choice for this option is an elevated pressure cycle—a cycle with higher power requirements for the main air feed compressor—later offset by higher power production in the combined cycle system. An example of this arrangement is the original Tampa Electric IGCC plant in Polk County, Florida (USA).
- **Full air/full nitrogen integrated**, in which the entire air supply to the ASU is taken as a bleed stream from the gas turbine air compressor and pressurized nitrogen is returned to the gas turbine combustor. An example is the original Demkolec project in Buggenum, Netherlands.
- An intermediate variant is a **partial air-integrated cycle** in which part of the feed air supply to the ASU is supplied by a standalone air compressor and part by an extracted stream from the gas turbine air compressor. The former Elcogas IGCC plant in Puertollano, Spain, is an example. For IGCC applications, the partial air-integrated cycle also typically includes nitrogen return to the gas turbine combustor.

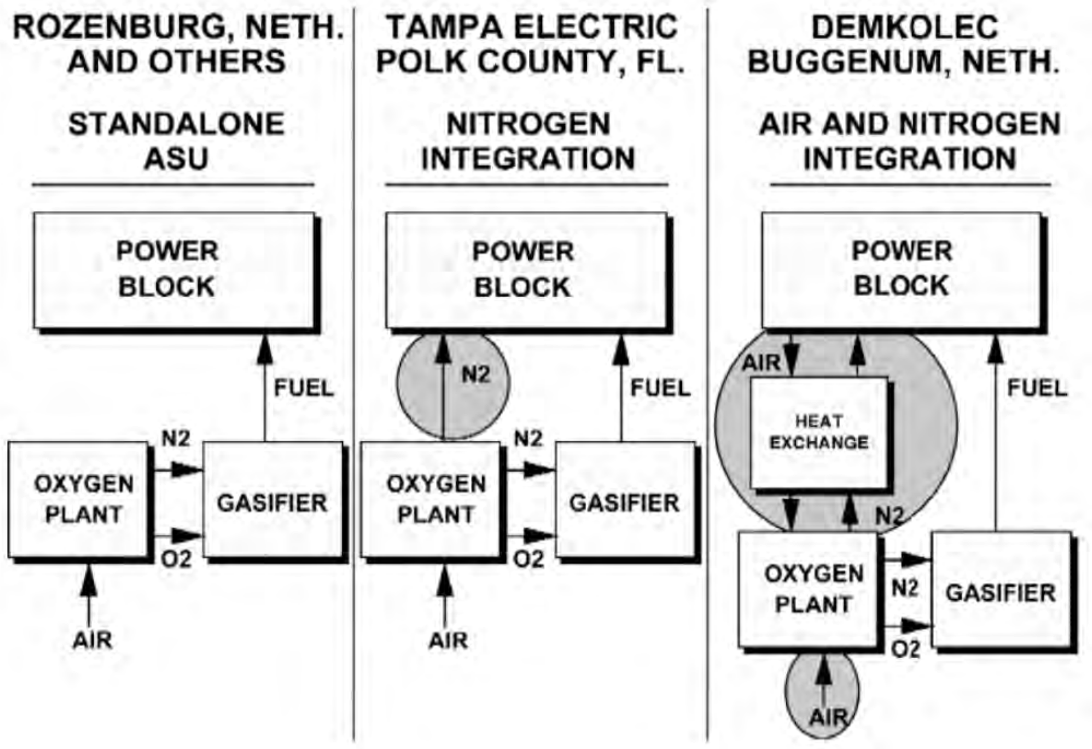


Figure 5-7. Commercial integration experience[4]

Non-cryogenic ASUs supplying oxygen and possibly nitrogen to applications also offer natural opportunities for integration (e.g., for high temperature applications where the air separation process also involves high temperature, such as ceramic ion transfer membranes). Other examples of integration of non-cryogenic ASUs with applications that could be based on coal gasification include the following:[4]

- Chemical air separation and partial oxidation process integration.
- Extracted air heat recovery for solvent-based processes.
- Extracted air heat recovery for adsorption-based processes.

5.8 PROCESS AND ANALYTICAL CONTROL REQUIREMENTS

Typically with specification/guidance from the project developer's controls specialist for the overall facility to ensure systems compatibility, the process and analytical control requirements for the air or oxygen supply module are determined by the potential suppliers for responding to and complying with (1) module functional requirements including startup, shutdown, turndown and ramping; (2) requirements for performance testing to demonstrate compliance with contract requirements; and (3) the need to support and be compatible with the remainder of the facility's analytical and control systems. Examples of important functional requirements for the ASU and the overall project (note that specifics will vary from project to project), which the design, operation, control, and performance testing of the air or oxygen supply module must support, include:

- For any gasification project, the delivery of required oxygen flow, pressure, and purity during normal operations, turndown, or ramping, all at very high reliability/availability. For an IGCC project, performance during ramping and turndown can be particularly important, given the need to accommodate an increasing amount of renewables-based power coming onto the grid and support other grid management requirements.
- For an IGCC project with co-production of one or more byproducts, functional performance as necessary to support economic optimization of the overall project's product slate and operation (given certain daily, weekly, or seasonal variations in power and co-product values).
- For a partial air-integrated IGCC cycle, in which part of the feed air to the ASU is supplied by a standalone compressor and part by a stream of extracted air from the gas turbine air compressor, the controls need to ensure that the two compressors feeding into the feed air supply line operate stably and compatibly during steady state operation and during start up, turndown, ramping (up and down), or step changes.
- The controls on the oxygen backup system must support the objective of minimizing or eliminating any time gap between an interruption of process oxygen flow and the delivery of backup vaporized LOX.
- For a nitrogen-integrated IGCC plant, the controls must support the objective of minimizing or eliminating any time gap between an interruption of process nitrogen gas flow and the delivery of vaporized LIN or other possible interim backup means of filling the time gap until backup steam diluent is available.
- Plant instrumentation must accommodate performance testing of the oxygen (or air) module in accordance with procedures and calculations agreeable to both the project developer and module supplier.
- For all gasification applications, the requirement for safe operation of the facility during all modes of operation (e.g., startup, steady state, shutdown [planned and unplanned], turndown, ramping, upset, and standby).

5.9 PROCESS AND OPERATOR SAFETY REQUIREMENTS

The air separation industry has extensive experience with safety specifications and considerations that impact the design, equipment procurement, construction, and O&M of the ASU. This section considers safety considerations related to the ASU/oxygen supply module as they should be addressed in accordance with applicable codes, standards, and knowledge bases. The project developer should have potential suppliers specify the industry codes and company standards that form the basis of their design, procurement, construction, and O&M procedures. It is also helpful to have potential suppliers note industry- and company-specific lessons learned from past safety-related events and how those lessons would be incorporated into the current project. At the appropriate stage of project development, input from potential suppliers should be reviewed by the project developer in consultation with an industry consultant having expertise in air separation safety. Once a project is underway, the detail pursued can be adjusted relative to the stage of project development (e.g., an early-stage safety review is a Preliminary Hazards Analysis [PrHA]). The oxygen module supplier should participate in that analysis as well as further reviews as the project progresses (e.g., the overall project hazards analysis [PHA], hazard and operability [HAZOP] study, and quantitative risk assessment [QRA]).

Examples of oxygen module safety-related design, procurement, construction, and O&M topics are included below (*this list, by definition, should be considered as representative only*).

- The avoidance of any hydrocarbon buildup in the air separation module (e.g., in liquid oxygen reboilers, liquid oxygen adsorbers, distillation column liquid oxygen sumps, or dead legs of piping where dead-end boiling can occur). The air separation industry has extensive experience in mitigating this hazard, but typical operations need to consider what chemicals may be present at or near the ASU intake.
- Piping and system components that encounter oxygen need to undergo special cleaning for oxygen service and be maintained in the "oxygen-clean" condition. Oxygen must not encounter hydrocarbons, including preservatives and lubricants. Cleaned for oxygen service standards are required not just for the ASU, but also for any oxygen piping in the gasifier. Allowed valve types and other components may be limited for oxygen service; the ASU industry will have developed pipe and valve specifications for oxygen systems.

- There are conditions where metals, including carbon steel and aluminum, can burn in oxygen, even if there are no hydrocarbons present—systems must be designed, operated, and maintained to prevent such events. For this reason, limiting the access to the air intake of an ASU is important, because anything that concentrates in the distillation system with the oxygen can cause fires or explosions in the ASU.
- Nitrogen gas is an asphyxiant, so vents, potential leaks, and other discharges must not be allowed to endanger breathing atmospheres.
- When cryogenic liquids (e.g., liquid oxygen, liquid nitrogen, and liquid argon) are spilled and vaporize, the cryogenic temperature vapors prior to warm up are heavier than air and can concentrate near the ground or other operating surfaces until they warm up and disperse. Designs and operating procedures need to account for and prevent hazards from any such events. The ASU should have a dump fan associated with routine vents that will help to heat and disperse any such vented gases.
- Nitrogen systems connected to oxygen systems for purging purposes must be designed and maintained so that they cannot become a means for foreign materials of any kind to enter the oxygen systems.
- Carbon steel systems cannot tolerate cryogenic temperatures and can be expected to fail if subjected to such temperatures. System designs and O&M procedures must prevent such events. An example to consider is a liquid oxygen storage, pumping, and vaporization backup system for situations where gaseous oxygen production is interrupted. Such systems may tie into carbon steel piping carrying oxygen gas to the gasifier and other oxygen-using systems in the plant. It is necessary that the backup system and what it ties into be designed and operated so that cryogenic liquids or gases do not enter plant sections that contain carbon steel, because of the potential for failure of a control valve, instrument, other system failure, or other unexpected operating event.
- As with any system carrying cryogenic liquids or gases, designs must prevent system sections from becoming blocked in without adequate thermal relief and proper discharge to relieve system pressure buildup when cryogenic liquids vaporize and/or cold gases warm up. For example, a section with two block valves in series must be designed with such relief and discharge.
- Safety reviews related to the ASU must also include considerations related to other units in the facility, which could contribute to ASU safety issues. An example is the situation that can occur with a partial air-integrated process for an IGCC plant. In such an arrangement, part of the air feed to the ASU is from the gas turbine air compressor. This portion of feed air must be subject to the same scrutiny regarding potential contaminants as the standalone air compressor supplying the remainder of the air feed. Another example is the location of the feedstock pile in relation to the air compressor intake.
- Designs and O&M procedures must account for contamination of the atmospheric air feed to the air separation module, such as from airborne hydrocarbons. Small amounts of hydrocarbons are expected in atmospheric air and a traditional part of the air composition design basis. These baseline concentrations, as well as excursions above the baseline design amounts, including species such as soot from fires, must be accounted for in plant designs and O&M protocols.
- Design and O&M procedures for the air separation module must ensure that any oxygen compressors be located inside properly designed personnel barriers and that personnel are not allowed inside such barriers when the compressor(s) are compressing oxygen.

In addition, safety-related elements of the ASU design, construction, and O&M requirements that are not specific to air separation or its products (e.g., ladders, piping clearances for operator access, other applicable OSHA requirements) must also be reviewed and appropriately acted upon to ensure safety.

Considerations and issues related to the specific project at hand should be considered by the appropriate team of participants to assure a properly tailored list of project-specific issues. Furthermore, industry codes and standards related to air separation safety are always subject to updating and revision. Current versions of such codes and industry knowledge bases should be referred to consistent with project timing.

Numerous examples of air separation safety-related papers, guides, and codes are publicly available.

- European Industrial Gas Association (EIGA) Doc. 65/13 Safe Operation of Reboilers/Condensers in ASUs.[5]
- Reboiler/Condenser Safety.[6]
- EIGA Safe Practices Guide for Cryogenic Air Separation Plants IGC Document 147/13/E.[7]
- Safe Design and Operation of a Cryogenic ASU.[8]
- Guides, codes, and standards subject to updating and revision, so users should consult the latest versions in their safety reviews. Some examples of industry guides and documents in draft form (at the time of publication) include:
    - Compressed Gas Association (CGA) Safe Practices Guide for Cryogenic Air Separation Plants (D-17026).[9]
    - CGA Oxygen Pipeline and Piping Systems (D-16039).[10]

5.10 ADDITIONAL DESIGN AND OPERATING ISSUES

In addition to the guidance noted in earlier sections of this chapter, the following issues are examples of other considerations for the project developer when defining the air or oxygen module:

- The track record of reliability/availability for cryogenic ASUs is generally very good and includes many examples of very high on-stream factors. Among the many elements that support such performance, spare parts provisioning decisions for large items without which the plant cannot operate are very important (e.g., the rotor on the main air compressor).
- In defining the maximum turndown point for a cryogenic air separation plant, the project developer should engage with potential suppliers to ensure that designing for stable operation at maximum turndown will not affect the design in a way that compromises efficient operation at the design point

(e.g., by the supplier reducing the diameter of a distillation column to build in more pressure drop). In some cases, the project developer may choose to accept such a compromise, but it should be a choice with knowledge of relative cost/performance.

- The project developer should exercise caution regarding the mechanical specifications to be applied to the oxygen module, as mechanical specifications from other industries can result in unnecessarily higher costs that could be avoided by accepting industrial gas industry specifications, with appropriate review by the project developer.

5.11 SUMMARY

The guidelines discussed throughout this chapter enable a gasification project developer to efficiently engage with potential oxygen module suppliers to achieve objectives in the most productive manner regarding time, cost, and outcome. The air separation industry is competitive, with multiple equipment suppliers offering ASU or oxygen production unit solutions for a wide range of plant sizes and applications. The major air separation technologies include cryogenic separation, sorbent-based separation, and polymeric membrane-based separation units. With the objective of further reducing the cost of oxygen for particular applications and smaller scales, individual companies have been developing advanced oxygen production technologies, such as high-temperature ceramic membranes, chemisorption, and facilitated transport membranes, along with an increasing effort on process intensification.

The oxygen module with the best cost/performance for a specific application will result from the optimization of many variables and tradeoffs that go into its design. Therefore, the most efficient way for a project developer to explore oxygen supply module options for a project will be to engage with one or more potential ASU suppliers who can suggest potential options on a cost- and time-efficient basis.

Due to the critical need to optimize the cost/performance while maintaining process safety and reliability, the scale of individual process units that comprise the overall facility (based on current technology for most gasification applications) tends to be large to achieve necessary economies of scale and competitive economics. This generalization also applies to the air separation module that produces oxygen for the gasifier and other possible oxygen consumers in the facility.

Endnotes

1 J. Jiang and X. Feng (2019) "Energy Optimization of ammonia synthesis processes based on oxygen purity under different purification technologies", Elsevier, Energy 185(C) abstract.

2 Universal Industrial Gases, Inc. (2016) "Non-Cryogenic Air Separation Processes", *http://www.uigi.com/noncryo.html*.

3 T. Banaszkiewicz. M. Chorowski. W. Gizicki (2013) "Comparative analysis of oxygen production for oxy-combustion application", Proceedings of the 7th Trondheim CCS Conference. Trondheim, Norway, June 5-6, 2013, *https://core.ac.uk/download/pdf/82425913.pdf*.

4 A.R. Smith and J. Klosek (2001) "A review of air separation technologies and their integration with energy conversion processes," Air Products and Chemicals, Inc., Fuel Processing Technology 70.

5 Asia Industrial Gases Association (n.d.) "Safe Operation of Reboilers/Condensers in Air Separation Units, AIGA 035/13", AIGA.

6 W. Schmidt (February 2, 2006) "Reboiler/Condenser Safety," Air Products and Chemicals, Inc., European Industrial Gas Association Safety Seminar.

7 Asia Industrial Gases Association (n.d.) "Safe practices guide for cryogenic air separation plants, AIGA Document 147/13/E", AIGA.

8 W.P. Schmidt, K.S. Winegardner, M. Dennyhy, H. Castle-Smith (April 2001) "Safe Design and Operation of a Cryogenic Air Separation Plant," Air Products and Chemicals, Inc., presented at the AIChE 35th Annual Loss Prevention Symposium.

9 CGA, "Safe Practices Guide for Cryogenic Air Separation Plants" (D-17026), CGA.

10 CGA, "Oxygen Pipeline and Piping Systems" (D-16039), CGA.

6.0 GUIDELINES FOR THE DESIGN OF A SYNGAS CLEANUP AND/OR CONVERSION MODULE

6.1 INTRODUCTION

The objective of the syngas cleanup process is to integrate the gasification and syngas conversion processes and for the removal of syngas contaminants to achieve the environmental emission targets for the overall plant. To accomplish this objective, the syngas cleanup system must transform the raw syngas product leaving the gasification process into a treated syngas stream that is suitable to produce the target product via the selected syngas conversion process.

The objective of transforming the raw syngas into a suitable clean syngas stream for the downstream syngas conversion process and separating the contaminants for environmental emission goals requires:

- Removal of entrained solids.
- Removal of gaseous contaminants (e.g., chlorides, sulfur).
- Shifting the syngas composition to achieve a specific hydrogen (H_2)-to-carbon monoxide (CO) ratio.
- Adjusting the syngas temperature.
- Adjusting the syngas pressure.

Although all these transformations are required for both power and chemical production, the specific transformation requirements differ for power and chemical production. Some of these requirements are mandated by regulations, while others are defined by catalyst, process, or material specification requirements.

When the design of the syngas cleanup system begins, the following design information is available to a project developer:

- The specific feedstock for the plant.
- A defined gasification process that produces a syngas with known composition, temperature, and pressure (based on the specified fuel).
  - The gasification process may include an air separation unit (ASU) based on the selected conversion process and process economics for the plant.
- The selection of an appropriate syngas conversion process to generate the target end product(s).

The available information on permissible contaminant/emission limits for gasification is summarized in Table 6-1. The table includes the regulatory requirements established by the U.S. Environmental Protection Agency (EPA) (EPA Emission Limits [May 3, 2011]) and the corresponding concentration of these contaminants in a coal-derived syngas mixture (see column titled “Syngas Concentration Limits”). In the columns with regulatory requirements, “NR” indicates the specific contaminants that do not currently have a regulatory requirement. The remaining columns provide contaminant concentrations that can be tolerated in power systems (integrated gasification combined cycle [IGCC]), methanol production, Fischer-Tropsch (F-T) hydrocarbon synthesis, and solid oxide fuel cell (SOFC) applications. In general, the syngas contaminant concentration limits for chemical production are lower than for power production, indicating higher levels of syngas cleanup are required for chemical applications (Table 6-1).

Table 6-1. Syngas contaminants in gasification systems

SYNGAS CONTAMINANT	EPA EMISSION LIMITS (MAY 3, 2011)	SYNGAS CONCENTRATION LIMITS	TOLERANCE LIMITS BY APPLICATION			
			POWER SYSTEMS (IGCC)	METHANOL	FISCHER-TROPSCH	SOFC
Sulfur (in H_2S, COS)	New Source Performance Standards (NSPS) (SO_2): 1.0 lb/MHh gross; or 1.2 lb/MWh net; or 97% reduction[1,a]	500.0 ppm$_w$ (S)[b,c]	750.0 ppm$_v$[2,d]	0.060 ppb[2]	10.0 ppb[3,4,e]	60 ppbv[3]
Nitrogen (in NH_3, HCN)	NSPS (NO_X expressed as NO_2): 0.70 lb/Meh gross; or 0.76 lb/MWh net[1,a]	213.0 ppm$_w$ (N)[b,f]	40.0 ppm$_v$[2]	0.01-10 ppm[2,5]	20.0 ppb[3,4]	5,000 ppm[6]
Particulates	NSPS (Filterable PM): 0.090 lb/MWh gross; or 0.097 lb/MWh net[1,7,a,g]	90.0 ppm$_w$[b]	700.0 ppb$_v$[2]	100 ppb$_w$[2]		
Mercury (Hg)	Mercury and Air Toxics Standards (MATS): 0.0030 lb/GWh gross[7]	3.0 ppm$_w$[b]				0.1-180 ppm[8,9]
Halogens (Cl, Br, F, I)	MATS: Option 1: HCl – 0.0020 lb/MWh; or Option 2: SO_2 – 0.40 lb/MWh[7]	1.9 ppm$_w$ (Cl)[b,h]	5.0 ppm$_v$[2]	10 ppb$_v$[2,10,11]	10.0 ppb[12,13]	100.0 ppb[14,15,j]
Alkalis (Na, K, Li)	NR	NR	100.0 ppb$_v$[2]	10-40 ppb$_v$[2,5]	10.0 ppb[12,13]	No effect[15,j]
Phosphorus (P)	NR	NR	300.0 ppb$_v$[2,c]			<1 ppm[8,16,17,k]
Arsenic (As)	0.02 lb/GWh gross[7]	20.0 ppb$_w$[b]			1-10 ppb$_v$[18,i]	<0.1ppb[3,8,9,19,k]
Selenium (Se)	0.30 lb/GWh gross[7]	300.0 ppb$_w$[b]				<200ppb[9,k]
Cadmium (Cd)	0.0020 lb/GWh gross[7]	2.0 ppb$_w$[b]				No effect[15,m]
Vanadium (V)	NR	NR	10.0 ppb$_w$[3,20,n]			No effect[15,m]
Calcium (Ca)	NR	NR	40.0 ppb$_w$[2,3,20,n]			
Barium (Ba)	NR	NR	100.0 ppb$_v$[2,n]			
Lead (Pb)	0.0090 lb/GWh gross[7]	9.0 ppb$_w$[b]	20.0 ppb$_w$[2,3,20,n]			No effect[15,m]
Manganese (Mn)	0.020 lb/GWh gross[7]	20.0 ppb$_w$[b]	200.0 ppb$_v$[2,n]			
Iron (Fe)	NR	NR	40.0 ppb$_v$[2,o]	100.0 ppb[2]		
Nickel (Ni)	0.070 lb/GWh gross[7]	70.0 ppb$_w$[b]	40.0 ppb$_v$[2,o]	100.0 ppb$_v$[2]		
Magnesium (Mg)	NR	NR	40.0 ppb$_w$ [3,20]			
Antimony (Sb)	0.020 lb/GWh gross[7]	20.0 ppb$_w$[b]				No effect[15,m]
Beryllium (Be)	0.0010 lb/GWh gross[7]	1.0 ppb$_w$[b]				No effect[15,m]
Chromium (Cr)	0.040 lb/GWh gross[7]	40.0 ppb$_w$[b]				No effect[15,m]
Cobalt (Co)	0.0040 lb/GWh gross[7]	4.0 ppb$_w$[b]				No effect[15,m]

Note: NR = No existing regulatory requirement(s).

a. The emission resulting from the combustion of fuels in a stationary combustion turbine are subject to 40 CFR subpart GG.

b. Assumes a heat rate of 7,000 British thermal units (Btu)/kWh and a syngas heating value of 7,000 Btu/lb.

c. This value represents the sulfur content of the fuel, while the emissions limits represent sulfur dioxide (SO_2).

d. Sulfur tolerance limit for IGCC relates to power island heat recovery steam generator (HRSG) protection; specification exceeds emission limits.

e. Co- and Fe-based catalyst are sensitive to sulfur. An exception is cobalt/molybdenum-based sour-gas shift (SGS) catalyst, because the active component is cobalt-molybdenum sulfide.

f. Assumes all nitrogen oxides (NO_X) due to nitrogen in fuel and does not include thermal NO_X formation and emissions.

g. MATS (filterable PM) limits have three options: (1) 0.070 lb/MWh gross electric output for duct burners on syngas; (2) 0.090 lb/MWh gross electric output for duct burners on natural gas or 20 total non-Hg hazardous air pollutants (HAP) metals 0.04 lb/GWh; or (3) Individual metals limits shown in table.

h. This value represents the chlorine content of the fuel while the emissions limits represent HCl.

i. A 13 to 50% loss[5] of SOFC performance at hydrogen chloride (HCl): 20 to 160 parts per million (ppm); halogen limits.[16]

j. Alkalis form a condensed phase under typical cleanup conditions and can be effectively removed before reaching the SOFC anode.

k. Multiple lab studies show that 1 to 20 ppm phosphine (PH_3) or HPO_2 causes severe, irreversible anode degradation accompanied by a substantial ohmic and electrode cell loss in SOFC and limits less than 1 ppm are recommended.[8,16,17] Studies for arsine (AsH_3) as low as 5 parts per billion (ppb) severely degrade the SOFC electrode by irreversible reaction with the electrode Ni materials.[3,8,9,17,19] Thus, As limits must be at the most less than 1 ppb. Studies of Se at 0.2 and 5 ppm showed significant degradation of SOFC performance.

l. As is potent F-T catalyst poison (i.e., forms Fe and Co arsenides at operating conditions of F-T).

m. Study on anode degradation[15] showed no effect on performance at low ppm levels for Cd (30 parts per billion by volume [ppbv]); SbO (8 ppm); Pb, Co, V, Cr, and Be.

n. For V, Ca, Ba, Pb, Mn, and P in IGCC: Turbine deposition/corrosion should be removed by condensation in cleanup system; V should not be an issue due to low content in most coals, but could be an issue in pet-coke-fired coproduction systems.

o. For Ni and Fe in IGCC: Turbine deposition/corrosion possible issue—Fe carbonyls form at temperatures less than 400°F and Ni carbonyls form at temperatures less than 900°F.

The information on the four applications (Table 6-1) is based on a literature review, which includes information from lab-scale studies aimed at identifying and quantifying potential contamination issues. However, the ultimate test is prolonged commercial operation, where the full economic impact of the contaminant concentrations can be quantified. Many end-users consider actual practical experience under realistic commercial operating conditions as the true test of a syngas cleanup system's performance for producing a treated syngas, especially for chemical production applications. Emerging syngas cleanup technologies that have limited operational experience with actual coal-derived syngas are considered riskier than established commercial systems. Project developers need to consider this risk when selecting and designing the syngas cleanup system for modular plants as well as technical and economic factors impacting the design for the syngas cleanup system.

For a plant considering co-production of power and chemicals, one of the two options for the syngas cleanup is to have a bulk syngas cleanup system that removes all syngas contaminants required to meet power generation specifications and a polishing system that removes all remaining contaminants required to meet chemical production specifications. This approach adds equipment and process complexity associated with the integration of the bulk and polishing cleanup processes with any secondary treatment needed for their contaminant streams. A second option for power and chemical co-production plants is to design the syngas cleanup system so that all the syngas meets the chemical production requirements. The disadvantage of this option is that extra production cost is included in the syngas used for power production that cannot be recovered. Project developers considering a power and chemical co-production plant should evaluate the economic impact of these two options as part of the syngas cleanup design process.

6.2 DESIGN APPROACH FOR SYNGAS CLEANUP SYSTEM

To design the syngas cleanup system, a project developer needs to know (1) the desired temperature, pressure, and composition of the treated syngas, which specifies all concentration limits for any contaminant that adversely affects the syngas conversion process and (2) the temperature, pressure, and composition of the incoming raw syngas. The composition of the raw syngas contains details for all contaminants based on the specified fuel source selected. For the main syngas components (i.e., H_2, CO, water [H_2O], methane [CH_4], and carbon dioxide [CO_2]) and bulk contaminants (i.e., particulates, tars, sulfur, nitrogen species, and chlorine), there is a large body of information about the concentration of these species. However, less information is available for many of the trace contaminants. The measurement of these trace contaminants in the raw syngas is challenging for the following reasons:

- Extreme temperature and pressure conditions of the syngas leaving the gasifier.
  - Conditioning of the raw syngas to utilize existing analytical systems involves cooling the syngas sample, which can result in the condensation of some key trace elements, thereby invalidating their measurement.
- Extremely low concentrations of contaminant present, which challenge the capabilities of existing analytical systems.
- The only available analytical techniques for some trace contaminants are based on elemental analysis, whereas chemical species define which removal mechanisms/processes are effective for trace contaminant removal.
- The variability of feedstock composition.

The data typically collected throughout the syngas cleanup process focuses on the chemical species undergoing conversion and generally not contaminants that may be present. Consequently, limited information is available on the interaction of trace contaminants with many of the syngas processing systems, which creates a gap in the available knowledge of how much, where in the process, and how to effectively and selectively remove many of the trace contaminant species. This is the primary reason why only syngas cleanup systems that have been commercially proven under actual operating conditions are considered to have acceptable technical and economic risk for new chemical plants based on gasification.

With the raw and treated syngas compositions, design of the syngas cleanup system begins by identifying the specific processes that are necessary to convert the raw syngas into the treated syngas. As with any process, the removal and chemical transformation processes needed have specific requirements for operating temperature, pressure, chemical species, and other conditions that adversely affect the performance of their process. The challenge for designing the syngas cleanup system involves optimizing an assembly of individual processes to:

- Maximize the plant energy efficiency.
- Maximize the performance of each individual gas cleanup process.
- Minimize the size and number of required pieces of equipment.
- Minimize the generation of waste products of each individual gas cleanup process.
- Minimize the capital and operating costs for overall plant (including balance of plant systems).

6.3 SYNGAS CLEANUP PROCESS

To begin identifying the individual gas cleanup technologies required for the syngas cleanup system, Table 6-2 contains potential contaminants in raw syngas and the design considerations and constraints associated with these contaminants. The design process will be iterative, because there are multiple design configurations that technically work for any choice of gasification system and syngas conversion process. A process configuration for a syngas cleanup system will be developed that meets the technical requirements for the gasification and syngas conversion choices. This system configuration is evaluated against operability, maintenance and repair, and the project's economic objectives. The results from these evaluations inform the selection of a configuration that optimizes the technical requirements, operability, and economic objectives. The general information in Table 6-2 facilitates the evaluation of process configurations by describing criteria for removal and transformation processes that must be considered based on the choice of fuel, gasification system, and syngas conversion processes. The subsequent sections examine the individual removal processes in more detail.

Table 6-2. Design considerations for standard contaminants in syngas mixtures

CONTAMINANT	DESIGN CONSTRAINTS OR CONSIDERATIONS
Particulates	For gasifiers operating at temperatures high enough to melt the ash, any particulates carried out in the syngas will probably be molten. To effectively separate these particulates from the syngas, the temperature of the syngas must be below 900°F. At this temperature, all the refractory elements and alkali metals have condensed onto the particulates allowing for effective removal.
Refractory Elements (Mg, Ca, Ba, Pb, Ni, Fe, Sb, Mn, V, Cd, Cr, Be, and Co)	Refractory elements typically become a syngas contaminant for high-temperature, entrained gasification where the ash is molten. Most of these elements released from the molten ash condense on the surface of particulates when the syngas temperature drops to 900°F.
Alkali Metals (Li, Na, and K)	These elements are present as volatile species down to temperatures of 900°F. This has been a factor in establishing the target syngas temperature to maximize the removal of the species that will condense on the particulate solids, enabling their removal with the solid particulates. Biomass-derived syngas tends to have higher concentrations of alkali metals.
Halogens	The most prevalent halogen is chlorine. In the reducing environment in the gasifier, HCl is one of the thermodynamically stable chlorine compounds. HCl is extremely reactive and corrosive, leading to the destruction of downstream equipment and catalysts. Preferential removal of chlorine occurs as close to the gasifier as possible. The two main options are sorbents based on alkali metal carbonates or water quenching/scrubbing. Both options require lower syngas temperatures (to avoid alkali vaporization for sorbents and to permit the presence of liquid water), which hinders the effective recovery of the syngas' sensible heat. In addition, these removal options generate secondary waste stream treatment issues. For the sorbents, there is a disposal issue. Scrubbing increases downstream wastewater treatment requirements.
Hydrocarbons	As the gasification temperature increases, the effectiveness of the gasifier to break down hydrocarbons increases. For the gasifiers with the highest operating temperatures, the primary hydrocarbon present is CH_4 and its syngas concentration is usually low. However, as the operating temperature of the gasifier drops, the hydrocarbon concentration in the syngas increases. For gasifiers with the lowest operating temperatures that use lignite and biomass, the hydrocarbon concentration and diversity increase and result in a condensable mixture of tars. The heaviest of these tars have dewpoints of approximately 350°C (662°F), which constrains the treatment systems to operate with higher inlet syngas temperatures and increases the need for treatment as close to the gasifier as possible.
H_2-to-CO Ratio	Although this is not a contaminant, this ratio is important for chemical production applications. The primary factor affecting this ratio is the H-to-C ratio of the fuel used for the gasification process. The gasifier type and operating conditions further influence this ratio based on the operating temperatures and presence of steam (added or included as in slurry feed) in the gasifier. For chemical applications, suitable H_2-to-CO ratios are greater than or equal to 2.0, which requires use of the water-gas shift (WGS) reaction ($CO + H_2O \leftrightarrow CO_2 + H_2$) to achieve the target ratio. The exothermic equilibrium-limited WGS reaction requires a suitable catalyst, temperature, and steam concentration for optimal cost-effective operation.
Mercury (Hg)	In gasification, the reducing conditions and Hg's low saturation temperature ensure that Hg is almost exclusively present as elemental Hg vapor rather than in the form of mercury oxides as occurs during combustion. For gasification, carbon impregnated with sulfur has been demonstrated to effectively remove Hg from syngas at temperatures below 260°C (500°F).
Arsenic	Under the reducing conditions in the gasifier, thermodynamics favor the reduction of As to Arsine (AsH_3), which is a relatively stable compound. Arsine does not condense out with the other refractory elements on the particulates at higher temperatures. However, AsH_3 does poison many catalysts and does collect on metal surfaces when the syngas is cooled for solvent-based acid gas treatment. Special precautions are required to protect workers during the maintenance of exposed equipment.
Sulfur Species	As with the hydrocarbon contaminants, the operating temperature of the gasifier dictates what specific sulfur species are present in the syngas. At higher temperatures, most of the sulfur will be in the forms of hydrogen sulfide (H_2S) and carbonyl sulfide (COS), with H_2S making up 93 to 98% of the total sulfur. At lower gasification temperatures, where the reactivity is insufficient to breakdown hydrocarbon species, sulfur species found in the syngas include H_2S, COS, carbon disulfide (CS_2), and organic compounds like mercaptans and thiophenes, which complicates the sulfur removal process. Due to differences in the selectivity between H_2S and COS in most acid gas absorption processes, achieving acceptably low sulfur concentrations (less than 50 parts per million by volume [ppmv]) often requires hydrolysis to convert the COS into H_2S.
Nitrogen Species	The rule of thumb is that most of the nitrogen compounds are generated from nitrogen in the feedstock, rather than being thermally generated due to the high-temperature gasification conditions. For feedstocks with little or no nitrogen compounds, the concentration of nitrogen species in the syngas are almost exclusively generated by thermal reactions of nitrogen in the gasifier. The two predominant nitrogen species are hydrogen cyanide (HCN) and ammonia (NH_3). The HCN can be hydrolyzed to NH_3 by WGS catalysts and COS hydrolysis catalysts and the NH_3 subsequently removed by certain acid gas absorption systems. The NH_3 is removed based on its high solubility in water and solvents in acid gas absorption processes.
Carbon Dioxide (CO_2)	CO_2 is typically removed from the syngas for practical reasons for chemical production applications. These practical reasons include decreasing the volume of the syngas to enable use of smaller equipment and limit potentially inhibiting effects of the chemical conversion reactions. For methanol production, the optimal syngas composition is defined by the stoichiometric ratio ($([H_2] - [CO_2])/([CO] + [CO_2])$), with a value slightly greater than 2 and CO_2 concentration ranging from 2.5 to 3.5 mol%. For power production, CO_2 removal only becomes important when CO_2 capture is required. The current regulatory environment does not require CO_2 capture for power applications (except for hydrogen production for fuel cells). However, local siting issues can be restrictive. Additional information on CO_2 removal is provided in Chapter 12.

6.3.1 PARTICULATE REMOVAL

The target temperature for particulate removal is less than 900°F to ensure that the particulates are solids and the condensable refractory elements and alkali metals have condensed onto the available solid particulates (see Table 6-2). The three primary classes of particulates consist of fine slag, fine ash, and soot. The amount of carbon present in these particulates depends on the efficiency of the gasification process. Another factor to consider is leachability testing to determine if targeted elements can be leached out of the particulates at amounts exceeding regulated limits. Recycling some or all the particulate product may be practiced for increasing carbon conversion efficiency, providing a suitable chemistry for slag formation, and maximizing vitrification of refractory elements in the slag, which inhibits leaching from the particulate waste stream. Although the sale of the particulate product may be more beneficial, the specifications for the particulate product may be difficult to meet without additional processing, which could make this option economically unattractive.

The two basic processes for particulate removal are wet and dry (Table 6-3). The wet removal processes involve scrubbers and/or wash towers. The dry processes use either cyclones or filters. Based on the standard operating temperatures of these particulate removal processes, most of the refractory elements and alkali metals can also be removed by these processes. The wet removal processes can also result in the removal of other gas phase species (e.g., chlorides and other halogen compounds) due to their solubility in water. These wet processes are not effective for ammonia (NH_3) removal if wastewater saturated with NH_3 from prior use is reused. Higher pressure removal increases gas solubility, which means more intensive treatment is needed before the wastewater generated can be released to the environment.

Table 6-3. Particulate removal processes

PARTICULATE REMOVAL TECHNOLOGY	TYPE	OPERATING TEMPERATURE	DESCRIPTION
Cyclone	Dry	300 to 500°C[a]/ 572 to 932°F	The momentum of the particulates separates the particulates from the gas as the mixture travels in a circular path. This process is typically used as a bulk removal process. Filtration is used for final/polishing particulate control to meet specifications for most applications (power and chemicals).
Filtration	Dry	300 to 500°C[b]/ 572 to 932°F	Filtration is performed by either ceramic or sintered metal filter candles. Ceramic candles allow higher operating temperatures, but are more susceptible to failure due to thermal or mechanical shocks. The sintered metal elements are more mechanically rugged, but more susceptible to failure due to temperature and/or interaction with the syngas components. Continuous filtration is carried out with periodic back pulsing to remove the filter cake from the candles and collect it in a lock hopper. The gas used for back pulsing the filter needs to be compatible with syngas (preferably nitrogen or dry syngas) to avoid reactions or two-phase flow. Also, this back-pulse gas needs to be available at pressures 1.5 to 2 times the filter operating pressure.
Scrubbers/Wash Towers	Wet	Below Syngas Dewpoint (200 to 500°C/ 392 to 932°F)[c]	The temperature is fixed to ensure that the finest particulates serve as condensation sites for the condensing liquids, increasing removal efficiency. Due to the particulates being collected in water, leachable species like lead, zinc, cadmium, and others are collected in the aqueous phase, making subsequent wastewater treatment more complex and costly. For slurry-fed gasifiers, the collected fines can be recycled to slurry preparation, where they will end up in the slag. Being a wet process, this process allows for the removal of highly soluble gas species. This is particularly effective for chlorine removal based on the high solubility of chlorides. The wet process is also an effective way to increase the syngas steam concentration for any WGS requirements.

a. This temperature range is below alkali metal condensation points, but above that required for the formation of sticky species like ammonium chloride (NH_4Cl).
b. In addition to the conditions mentioned in note "a," this temperature is suitable for operation of ceramic and sintered metal candles.
c. This temperature range is set by the water dewpoint, which is a function of temperature and pressure. Higher pressure gasifiers enable higher temperature dewpoints.

6.3.2 HALOGEN REMOVAL

The most common halogen present in most fuels is chlorine. During gasification, the reducing environment promotes the conversion of the chlorine in the fuel into HCl. The metal surfaces of the process equipment have a strong tendency to react with HCl, resulting in corrosion. The HCl also reacts with syngas compounds, like NH_3, resulting in the formation of NH_4Cl, which results in equipment fouling as the NH_4Cl condenses out of the syngas. The HCl will also react with catalysts, like WGS catalysts, causing rapid deactivation. To prohibit and/or eliminate these issues, the removal of chlorine and other halogens from the syngas needs to occur as soon as possible after the raw syngas leaves the gasifier. Alkali carbonate sorbents use the reactivity of the halogen compounds to react with them and form solid alkali halides. To avoid the loss of the alkali compounds through vaporization, this sorbent-based adsorption process operates at temperatures below the dewpoint for alkali species (less than 900°F). Although the strong reaction between the alkali carbonate and halogen reduces the halogen species concentration in the syngas, the reaction is not reversible. When the alkali carbonate in the sorbent is consumed, the sorbent must be replaced. This requires shutting down the process or a multiple sorbent bed system with one sorbent bed in operation and another offline waiting to be brought into service. The use of sorbents for halogen removal also requires upstream particulate removal to prevent accumulation within and/or plugging of the sorbent bed.

Alternatively, a scrubber can be used for halogen removal. The scrubber can be used for particulate and halogen removal, which enables process intensification. This is one of the most common commercial contaminant removal processes for gasification, especially for chemical production. The use of water as a scrubbing agent ensures that the product gas is saturated with water vapor. This high concentration of water vapor in the effluent syngas provides all the steam needed to achieve the required H_2-to-CO ratio for most chemical production applications. This eliminates the need for additional equipment to generate high-pressure steam for the water-gas shift (WGS) process, but reduces the high-value sensible heat of the syngas. The reuse of wastewater for scrubbing concentrates the contaminants in the wastewater to be treated, reducing the volume of wastewater that needs treatment. However, a scrubbing process introduces higher water consumption for the plant and a more complicated wastewater treatment system capable of removing particulates, halogens, and other gas species. For large, commercial gasification plants, this process-intensive scrubber results in the lowest cost for bulk syngas cleanup.

6.3.3 HYDROCARBON REMOVAL

Gasifier selection defines gasification temperature and residence time, which influences the gasifier conversion chemistry and has a major impact on the concentration of hydrocarbons and tar in the raw syngas produced (see Chapters 2 and 3).

For high-temperature gasifiers, like entrained flow gasifiers, the primary reaction mechanisms are gasification and partial oxidation reactions. Under these conditions, the gasifier converts the larger hydrocarbons into H_2, CO, CO_2, and H_2O. The primary hydrocarbon present in the raw syngas from these high-temperature gasification processes is CH_4 (typically 1 vol% or less), with only trace concentrations of larger, more complex hydrocarbons. With these gasification systems, no additional hydrocarbon treatment is required. As the operating temperature in the gasification system drops, the hydrocarbon concentration in the raw syngas increases. Except for low-temperature gasification systems, which generate tars, the hydrocarbon concentration in the raw syngas is generally small enough that no dedicated hydrocarbon removal system is required. However, these trace hydrocarbon species collect in standard solvent-based acid gas systems. If the hydrocarbon concentration is high enough, pre-treatment of the syngas is required to avoid operational issues with the acid gas absorption process. Standard pretreatment options are available based on commercial applications used for natural gas processing.

For gasifiers operating at the lowest temperatures with biomass and lignite fuels (gasified at lower temperatures due to their higher reactivity), pyrolysis rather than oxidation reactions dominate, resulting in larger hydrocarbon/tar concentrations in the raw syngas. For these gasification systems, the tars need to be removed from the raw syngas to avoid equipment fouling and enable optimal syngas conversion downstream.

The removal of these tars from the raw syngas reduces the overall carbon conversion rate, so the decomposition of the tars into valuable syngas components, like H_2 or CO, is preferable. The three potential reaction mechanisms include partial oxidation, tar cracking, and steam reforming (more details about these three reaction mechanisms are provided in Table 6-4). The sensible heat of the raw syngas can be used for these reactions by integrating this process immediately after the gasifier. However, in this integration configuration, the decomposition process needs to handle the full load of syngas contaminants, including particulates, halogens, and sulfur.

Table 6-4. Potential tar decomposition reaction mechanisms

REACTION	DESCRIPTION
Partial Oxidation	This mechanism is most suitable as the last major reaction process to occur before cleaning the syngas. Oxidation catalysts could be used, but these would have to be tolerant of the particulate, halogen, and sulfur contaminants in the raw syngas. The preferred reactant would be oxygen (O_2). The exothermic nature of the oxidation reaction may require an inert diluent to be included with the O_2 to provide additional thermal mass to help control the reaction temperature. Although air could be used as the oxidant, this would result in the addition of nitrogen to the syngas. This additional nitrogen would be useful for temperature control of the exothermic oxidation reaction, but would require an additional removal process for the nitrogen. If air is not used, an ASU is needed to separate the O_2 from the nitrogen. If an inert diluent is required for exothermic temperature control, the preferred diluent would be CO_2, especially if CO_2 removal is already part of the syngas cleanup system.
Volatile/Tar Cracking	The cracking reactions typically result in the release of hydrogen-rich species and the deposition of carbon-rich species. These reactions are catalyzed by noble metals; nickel; and "natural catalysts," like dolomite, magnesite, calcite, and olivine. The volatile/tar cracking reactions are highly endothermic and require syngas temperatures greater than or equal to 750°C (1,382°F). The catalyst needs a periodic oxidation to regenerate the catalyst, as these reactions also result in the deposition of carbon-rich species on the catalyst.
Steam Reforming	In this process, the steam reforming reaction ($C_nH_{2n+2} + nH_2O \leftrightarrow nCO + [2n+1]H_2$) (for saturated hydrocarbons) is used to breakdown the tars. These steam reforming reactions are endothermic and consume a significant amount of steam. The reactions also require a catalyst. The most common catalyst is nickel, although it is easily poisoned by the sulfur compounds present in the syngas. Other noble metal catalysts can be used, but this increases the cost. Another catalyst deactivation mechanism is coking. The rate of coking is much lower than for tar cracking and can be reduced by the addition of steam; when combined with catalyst poisoning by syngas contaminants it results in the need to replace the catalyst, which should be included in the plant operational plans. High-temperature operating conditions support the endothermic steam reforming reactions, but additional heat energy needs to be provided to optimize conversion.

Alternatively, the tar can be removed from the syngas in an oil wash. The oil-based gas washer (OLGA) is an oil wash developed by the Energy Research Centre of the Netherlands (ECN) and Dahlman Industrial Group for tar treatment in biomass-derived syngas.[21, 22] The raw syngas feed enters the process at a temperature above the tar dewpoint (approximately 350°C [approximately 662°F]). A cyclone is used for bulk removal of entrained particulates. An initial absorption process is used to cool the syngas and capture heavier tars. These tars are recycled back to the gasifier for conversion. In a second absorption process, lighter tars are separated from the syngas as separate byproducts. These lighter tars are recovered using steam or air stripping of the washing oil. The OLGA process has been demonstrated with smaller scale gasification systems, including several integrated with the ECN MILENA indirect steam-blown gasifier, typically less than 12-megawatt-electric (MWe). Lab-scale pilot plant testing has demonstrated successful use of the OLGA process with F-T and substitute natural gas (SNG) synthesis.

6.3.4 HYDROGEN-TO-CARBON MONOXIDE RATIO ADJUSTMENT

The H_2 and CO in the syngas are not contaminants, rather they contain the chemical energy that makes the syngas valuable. For applications that use combustion to extract the chemical energy from these compounds as electric power, no adjustment of the H_2-to-CO ratio is required (though CO is preferred due to its higher molar heating value). The exception is electricity production with CO_2 capture, where the CO needs to be shifted into H_2 and CO_2 to enable capture of the CO_2 prior to combustion (more details on power production with CO_2 capture are provided in Chapter 12).

For applications where the H_2 and CO are used as reactants to produce chemical products, the H_2-to-CO ratio needs to be optimized for the targeted chemical conversion process. The optimum H_2-to-CO ratios for some of the main chemical production applications are provided in Table 6-5. As the H-to-C ratio in the chemical product increases, the H_2-to-CO ratio increases.

Table 6-5. H_2-to-CO ratios for production of different chemical products

CHEMICAL PRODUCT	H_2-TO-CO RATIO	CO CONVERSION
Methanol	2	~52%
Fischer-Tropsch Liquids	2	~52%
Methane	3	~64%
Hydrogen	~∞	~100%

The H_2-to-CO ratio is adjusted via the WGS reaction ($CO + H_2O \leftrightarrow H_2 + CO_2$). Although the gasification system impacts the composition of the syngas, the fuel provides the carbon and hydrogen for generation of the syngas. Consequently, the H_2-to-CO ratio is established based on the H-to-C ratio available in the fuel. For most coals, the H-to-C ratio is less than 1, so the H_2-to-CO ratio for syngas derived from coal is typically less than or equal to 1. An estimate for the CO conversion by the WGS reaction for coal-derived syngas mixtures to achieve the targeted H_2-to-CO ratio is also included in Table 6-5. For most hydrocarbon products with a H-to-C ratio of around 2, around 50 to 55% CO conversion is required to enable the syngas conversion chemistry. At these CO conversion levels, the typical WGS process would consist of a single WGS reactor with a syngas bypass loop (Figure 6-1) for controlling the H_2-to-CO ratio in the feed gas for the syngas conversion process.

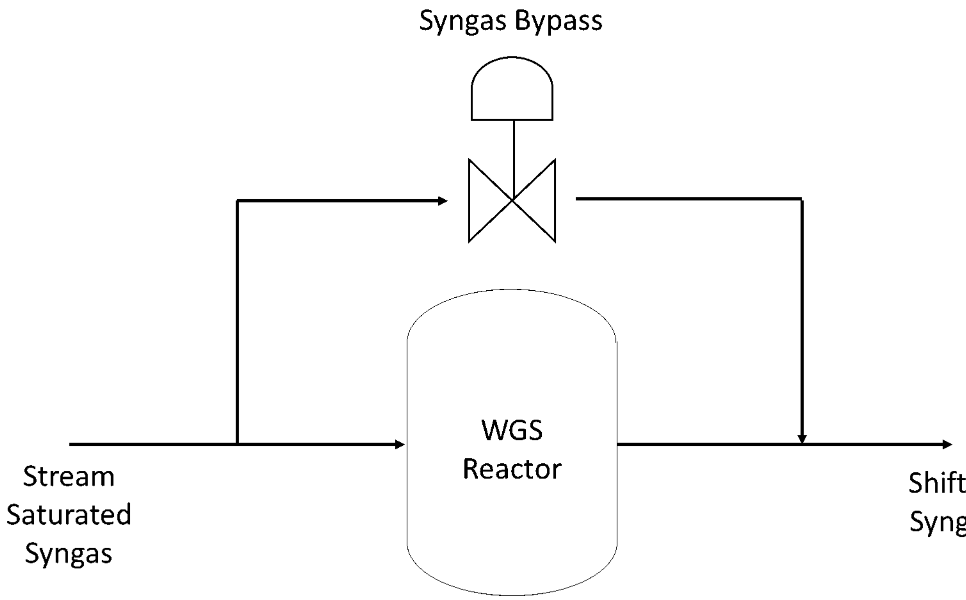


Figure 6-1. Single WGS reactor with syngas bypass

For a hydrogen product, which is the chemical product with no carbon, essentially all the CO must be shifted to CO_2 and then captured. Near complete CO conversion is also required for this application, because hydrogen is the target fuel for power production with high CO_2 capture (see Chapter 12). At this high level of CO conversion, the WGS process typically consists of multiple adiabatic reactors with inter-reactor cooling (Figure 6-2).

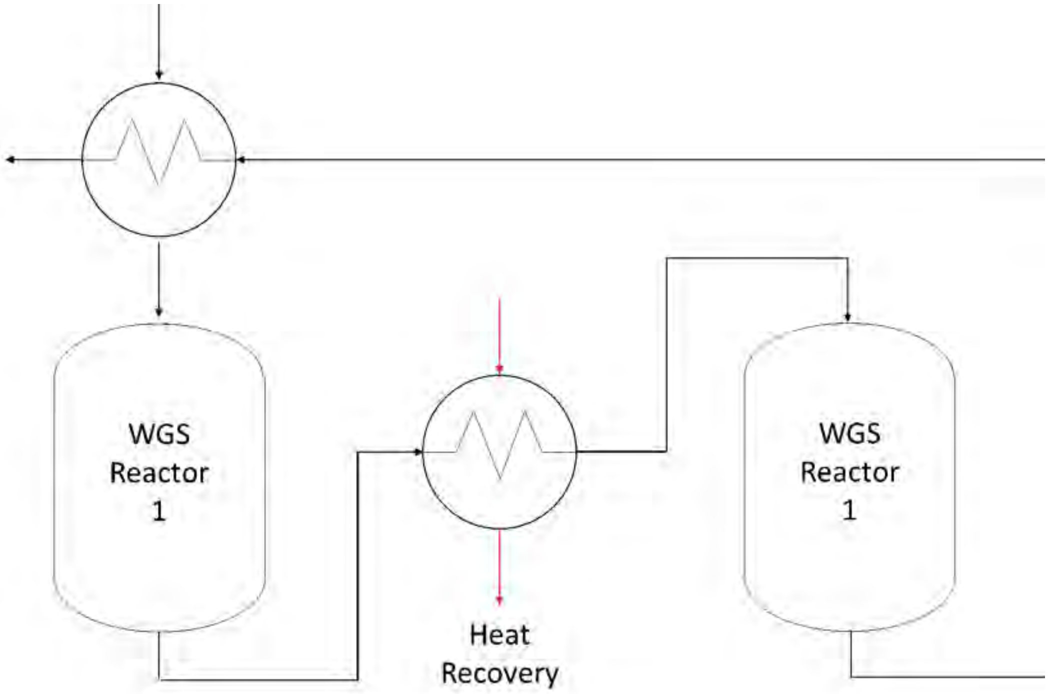


Figure 6-2. Multiple adiabatic WGS reactors with inter-reactor cooling

The forward WGS reaction is an exothermic (ΔHf = -41 kJ/mol), equilibrium-limited reaction. The WGS reaction equilibrium is essentially pressure insensitive, because there is no change in the number of reactant and product molecules. The optimal condition for maximum conversion is lower operating temperature, but the exothermic reaction heat increases the reaction temperature, reducing CO conversion under adiabatic conditions. The two process configurations to address this issue are multiple adiabatic reactors with inter-reactor cooling (Figure 6-2) or isothermal reactors where heat removal is integrated into the reactor itself. The isothermal reactor reduces the number of reactors required, but increases the complexity of the reactor internals. The multiple reactor process has several reactors, but their construction is simple. The simple reactor design also facilitates the loading/unloading of the catalyst from the reactors. The project developer should evaluate the process costs for both approaches to find the most cost-effective approach for the design of the modular plant.

Catalysts are needed to accelerate the reaction rates to enable reaching equilibrium at acceptable rates. There are three commercially available types of WGS catalyst, including high-temperature shift (HTS), low-temperature shift (LTS), and sour-gas shift (SGS). The difference between these catalysts is their sulfur tolerance. The LTS catalysts are the most sulfur sensitive, with a sulfur specification of less than 0.1 parts per million by volume (ppmv) in the feed gas. HTS catalysts can tolerate less than 100 ppmv of H_2S. For SGS catalysts, the active phase of the catalyst is a sulfided species. These catalysts require a high level of sulfur in the feed gas to maintain the sulfided active phase. Therefore, SGS catalysts are the only WGS catalysts that can be used directly with raw syngas. The HTS and LTS catalysts require some sulfur removal prior to the WGS reactor (i.e., “sweet-shift”). Typical solvent-based absorption processes for sulfur removal produce a treated syngas, which has a low temperature (less than 100°F) and minimal steam concentrations (low steam saturation at low syngas temperature). When sulfur removal occurs upstream of the WGS process, any potential process intensification using a single solvent-based absorption process for full simultaneous sulfur and CO_2 removal is lost, because additional CO_2 removal equipment is required downstream of WGS process (i.e., sweet-shift carbon capture and storage [CCS]).

The operating requirements for the different commercial WGS catalysts are provided in Table 6-6. The wider operating temperature range for the HTS and SGS catalysts reflects that these catalysts are typically used for bulk conversion in process streams with large CO concentrations and where high levels of CO conversion are needed. LTS catalysts are typically used as a final polishing step to achieve effluent CO concentrations of less than 1 mol% due to their high activity and favorable equilibrium at lower temperatures. When HTS and SGS catalysts are operated at high temperatures to achieve high CO conversions, high concentrations of steam in the syngas are required to minimize deactivation of the catalyst by coking. A rule of thumb for the steam-to-carbon ratio is 2, but some manufacturers recommend a steam-to-dry gas ratio of 2.

Table 6-6. Information on commercial WGS catalysts

CATALYST	ACTIVE COMPONENT	OPERATING TEMPERATURES	SULFUR TOLERANCE
HTS	Fe	600 to 900°F	≤100 parts per million
LTS	Cu and Zn	390 to 575°F	≤0.1 parts per million
SGS	Sulfided Co/Mo	400 to 900°F	≥250 parts per million

The sulfur tolerance of these catalysts is one of the defining operating requirements for these catalysts. Additional poisons for HTS and LTS catalysts include halogens and arsenic (As) compounds. The SGS catalyst is more tolerant of the contaminants in raw syngas, although the accumulation of As in the SGS catalyst can occur. Consequently, the unloading of spent SGS catalyst must include protection of plant personnel for exposure to the As contained in the spent catalyst for fuels with As concentrations. The SGS catalyst speeds up the hydrolysis of COS into H_2S and of hydrogen cyanide (HCN) into NH_3. HTS and LTS also catalyze the hydrogenation of HCN into NH_3.

In general, all the WGS catalysts are best suited for processing syngas mixtures with high steam concentrations. This steam serves several functions, including providing extra thermal mass to assist in limiting the temperature increase due to the reaction heat, extra reactant to drive the equilibrium toward high CO conversion, and catalyst protection against coke deposition.

The manufacturers' recommendation for steam-to-carbon ratios of 2 requires syngas steam concentrations of greater than 45 mol% for the CO-rich syngas generated from coal gasification. Based on typical syngas steam concentrations in the raw syngas, additional steam usually must be added to the syngas for the WGS process. The integration of a quench/scrubbing process for particulate removal immediately upstream of an SGS-based WGS uses the sensible heat of the raw syngas to generate the high steam concentration at the operating pressure while reducing the syngas temperature, particulates, and selected contaminant concentrations.

For syngas from a low-temperature, solvent-based absorption process for acid gas removal (AGR), water vapor saturation limits the steam concentration. The integration of a WGS process with this low temperature syngas requires preconditioning of the syngas to increase the syngas temperature and in turn raise the steam concentration to meet the WGS process requirements. Preconditioning increases equipment and parasitic energy load for the syngas cleanup system. These differences make integration of the SGS-based WGS process upstream of the sulfur and CO_2 removal processes the preferred commercial configuration for syngas cleanup for chemical production.

For chemical production applications that require less than 60% CO conversion, the standard WGS process configuration consists of a single adiabatic reactor with a syngas bypass loop for controlling the H_2-to-CO ratio of the syngas feed to the syngas conversion process. To produce hydrogen or other low carbon products (power with CO_2 capture) that need high CO conversion, the standard WGS configuration includes multiple adiabatic WGS reactors with inter-reactor cooling. Although an isothermal reactor would potentially reduce the number of reactors, adiabatic reactors generate treated syngas at higher temperatures, which contains higher value heat for recovery.

The project developer's selection and integration of a WGS process into the syngas cleanup system for H_2-to-CO ratio adjustment requires evaluation of technical and economic factors to obtain an effective H_2-to-CO ratio adjustment, as well as minimize the capital and operating cost (including parasitic energy losses) of the syngas cleanup system.

6.3.5 MERCURY REMOVAL

Under gasification conditions, the mercury (Hg) present is reduced to elemental Hg. This elemental Hg has low reactivity, which is demonstrated by little or no interaction with any materials at temperatures greater than 200°F. The release of Hg to the environment in flue gas during power production has become a concern, because of the toxic nature of mercury compounds. Mercury can be removed from syngas to concentrations below 0.1 $\mu g/m^3$ by adsorption in a sulfur-impregnated, activated carbon bed. The commercial operation of sulfur-impregnated, activated carbon at Eastman Chemical in Kingsport, Tennessee (USA) demonstrated in the range of 90 to 95% Hg removal during more than 30 years of operation.[23] Since these sulfur-impregnated, activated carbon beds are integrated upstream of the acid gas absorption processes, the Hg adsorption occurs at temperatures less than 200°F.

Mercury is also removed by the Rectisol® process for acid gas treatment. However, the Rectisol® process is typically not designed to recover the Hg, which accumulates as liquid Hg in cold sections of the process. As the Hg concentration in the raw syngas is in the ppb range, the accumulation of significant amounts of Hg occurs over extended periods. During periodic shutdowns of the Rectisol® plant used for the treatment of syngas, maintenance plans should include the evaluation and removal of any elemental Hg collected in the cold zone.

6.3.6 ARSENIC REMOVAL

There appears to be a general misconception in the scientific community that As is converted to arsine (AsH_3) during gasification. Arsine decomposes at temperatures of 230 to 300°C (446 to 572°F) and As sublimates at 614°C (approximately 1,137°F). Thus, it follows that arsenic mostly converts to elemental As vapor after gasification, although some arsine may be formed depending on the thermodynamic equilibrium between As vapor and As H_3 in the reducing environment. After the syngas is cooled, the arsenic vapors solidify and accumulate on the SGS catalyst; exposed surfaces in heat exchange equipment used for cooling the syngas; and the sulfur-impregnated, activated carbon used for Hg removal. Due to the toxicity of As, special maintenance procedures must be used when accessing surfaces or materials that are coated and/or trap As present in the syngas.

Arsine is a known poison for many low-temperature catalysts, and yet arsine poisoning of catalysts used for chemical applications with coal gasification have not been observed in practice, even though no specific arsine removal process was included in these plants. This suggests that arsenic solidifies and drops out on surfaces upon cooling before reaching the catalytic sites. The assumption is that As is separated as a secondary contaminant in the removal processes, such as the SGS, activated carbon-based Hg removal, or Rectisol® processes.

6.3.7 SULFUR SPECIES REMOVAL

The sulfur species found in the syngas depends on the gasification process and the fuel chosen. For lower temperature gasification processes with biomass pyrolysis reactions account for a large portion of the gasification reactions. As a result, organic compounds are only partially broken down by thermal decomposition. The resulting syngas contains significant concentrations of tars and organic sulfur species like CS_2, mercaptans, and thiophenes, in addition to H_2S and COS. As the temperature of the gasification process increases, gasification and oxidation reactions dominate and the main sulfur species in the syngas become H_2S and COS. For gasification conditions that produce only H_2S and COS, equilibrium conditions between H_2S and COS result in approximately 93 to 98% of the sulfur being H_2S, with the rest as COS.

Different sulfur removal technologies have different abilities to simultaneously remove H_2S and COS from syngas. For sulfur removal processes that are incapable of removing enough of the COS to meet target sulfur specifications, the COS must be converted into H_2S upstream of the sulfur removal process. This conversion of COS into H_2S can be performed by either a dedicated COS hydrolysis process or as a parallel hydrolysis reaction in the WGS process. These two processes are detailed in Table 6-7. The design decision about the need for COS conversion heavily depends on the sulfur removal technology selected and if the final syngas conversion process requires the use of WGS to adjust the H_2-to-CO ratio.

Table 6-7. COS conversion processes

PROCESS	TEMPERATURE (°F)	TECHNICAL DESCRIPTION
Hydrolysis Only	300 to 575°F	The COS is converted by a hydrolysis reaction ($COS + H_2O \leftrightarrow CO_2 + H_2S$). The optimal operating temperature range for the lowest COS concentrations in the treated syngas is 300 to 400°F. Acceptable catalysts include promoted chromium oxide-alumina, pure activated alumina, and titanium oxide. The active phase of these catalysts is a sulfide requiring use with syngas mixtures containing sulfur. These catalysts also tolerate heavy metals and arsenic. Halogens poison these catalysts. These catalysts promote the decomposition of nickel and iron carbonyls. To protect the catalyst bed from deposition of iron and nickel, a heated guard bed can be added. These catalysts also promote the hydrolysis of HCN into NH_3. To avoid introducing liquid water into the catalyst bed, which damages the catalyst, the syngas is superheated to 50°F above its dewpoint prior to entering the catalyst bed.
Simultaneous Hydrolysis and Water-Gas Shift	400 to 900°F	The only WGS catalysts suitable for this application are SGS catalysts, because their sulfided active phase allows operation in a syngas with high sulfur concentrations. The SGS catalyst can also convert other more complex organic sulfur compounds into H_2S.

Most commercially available sulfur removal technologies are adaptations of technologies developed for removing acid gases, which include H_2S and CO_2, from raw natural gas and other process gases. Simultaneous removal can be a benefit when the removal of H_2S and CO_2 is required to meet the treated syngas specifications or a disadvantage when the CO_2 (its mass flow rate) is an integral component of performance, as in power output from the gas turbine in IGCC. For project developers, the first selection criterion for a sulfur removal technology is its selectivity for sulfur. The second criterion is its ability to also meet CO_2 removal requirements.

Another selection criterion is the sulfur removal technology's ability to produce a suitable sulfur byproduct stream that is suitable for disposal or sale either directly or with minimal additional processing requirements. Most sulfur removal technologies generate a sulfur byproduct stream containing H_2S and COS. This sulfur byproduct stream is then typically treated in a Claus sulfur treatment system to convert the H_2S and COS into elemental sulfur, but can also be oxidized and then converted into sulfuric acid. A requirement for a standard Claus sulfur treatment system is a H_2S concentration of greater than 30 mol%. This requirement can be reduced for an oxygen-blown Claus sulfur removal system depending on the availability of surplus oxygen from an ASU. The tradeoff between accepting a sulfur removal technology that can only produce lower concentration sulfur byproducts and increasing the capacity of an existing ASU to allow the use of an oxygen-blown Claus sulfur removal system is contingent on the economic benefits.

6.3.7.1 ABSORPTION TECHNOLOGIES

The largest category of acid gas treatment options is based on absorption technologies. In these absorption technologies, the acid gas is absorbed into a liquid solvent. After absorption, the "rich" liquid solvent is regenerated by flashing, stripping, and/or thermal regeneration. The "lean" solvent is then recycled to the absorption column. A simplified process flow diagram for this absorption process is shown in Figure 6-3. Acid gas capacity is based on the chemical bonding between the acid gases and solvent for chemical solvents and physical solubility for physical solvents. The acid gas capacity of physical solvents can be increased by operating conditions, including lower temperatures and higher pressures. As a result, process economics favor physical solvents in higher pressure gasification systems.

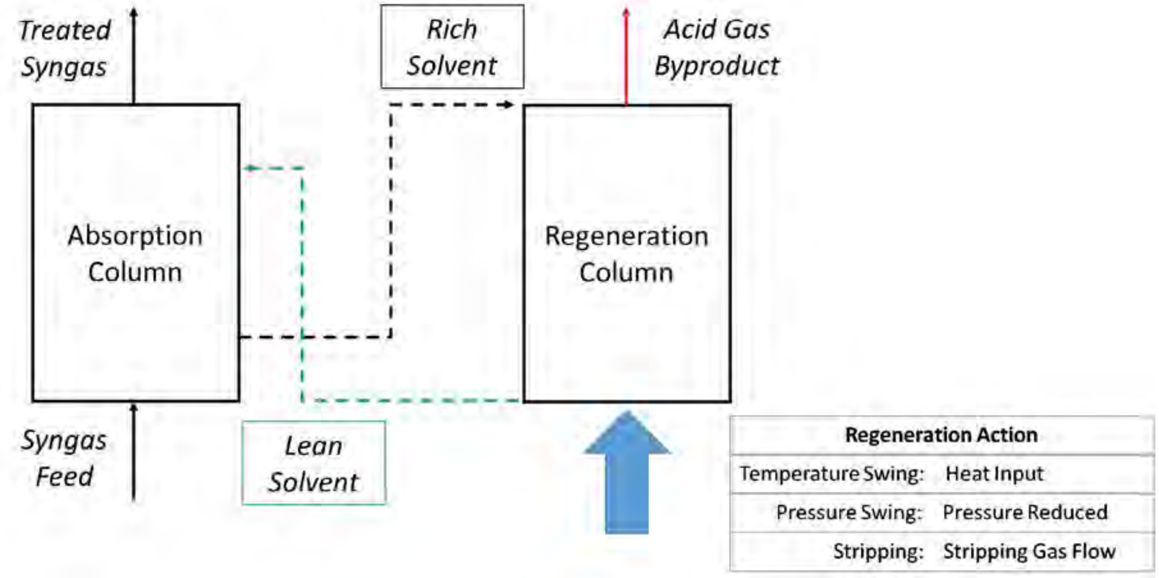


Figure 6-3. Process flow diagram for acid gas absorption process

A noteworthy operating feature of these absorption systems is that they operate at temperatures below 250°F to enable operation with the various liquid (typically aqueous) solvent solutions. These low temperature operating conditions create integration issues. For power and chemical production applications, the sulfur must be removed upstream of the syngas conversion processes, but these syngas conversion processes operate at much higher temperatures than the absorption processes. This results in the need for additional heat exchange equipment to recover the heat in the syngas prior to the absorption. This configuration increases the need for process equipment and thermal inefficiencies of the plant. These factors impact process economics; therefore, the project developer's evaluation of the process economics is critical during the selection of a sulfur removal technology for the syngas cleanup system.

Other factors that impact the absorption process economics include:

- Availability of the solvent in quantities large enough and at a price to meet market needs.
- Low volatility of the solvent to minimize loss during thermal regeneration.
- Low viscosity to minimize power required for pumping the solvent through the system.
- Low regeneration energy for regenerating the solvent.
- Low toxicity to minimize hazards for operating plant personnel.

Due to the maturity of the absorption-based, AGR technologies, a variety of available commercial solvents can be used to optimize the technical performance and process economics for modular gasification applications.

6.3.7.1.1 CHEMICAL SOLVENTS FOR ABSORPTION TECHNOLOGIES

The standard chemical functionality of a chemical solvent is provided by an amine functional group. The amine functional group undergoes an acid/base reaction with H_2S and CO_2, separating the acid gases from the process gas. Methyldiethanolamine (MDEA) is the most widely used solvent for coal gasification applications due to its higher selectivity for H_2S solvent bonding compared to primary amines like monoethanolamine (MEA) and secondary amines like diethanolamine (DEA). Proprietary formulations are available to address corrosion issues, selectivity for H_2S and CO_2, and high-energy requirements for regeneration.

Hot aqueous solutions of potassium carbonate have also been used as chemical solvents for acid gases. The challenge with using hot potassium carbonate for treating some syngas is achieving enough sulfur removal to meet the target sulfur concentrations in the treated syngas. Hot potassium carbonate solutions have been successfully used for pure CO_2 removal from sulfur-free syngas generated in steam reformers. However, amine-based absorption processes are more cost competitive for this application.

For simple applications where the primary objective is AGR from the syngas, absorption processes based on chemical solvents are an option for syngas cleanup. However, when there is a need to treat H_2S and CO_2 differently for either removal from the syngas or the byproducts generated, chemical solvents face technical challenges. The differences in selectivity for H_2S and CO_2 can be used to achieve preferential absorption of either H_2S or CO_2, but some of both species are removed. For IGCC power applications, the removal of CO_2 reduces mass throughput in the gas turbine, reducing power output. During the typical temperature swing regeneration process used for chemical solvents, H_2S and CO_2 are simultaneously released. This regeneration exclusively generates a byproduct stream with relatively fixed concentrations of H_2S and CO_2. Enriching the byproduct stream with H_2S to meet specifications for the Claus sulfur removal system is technically challenging. Similarly, generating a separate CO_2 byproduct stream with low enough sulfur concentrations for environmental release or to meet CO_2 utilization or storage applications also has challenges. These challenges have limited the use of chemical solvents for coal gasification applications.

6.3.7.1.2 PHYSICAL SOLVENTS FOR ABSORPTION TECHNOLOGIES

Commercial physical solvents have been developed for high solubility of the acid gases. Due to the differences in H_2S and CO_2 concentrations in syngas, commercial physical solvents also are more selective for H_2S over CO_2. The three main commercial physical solvents used for AGR are refrigerated methanol (Rectisol®), dimethyl ethers of polyethylene glycols (Selexol™, DMPEG), and n-methyl-pyrrolidone (Purisol, NMP). These absorption processes are detailed in Table 6-8. The difference in H_2S and CO_2 selectivity of these physical solvents provides flexibility in process configuration for non-selective and selective removal of the H_2S and CO_2 from the syngas into separate sulfur and CO_2 byproduct streams. Stringent specifications for the treated syngas and the byproduct streams create a more complex process configuration.

Table 6-8. Commercial physical solvents for acid gas removal

PROCESS	RECTISOL®	SELEXOL™	PURISOL
Solvent	Methanol	Dimethyl ethers of polyethylene glycol (DMPEG)	N-methyl-pyrrolidine (NMP)
Operating Temperature Range	-22 to -40°F	32 to 105°F	60 to 105°F
Relative H_2S solubility compared to CO_2 (CO_2= 1) at 25°C (77°F)	7.6[24,a]	8.93[25]	10.2[25]
Primary Commercial Syngas Application	Chemical production	Power production	Power production
Other Advantages	Process can achieve low concentrations of H_2S and COS (less than 0.1 ppmv) and CO_2 (less than 10 ppmv). Process also removes HCN, NH_3, hydrocarbons, Hg, and iron and nickel carbonyls. The treated syngas has been repeatedly demonstrated for chemical production from gasification systems, making it the standard application. The disadvantage for this process is the need for refrigeration to cool the syngas.	The higher operating temperature of this process reduces the thermal inefficiencies associated with cooling the syngas during syngas treatment. The limit for sulfur removal of about 1 ppmv limits this process to power applications. This technology can meet current and potential future regulatory CO_2 emission limits, because it can be configured to generate distinct H_2S and CO_2 byproduct streams.	This system is like Selexol™ in both performance and operations. It has the highest selectivity toward H_2S among all commercial solvents.

a. Data measured at -25°C (-13°F) to accommodate the higher vapor pressure of methanol.

The regeneration options for physical solvents include pressure swing and stripping, as well as temperature swing processes, because the removal mechanism is the solubility of the acid gases in the solvent (based on weaker Van der Waals interactions). This contributes to the flexibility in process configurations for physical solvent-based absorption processes. The optimization of the absorption process is achieved by maximizing solubility, which favors low-temperature and high-pressure operating conditions. These favorable operating conditions make physical solvent-based absorption processes most technically and economically effective for high-pressure gasification systems.

6.3.7.1.3 OTHER SOLVENTS FOR ABSORPTION TECHNOLOGIES

To simultaneously achieve the benefits of chemical and physical solvents, absorption processes using mixtures of chemical and physical solvents are available. Some examples of commercial mixed solvent absorption processes include Shell's Sulfinol and Lurgi's Amisol. The primary objective for mixed solvents is to increase acid gas loading to reduce solvent flow, which is proportional to the process costs. This combination offers improved hydrolysis of COS due to the interaction with the physical solvent. Although the performance of these mixed solvents is estimated based on the characteristics and properties of the solvents in the mixture, the secondary removal of contaminants like HCN, NH_3, and carbonyls is difficult to predict. The limited commercial experience available with these mixed solvents for syngas has inhibited confidence in these systems and their ability to achieve the same level of syngas cleanup provided by Rectisol® for chemical production applications. Mixed solvents may be advantageous for modular systems due to the potential for creating a solvent mixture that has better removal performance for multiple contaminants like HCN, NH_3, carbonyls, arsenic, and Hg. This would enable process intensification for syngas cleanup, which would reduce process equipment in the syngas cleanup system and improve modular plant economics without sacrificing removal efficiency.

There are also solution-based absorption processes where components in the solvent react with the sulfur species in the syngas and convert them into elemental sulfur. This chemical transformation of the sulfur species into elemental sulfur differentiates them from typical chemical solvents that bond with the sulfur species, but release the same sulfur species upon regeneration. In the absorption column, the active component in the solution reacts with the sulfur species in the syngas to oxidize them to elemental sulfur, which precipitates from the solution. The active components for this reaction are highly oxidized metal ions. In the early stages of the development of these processes, vanadium-based, active components were used; however, environmental and toxicity issues associated with these vanadium-based components made their use unacceptable. Current commercial processes like Lo-Cat, SULFCAT, and Sulferox use chelated or encapsulated iron species as active components. The reaction for these iron active components with the sulfur species in the absorption column is: $HS^- + 2Fe^{+3} \rightarrow S^0 + 2Fe^{+2} + H^+$.

The spent solution leaves the absorption column loaded with precipitated elemental sulfur and ferrous ions (Fe^{+2}) and is transferred to an oxidizer vessel. In the oxidizer, oxygen from air is used to re-oxidize the ferrous ions into ferric ions (Fe^{+3}). The precipitated sulfur settles by gravity to the bottom of the oxidation vessel. A filtration system is used to separate the elemental sulfur from the solution. The re-oxidized solution is then recycled back to the absorption column.

The raw sulfur byproduct generated from this process typically contains 65 to 85% sulfur, with water and dissolved salts making up the balance. The additional purification of this sulfur byproduct is necessary for the "bright yellow" color specification for commodity sulfur. Alternatively, the disposal of this sulfur byproduct requires additional water removal to meet suitable standards.

Based on the oxidation reaction to elemental sulfur, this absorption process is specific for the removal of H_2S and ineffective for the removal of COS or CO_2. This process is effective for removing H_2S from CO_2 and for process gas mixtures with low H_2S partial pressures. Commercial applications of this process are generally for smaller-scale applications. This process is used as a substitute for a Claus sulfur treatment system in situations where the sulfur concentration is too low. It has also been used for tail gas treatment in Claus plants.

6.3.7.2 ADSORPTION TECHNOLOGIES

In the chemical industry, the adsorption of contaminants onto sorbents is the standard approach to protect catalyst beds from poisons present in the process gas. For sulfur species, one of the key sorbents used is based on zinc oxide (ZnO). Zinc oxide reacts with the H_2S in the process gas to form zinc sulfide (ZnS). At typical operating conditions, the equilibrium concentrations favor treated gas compositions with less than 0.1 ppmv of H_2S. Although ZnO is selective for H_2S, its performance in the removal of COS and mercaptans is poor. Consequently, COS and mercaptans need to be converted to H_2S for effective treatment with ZnO. Based on their development for guard bed applications, commercial ZnO-based sorbents are designed for a single use with a sulfur capacity of approximately 20% by weight.

Due to this design, commercial ZnO-based sorbents are exclusively used for polishing and/or guard bed applications where the H_2S concentration in the process gas is less than or equal to 1 ppmv. Under these conditions, sorbent beds of acceptable design size can last for one to two years on a single sorbent charge. For the high-sulfur concentrations (greater than 1,000 ppmv) in syngas, the replacement cost of commercial ZnO-based sorbents is too high for competitive operation.

For most catalytic applications, the sulfur concentration after treatment with ZnO-based sorbents is low enough to meet the sulfur specification for most catalysts. For the more sulfur-sensitive catalysts, commercial copper oxide (CuO)-based sorbents can achieve sulfur concentrations of less than 50 parts per billion by volume (ppbv) in the treated process gas. For most syngas conversion processes where the catalyst is rapidly poisoned by an upset that allows sulfur into the treated syngas, a ZnO- and/or CuO-based sorbent bed is included upstream of the syngas conversion process to eliminate the potential for accidentally poisoning the catalyst bed. The size of this guard bed would be determined by the risk of an upset and the amount of sulfur that could reach the syngas conversion catalyst bed. For Rectisol®, the available commercial operating experience would support a smaller guard bed than a sulfur removal process with only pilot plant testing data.

The disadvantages of existing commercial ZnO-based sorbents are their fixed sulfur capacity and that they are single use, which limits their application to polishing and/or guard bed service. With regenerable ZnO-based sorbents, a continuous removal process for sulfur in syngas is possible using multiple fixed-beds of sorbent. Staggered batch operation of these sorbent beds enables at least one bed to be in adsorption mode. The other sorbent beds would be in different stages of purging, heating, and regeneration.

Two regenerable ZnO-based sorbents have been developed. One of these sorbents, RVS-1, was developed through in-house research and development (R&D) by the U.S. Department of Energy's National Energy Technology Laboratory (DOE/NETL). This sorbent has been extensively tested at lab-scale in a variety of syngas compositions, temperature, pressures, and regenerations conditions with the sorbent showing stable adsorption/regeneration performance in cyclic testing of adsorption and regeneration for up to 50 cycles.[26,27] The sulfur concentration in the treated syngas from this sorbent is less than 10 ppmv prior to breakthrough, which makes this sorbent well suited for power generation applications and, with an additional ZnO guard bed, suitable for chemical applications.

A second regenerable ZnO-based sorbent was developed by RTI International for their "Warm Gas Desulfurization Process (WDP)." Rather than a fixed-bed sorbent, RTI's sorbent was a fluidizable sorbent for use in a transport reactor system. By moving the sorbent through the adsorber and regenerator reactors, RTI's transport reactor system consists of a single adsorber and a single regenerator. RTI successfully demonstrated the performance of the transport reactor system and fluidizable regenerable ZnO sorbent in a 50-MW demonstration plant (Figure 6-4) with high-sulfur syngas generated from the gasification of a coal/petroleum coke mixture at the Tampa Electric IGCC plant in Polk County, Florida (USA). During this demonstration testing, the sorbent demonstrated simultaneous removal of H_2S and COS, reducing the sulfur concentration in the syngas (initially around 10,000 ppmv H_2S and around 600 ppmv COS) by more than 99.9%.[28]

Used with permission from Howe et al.[6]

Figure 6-4. 50-MW RTI syngas cleanup demonstration plant at Tampa Electric IGCC plant in Polk County, Florida

The regeneration of these ZnO-based sorbents is based on the oxidation reaction of $ZnS + 1.5\ O_2 \rightarrow ZnO + SO_2$. The enthalpy for this exothermic reaction is 300 kJ/mol. Controlling the reactor temperature during this reaction is critical to avoiding sintering and weakening of the sorbent support structure, as well as effective regeneration of the active ZnO phase. For the fixed-bed sorbents, the oxygen concentration in the regeneration gas is limited to less than 5 mol%. For the transport reactor, the added thermal mass of the sorbent enables the use of air while allowing reactor temperature control. In contrast to solvent-based absorption processes, the sulfur species in the sulfur byproduct stream is SO_2, rather than H_2S. Although modified versions of the Claus sulfur removal system can be used, this requires a slipstream of the syngas to reduce the SO_2 to H_2S to support the reaction chemistry. Alternatively, the SO_2 can be further oxidized into SO_3 to produce sulfuric acid.

Although these regenerable sorbent-based sulfur removal processes are not fully commercial, they have been extensively tested at a smaller scale that may be more compatible with modular gasification systems. RTI's system analysis also shows that these dedicated sulfur removal systems have better performance and process economics compared to commercial Rectisol® and Selexol™ systems for IGCC applications (with 90% CO_2 capture) and methanol and F-T synthesis (FTS).[29,30] These benefits occur because the dedicated sulfur and CO_2 removal systems are selective and simple compared to the combined sulfur and CO_2 removal in Rectisol® and/or Selexol™, which are further complicated by the need to separate the sulfur and CO_2 into separate byproduct streams.

6.3.8 NITROGEN SPECIES REMOVAL

The primary nitrogen compounds found in syngas are HCN and NH_3. The two sources for the HCN and NH_3 found in the syngas are the decomposition of organic nitrogen compounds present in the fuel source for gasification and the molecular nitrogen introduced with the oxidant and fuel for gasification. Most of the HCN and NH_3 present in the syngas is produced by decomposition of the organic nitrogen in the fuel, because the nitrogen bonds in the organic nitrogen compounds in the fuel consist of N-H or N-C bonds and are weaker than the N-N triple bond in molecular nitrogen. Minor amounts of HCN and NH_3 formed from molecular nitrogen are caused by the highly favorable reaction thermodynamics from the high-temperature conditions in the gasifier.

During combustion of the syngas for power generation, the HCN and NH_3 result in the formation of NO_X, which is a regulated contaminant for power plants. For methanol production, NH_3 present in the feed gas can be catalytically converted to amines, which are not permitted in commercial methanol specifications. HCN is also a poison for F-T catalysts.[31] HCN is effectively removed in amine-based absorption systems, but will degrade the amine by reacting with it, because of its solubility in water. The high hydrogen and steam content in syngas enables the catalytic conversion of HCN into NH_3 during any WGS or COS hydrolysis processes.

The removal of HCN and NH_3 from the syngas depends on the high solubility of HCN and NH_3 in water. This removal is typically not observed during quenching/scrubbing of the raw syngas, where recycled wastewater, which is saturated with these compounds, is used. During syngas cooling for the AGR process, a large portion of the steam content of the syngas is condensed. This condensed process water will absorb a significant amount of the HCN and NH_3 present in the syngas. Further treatment in an aqueous amine AGR process will separate additional HCN and NH_3. If the HCN concentration entering the amine process is too high, a pretreating column with water to remove the HCN should be considered to reduce the operation costs associated with replacing amine degraded by the reaction with HCN. The two most popular commercial physical solvents for gasification applications, Rectisol® and Selexol™, are also effective for the removal of HCN and NH_3.

6.3.9 CARBON DIOXIDE REMOVAL

For power applications, pre-combustion CO_2 removal reduces the net mass through and the power generated in the gas turbines in IGCC systems, which lowers plant thermal efficiency and increases power production costs. This approach is considered preferable to post-combustion CO_2 removal, which tends to have large electrical load requirements due to the relatively low pressure of the exhaust gases. These parasitic losses are many times greater than any power losses associated with pre-combustion capture. For chemical production applications, currently, CO_2 removal is performed to achieve more favorable thermodynamic conditions or avoid inhibiting effects associated with CO_2. Most current commercial CO_2 removal processes result in the release of the CO_2 to the environment at near ambient conditions. In this section, the discussion of CO_2 removal technologies focuses on the needs of current commercial chemical production applications.

For chemical production, the H_2 and CO concentrations in the syngas must be adjusted to support the chemistry of the syngas conversion process used to make targeted chemical products. The modification of the H_2 and CO concentrations is accomplished via the WGS reaction (see Section 6.3.4). Commercial gasification-based chemical production plants integrate the WGS process upstream of the sulfur removal process to take advantage of higher syngas temperatures for the WGS reaction, high syngas steam concentrations to force more CO conversion and reduce catalyst deactivation, higher sulfur removal efficiency in the downstream sulfur removal process due to COS hydrolysis catalyzed by the WGS catalysts, and simultaneous sulfur and CO_2 removal using commercial AGR technologies developed for natural gas processing.

A primary constraint for this AGR technology for CO_2 is the ability to produce treated syngas and sulfur byproduct streams with suitable specifications. The specifications for the sulfur byproduct stream are set by the Claus sulfur treatment process, which requires the AGR process to separate the sulfur from the CO_2 to generate a sulfur byproduct with greater than 30 mol% sulfur and a CO_2 byproduct that meets sulfur oxide (SO_X) emission requirements.

For chemical solvent-based acid gas absorption processes (e.g., amines), the chemical bonding that removes the H_2S and CO_2 typically requires thermal regeneration using temperature swing. As a result of the similarity of the chemical bonding between the solvent and both H_2S and CO_2, the conditions during thermal regeneration favor simultaneous release of H_2S and CO_2 during regeneration, inhibiting the possibility of generating separate high-purity H_2S and CO_2 byproduct streams without adding equipment and system complexity. Amine AGR processes are commercially used for syngas cleanup for chemical production in applications that do not require simultaneous sulfur and CO_2 removal.

The differences in the physical solubility of sulfur species and CO_2 in physical solvents allow these processes to generate a sulfur byproduct suitable for downstream Claus sulfur treatment and a CO_2 byproduct with sulfur concentrations that meet end-use specifications. The two physical solvent absorption-based AGR processes that have been successfully used in commercial gasification-based plants are Rectisol® and Selexol™. For chemical applications, Rectisol® is the standard process selected based on its ability to reduce sulfur compounds in the treated syngas to concentrations suitable for even the most sulfur intolerant syngas conversion process and its ability to remove trace contaminants like mercury (Hg), NH_3, and HCN. Although the standard Selexol™ process could be modified to achieve sulfur removal levels that meet chemical production specifications, the increase in operational complexity and operational costs are similar or higher than that for Rectisol®. This aspect, and the commercial success of Rectisol® for chemical production, has positioned Rectisol® as the standard for AGR of syngas for chemical production applications. However, the Rectisol® process requires that the syngas is reduced to near cryogenic temperatures and has higher auxiliary loads, and there is limited recoverable exergy to offset thermal efficiency losses for power applications. Thus, for power applications, Selexol™ is the more preferred technology.

Other potential technologies that could be used for CO_2 removal from syngas (e.g., adsorption, membranes) are not able to both effectively and simultaneously remove sulfur and CO_2 from the syngas and export these compounds into two separate, high-purity streams that meet end-use specifications. These alternative gas cleanup processes also lack commercial applications to effectively compete with the established Rectisol® or Selexol™ technologies. Although separate processes for sulfur and CO_2 removal could be used to produce suitable sulfur and CO_2 byproduct streams that could meet such specifications, the possibility of multiple processes having an economic advantage in capital and operating costs over a single process is unlikely.

However, RTI International has demonstrated that their proprietary desulfurization technology based on a regenerable sulfur sorbent coupled with a standard activated MDEA CO_2 removal process provides better economic value than either Selexol™ or Rectisol® for power and chemical production applications.[30] The separate individual processes are advantageous, because the high selectivity of the individual processes enable compact, simple removal configurations, whereas a single integrated process requires complex configurations to provide the sulfur and CO_2 selectivity that are needed to meet the specifications of the sulfur and CO_2 byproduct streams. This advantage is significant when CO_2 emission regulations are enacted and more stringent specifications are required for both the sulfur (with CO_2 being a key contaminant) and CO_2 byproduct streams. With only a large-scale demonstration of RTI's desulfurization and commercial amine technologies for separate sulfur and CO_2 removal processes at Tampa Electric Company's IGCC plant in Polk County, Florida (USA), there is insufficient commercial operating experience to compete with the commercial experience of Rectisol® or Selexol™. As separate sulfur and CO_2 removal processes may be more suitable for modular applications, this option should be considered by the project developer as part of the syngas cleanup design, recognizing the inherently higher risk associated with non-commercial technologies. Other sulfur removal technologies may be appropriate for modular gasification facilities based upon the amount of H_2S that has to be treated and options to deal with the sulfur-based product.

6.3.10 OTHER SPECIES REMOVAL

The contaminants discussed in the previous sections are present in the original raw syngas. During the syngas cleanup process, several additional contaminants are introduced into the syngas with chemicals used to treat the syngas (accidentally) or by the reaction between syngas components. Oxygen is one of the compounds that is introduced into the syngas accidentally, usually as a dissolved component in water used during scrubbing or promoting WGS reactions. Oxygen can also be introduced by air leaking into plants being operated near or below atmospheric pressure. This oxygen results in the oxidation of the WGS catalyst and other syngas components and could lead to explosive mixtures of syngas and oxygen. To avoid this problem, only deaerated water should be added into the process.

The high CO concentrations in the syngas enables the formation of nickel and iron carbonyls at lower operating temperatures. Carbonyls species have low solubility in water, so they tend to pass through scrubbers and most acid gas treatment systems. At higher temperatures, they decompose and deposit nickel and iron. This deposition results in increased pressure drop, undesired side reactions catalyzed by the iron and/or nickel, and the fouling of heat exchangers and turbine blades. These carbonyl compounds are also extremely toxic. Although carbonyls have a low solubility in water, they are more soluble in organic solvents. Acid gas absorption processes that use physical solvents, like Rectisol® and Selexol™, remove carbonyl species. Heated guard beds have also been used to decompose the carbonyls and capture the deposited iron or nickel to protect catalyst beds for COS hydrolysis. This heated guard bed approach could be used for the protection of other catalyst beds, heat exchangers, and turbines from iron and nickel carbonyls.

For high-pressure syngas at lower operating temperatures, thermodynamic conditions are favorable for the formation of formic acid from CO and H_2O via the reaction $CO + H_2O \leftrightarrow HCOOH$. In addition, formic acid may also be formed during COS hydrolysis. Formic acid formation can cause problems in the form of acidic corrosion of equipment and adverse solvent performance due to high formate concentrations in amine-based, acid gas absorption processes. Although the thermodynamically favorable conditions for formic acid formation may be present, its formation has typically not been observed in commercial gasification plants. The most probable reason is that other species in the syngas (e.g., NH_3 and chlorides) that are absorbed into the water buffer the pH of the water by reacting with any formic acid present, creating unfavorable pH conditions for formic acid formation.

6.4 INTEGRATION OF GAS CLEANUP PROCESSES FOR SYNGAS CLEANUP SYSTEM

The variety of different contaminants present in fuels mandates the use of multiple gas cleanup processes to remove these contaminants from syngas mixtures for either power or chemical production. Process operating requirements, including temperature, pressure, and concentrations of poisons in the syngas, for the individual gas cleanup processes introduce integration constraints. The goal for design of the syngas cleanup system is to integrate these multiple individual gas cleanup processes based on their individual integration constraints to meet technical and economic objectives. With this design goal, an iterative approach allows the evaluation of different integration configurations to select the optimized integration configuration and different gas cleanup processes for specific contaminants.

6.4.1 COMMERCIALLY DEMONSTRATED GAS CLEANUP PROCESSES

The methodology for creating integrated configurations focuses on sequential removal of the contaminants, starting with contaminants with the highest adverse impact on downstream operations and working toward the contaminant with the least adverse downstream impact while maximizing the thermal efficiency of the overall plant. This means avoiding the introduction of low-temperature removal processes upstream of processes that operate at higher temperatures to avoid the loss of thermal efficiency and the cost of additional equipment to reheat the syngas. Based on this methodology, the typical removal sequence for contaminants is listed in Table 6-9. The integration sequence is based on standard commercially available systems with actual operating experience with coal-derived syngas (Table 6-9).

Table 6-9. Typical integration sequence for contaminants

CONTAMINANT	DESCRIPTION
Particulates	By starting with the removal of particulates, the syngas temperature is reduced to condense out most of the refractory elements and have the alkali metals deposit onto the solid particulates. Initial removal of the particulates also facilitates downstream processing by eliminating blinding of catalyst beds with particulates, solids in liquid absorption systems, and the need for treatment of entrained solids in process water condensate. Removing these abrasive particulates also reduces maintenance costs for downstream piping and equipment.
Halogens	These species are known for their high reactivity, allowing them to poison most catalysts, degrade solvents, and corrode process equipment. Removing these species as soon as possible simplifies the design of downstream equipment and processes.
Tars	For fuels and gasification systems that generate significant concentrations of tars, these tars represent inefficient fuel conversion to product. Therefore, the tars need to be separated and converted to valuable products or recycled to the gasification system for syngas generation. At operating temperatures below its dewpoint, tar will condense and coat surfaces, reducing heat and mass transfer in heat exchangers and in catalyst beds and making removal difficult.

Table 6-9. Typical integration sequence for contaminants

CONTAMINANT	DESCRIPTION
H_2-to-CO Adjustment	This maximizes the use of the steam concentration available in the raw syngas and any additional steam added by quenching/scrubbing of the raw syngas for particulate removal. This also integrates the WGS process when the syngas is first cooled to an acceptable operating temperature. Alternative conversions that could be integrated include COS hydrolysis, HCN hydrolysis, and/or hydrogenation. Both conversions occur as secondary reactions in the WGS process. The SGS catalyst, which operates in a syngas with high-sulfur concentrations, will also help separate some of the arsenic species present in the syngas.
Water	In preparation for commercial solvent-based absorption processes, the syngas is cooled to less than 100°F, which removes a significant amount of the water vapor in the syngas and other highly water-soluble species, including NH_3.
Hg	At this point, the syngas has been cooled for commercial solvent-based absorption acid gas treatment. The temperature of this cooled syngas matches the specifications for Hg removal with commercial sulfur-impregnated activated carbon.
Sulfur	At a minimum, the sulfur species need to be removed for both power and chemical production. Typical commercial solvent-based absorption processes operate at less than 100°F. As this is one of the last removal processes, synergistic removal of any remaining trace contaminant concentrations is exploited as much as possible.
CO_2	If required, CO_2 removal is usually integrated with sulfur removal to exploit effective process intensification of AGR (i.e., H_2S and CO_2).
Polishing/Guard Bed	During preconditioning of the treated syngas, final polishing or guard beds are integrated to protect the syngas conversion process. In an IGCC power production application, a final filter is included to ensure that the particulate loading to the gas turbine does not exceed the specifications of the manufacturer. The capital and operating cost of this final filter are less than the replacement costs for a gas turbine, making it a wise insurance investment. Similarly, final guard beds for specific species (like sulfur) could be added to protect downstream catalyst/equipment.

Standard syngas cleanup system configurations have been developed based on the typical integration sequence for contaminants (Table 6-9), the available commercial processes, and experience—in particular, chemical applications with higher cleanup requirements than power production applications. A block flow diagram of the standard syngas cleanup system configuration for chemical production applications is shown in Figure 6-5. Treatment in this system configuration begins with syngas quench/scrubbing to remove particulates, separate halogens, and add enough steam for the WGS process. In the next step, a WGS process using an SGS catalyst adjusts the H_2-to-CO ratio for optimal syngas conversion. For most chemical conversion applications (excluding hydrogen production), this WGS process involves a single reactor with syngas bypass loop (Figure 6-1). This process also converts COS into H_2S and HCN into NH_3. The SGS catalyst will also separate a portion of the arsenic species. The syngas from the WGS process is cooled to less than 100°F for acid gas treatment, resulting in the condensation of most of the water vapor in the syngas and any highly water-soluble contaminants such as NH_3 remaining in the syngas. At this temperature, Hg removal with sulfur-impregnated active carbon can be performed. Finally, Rectisol® is used to separate the sulfur and CO_2 from the syngas. Rectisol® is the standard for chemical applications, as it achieves treated sulfur concentrations of less than 0.1 ppmv and treated CO_2 concentrations of less than 10 ppmv. These low concentrations meet or exceed the specifications for all chemical production applications. The Rectisol® process also provides final removal for any remaining Hg, HCN, NH_3, carbonyls, and hydrocarbons.

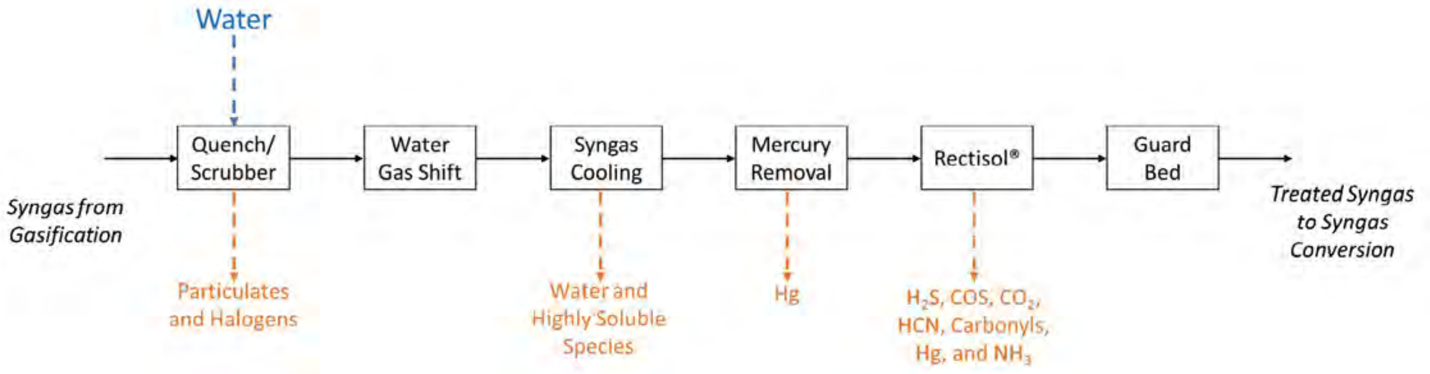


Figure 6-5. Typical syngas cleanup configuration for chemical production

For IGCC power production, the specifications are less stringent than chemical production. The important modifications to the syngas cleanup configuration include:

- Modification of particulate removal to recover heat at higher exergy and dry particulates.
- Elimination of the WGS process (unless CO_2 capture is required).
- The use of a solvent-based absorption process that operates at higher temperatures, increasing the overall thermal efficiency of the plant, and minimizes CO_2 removal from syngas (high sulfur selectivity)—unless it is required.

For syngas cleanup system configurations enabling CO_2 capture, the process reverts closer to the standard for chemical production. The difference from the standard configuration for chemical production is that the solvent-based absorption process for AGR is operated at a higher temperature to increase overall thermal efficiency and limit sulfur removal to meet the sulfur emission limits for power production. Selexol™ is one of the most promising process technologies for this application, because it offers higher operating temperatures; can achieve sulfur removal that meets environmental emission regulations for power production; and can simultaneously separate sulfur and CO_2, enabling CO_2 capture. Example syngas cleanup configurations for IGCC applications without and with CO_2 capture using Selexol™ for AGR are shown in Figure 6-6.

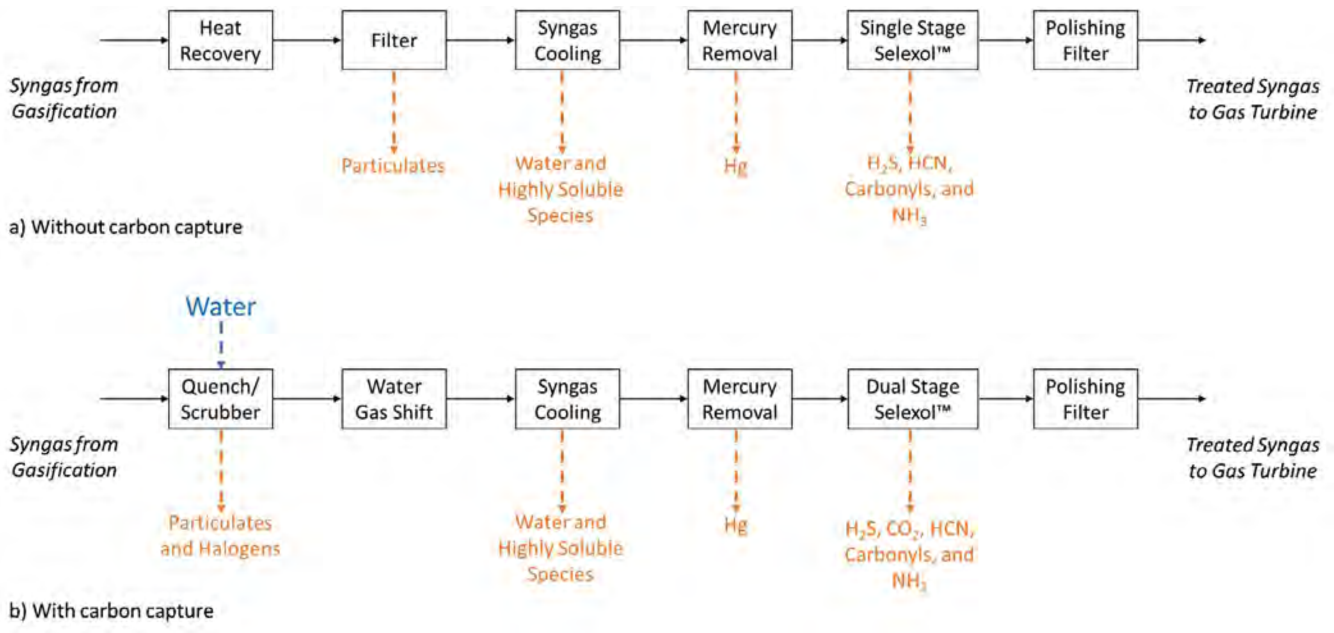


Figure 6-6. Selexol™-based syngas cleanup configurations without and with CO_2 capture for IGCC applications

6.4.2 DESIGN PROCESS

The emphasis of the proceeding sections has been the selection methodology for assembling the different gas cleanup processes into a syngas cleanup system. The methodology is to sequentially separate the most adverse to least adverse contaminant on downstream catalysts and equipment while simultaneously maximizing plant thermal efficiency. This effort results in a block flow diagram for the syngas cleanup system. For the configuration in this block flow diagram, heat and mass balances need to be developed to meet the technical removal objectives. These heat and mass balances need to be used to develop size equipment lists for the syngas cleanup system; these size equipment lists are then used to generate factored equipment and operating cost estimates, which can be compared with economic performance objectives. Based on the projected economic performance of the plant, modifications to the process configuration can be suggested to improve technical and economic performance. Using these steps in Table 6-10, an iterative design process can be implemented that allows optimization of the integration of the syngas cleanup system with the technical and economic performance. This approach can also be used to evaluate the risk(s) associated with the different technologies and evaluation of potential contingencies using methodologies employed by NETL in their system analysis.[32]

Table 6-10. Fundamental steps in iterative design approach

STEP	DESCRIPTION
Develop a Basic Block Flow Diagram	The diagram provides an integrated scheme for the different gas cleanup process technologies to meet syngas end-use requirements while maximizing removal process and thermal efficiency. Additional considerations are exploiting demonstrated commercial experience, disruptive potential of emerging technologies, and opportunities for cost beneficial process intensification.
Perform Heat and Mass Balances	Heat and mass balances for the process configuration are developed to meet the technical removal objectives for the proposed plant.
Develop Sized Equipment Lists	The heat and mass balances are used to develop sized equipment lists for the major process equipment in the block flow diagram.
Develop Factored Capital and Operating Cost Estimates	The sized equipment list is used to develop estimates for plant capital and operating costs.
Evaluate the Performance Economic Against Plant Economic Objectives	If the estimates for economic performance meet the plant's economic objectives, the design process is complete. Alternatively, modifications may be made to the process configuration to reduce costs and achieve economic performance objectives.
Suggest Modifications to the Process Scheme	Include modification(s) into the previous block flow diagram and repeat the process.

Once the iterative process has identified a syngas cleanup system that meets the technical and economic objectives for the proposed modular plant, a preliminary hazard review analysis should be performed to ensure that all major safety aspects of the syngas cleanup system have been incorporated into the design. Although safety should be part of the technical requirements included in the development of each process configuration, this additional step evaluates the safety of the proposed process configuration, ensuring that no additional safety equipment or requirements need to be added to the cost estimates. If the process configuration lacks important safety considerations, the economic and thermodynamic analyses should be redone with the new safety precautions incorporated to ensure that all economic and performance requirements are met.

A final recommendation for the preliminary design process is to evaluate non-standard process operations like startup and shutdown to optimize the performance of these steps. The first step is to develop a detailed strategy for startup and shutdown of the syngas cleanup system. Startup and shutdown sequences need to consider integration of the startup or shutdown of these different removal processes while minimizing the overall startup or shutdown time, because the syngas cleanup system is composed of multiple different removal processes. These plans should include estimates of the resources (e.g., additional steam, purge gas, heat, staff) and emissions (namely partially treated syngas), which must fall within the permitted limits to ensure the plan is realistic and feasible. Evaluation of these non-standard process operations as part of the preliminary design effort enables a better evaluation of the balance of plant systems and auxiliary equipment needs. This also permits additional iterative steps to modify the syngas cleanup systems to facilitate these non-standard process operations while the design is at a conceptual level and prior to the investment of heavy engineering resources in a detailed design.

With strategies for both the standard and non-standard operations, the control strategy for the individual processes during normal and non-standard operations should be included as supplemental detail in the block flow diagram. This enables an estimate of the amount and location of the required and supplemental instrumentation and analytical systems throughout the system. If these details are used to refine the cost estimates, the resulting cost estimates provide the best possible estimate based on a comprehensive preliminary design effort.

The final product should be a preliminary design for the syngas cleanup system that meets technical and economic performance objectives during operation, addresses all key and fundamental safety issues, and incorporates the auxiliary systems and solid plans for non-standard process operations (e.g., startup and shutdown). This design approach also defines the risk factors for the syngas cleanup system prior to proceeding with a large project investment.

6.5 SUMMARY

The goal of the syngas cleanup system is to transform the raw syngas generated in the gasification system into a treated syngas that enables optimal productivity of the syngas conversion process. The combination of the fuels and gasification process results in a raw syngas mixture that contains a mixture of entrained solid particulates and condensable and non-condensable gaseous species made from the large variety of elements found in the fuel. The optimal treated syngas for conversion would contain exclusively H_2 and CO in the proper ratio at a suitable temperature and pressure for the conversion process. Therefore, the transformation requires:

- Removal of entrained solids.
- Removal of contaminants.
- Shifting the syngas composition to achieve a specific H_2-to-CO ratio.
- Adjusting the syngas temperature.
- Adjusting the syngas pressure.

Different specifications for syngas conversion for chemical versus power applications result in differences in the syngas cleanup system. Chemical applications typically have more stringent specifications and require more complex syngas cleanup systems. However, the syngas specifications for power applications that require CO_2 capture are essentially identical to those of chemical applications (see Chapter 12 for discussion about CO_2 capture-based power applications).

The recommended design approach uses an iterative process where individual gas cleanup processes for the different contaminants are assembled into a process configuration with the objective of sequentially removing the contaminants in the order of greatest to least adverse impact on downstream processes while maximizing the overall thermal efficiency of the plant. Removal processes that have extensive commercial operating experience under actual conditions provide the lowest risk solutions with the lowest need for contingency. A potential challenge with these existing commercial removal processes is their optimization for large-scale commercial applications where economies of scale are realized. For smaller scale modular plants, these existing commercial processes may not be suitable. This may require the project developer to evaluate emerging gas cleanup technologies, which will introduce risk and require the introduction of duplicate and extra removal processes to ensure meeting treated syngas specifications. The extra equipment is primarily for the trace contaminants. For existing commercial removal processes, the secondary removal of trace contaminants has been demonstrated to meet the most stringent chemical production specifications. For most emerging removal processes, secondary removal of trace contaminants has not typically been demonstrated or evaluated. Thus, earlier applications of these emerging technologies in modular plants will include duplicate equipment to address technical risk with the emerging removal technology and extra equipment to ensure appropriate treatment of trace contaminants. As more experience is accumulated, systems with emerging removal technologies can achieve the technical and economic objectives with lower risk and less equipment.

Once a process configuration for the syngas cleanup system is complete, the configuration is used to develop the heat and mass balances to meet treated gas specifications. The heat and mass balances permit the development of a sized equipment list of the major pieces of equipment. Factored estimates for capital equipment and operating costs can then be developed from the sized major equipment list. With these cost estimates, the project developer can evaluate the economic objectives of the plan.

When both the technical and economic objectives for the plant have been achieved, several final evaluations are recommended for reducing project and plant risk. The first evaluation is a preliminary design hazard review on the proposed configuration for the syngas cleanup system. This review ensures that most of the safety and environmental concerns are addressed by the syngas cleanup system. The second recommendation is to evaluate the operating plans for non-standard operations like startup and shutdown to ensure that auxiliary equipment and balance of plant systems are adequate for startup and shutdown of the system. If modifications to the syngas cleanup configuration are required to address either of these recommendations, additional iterations of the design process are used to update the process economics and confirm technical and economic objectives.

This design approach allows evaluation of the many possible process configurations based on the variety of different available gas cleanup technologies and the variety of contaminants in the raw system that must be removed for optimal syngas conversion. Although this approach is valuable for larger applications based on existing commercial technologies, it is most useful for smaller, modular plants that need to use emerging gas cleanup technologies. The application of this design approach also reduces the risks associated with the project and establishes a basis on which to begin the detailed design of the syngas cleanup system.

Endnotes

1 *40 CFR Part 60 Standards of Performance for Fossil-Fuel-Fired Electric Utility, Industrial-Commercial-Institutional and Small Industrial-Commercial-Institutional Steam Generating Units, Final Rule,* EPA, December 16, 2011.

2 D. A. Horazak, R. A. Newby, E. E. Smeltzer, R. B. Slimane, P. Vann Bush, J. L. Aderhold and B. G. Bryan, “Novel Gas Cleaning/Conditioning for Integrated Gasification Combined Cycle 4: Volume 1 – Conceptual Commercial Evaluation Operational Program Final Report,” DOE Award Number: DE-AC26-99FT40674, December 2005.

3 K. Lui, C. Song and V. Subramani, “Hydrogen and Syngas Production and Purification Technologies,” AICHE Wiley and Sons, 2010.

4 Sasol, “Sasol Generic SHE specifications for Service Providers,” 2021. [Online]. Available: https://www.sasol.com/sites/sasol/files/content/files/10.8%20SSC-SAF-GPR-00016%20Sasol%20Generic%20SHE%20specifications.pdf. [Accessed 2021].

5 B. M. T. Turk, A. Lopez-Ortiz, R. Gupta, J. Portzer, G. Krishnan, B. Freeman and G. Fleming, “Novel Technologies for Gaseous Contaminants Control Final Report for the Base Program,” DOE Contract number DE-AC26-08FT40675, September 30, 2001.

6 Y. Yi, A. Rao, J. Brouwer and A. Samulsen, “Ammonia as a Contaminant in the Performance of an Integrated SOFC Reformer System,” in *4th Conference on Fuel Cell Science, Engineering and Technology, ASME,* Irvine, CA, 2006.

7 *40 CFR Part 63 National Emissions Standards for Hazardous Air Pollutants from Coal- and Oil-Fired Electric Utility Steam Generating Units,* EPA, December 16, 2011.

8 G. Krishnan, P. Jayweera, K. Lau and A. Sanjurjo, “Effect of Coal Contaminants on Solid Oxide Fuel System Performance and Service Life, Technical Progress Report 3,” SRI International, 2006.

9 F. N. Cayan, Z. Mingjia, S. R. Pakalapati, I. Celik, N. Wu and R. Gemmen, “Effects of Coal Syngas Impurities on Anodes of Solid SOFCs,” *Journal of Power Sources,* vol. 185, pp. 595-602.

10 C. N. Hamelinck, A. P. C. Faaij, H. den Uil and H. Boerrigter, “Production of FT Transportation Fuels from Biomass; Technical Options, Process Analysis and Optimization, and Development Potential,” *Energy and Fuels*, vol. 29, pp. 1743-1771, September 2004.

11 J. Clark, Q. Li, G. Lilik, N. Reed and C. Wang, “Novel Design of an Integrated Pulp Mill Biorefinery for the Production of Biofuels for Transportation,” EGEE, vol. 580, 4 May 2007.

12 Now G. Jacobs and B. Davis, “Sensitivity of Fischer-Tropsch Synthesis and Water Gas Shift Catalysts to Contaminants from Gasified Biomass and Coal/Biomass Mixtures,” Center for Applied Research University of Kentucky, DOE Award Number DE-PS$_2$6-0, October 16, 2009.

13 H. Liebold, A. Hornung and H. Seifert, “Gas Cleaning at High Temperatures,” *Powder Technology*, vol. 180, no. 1-2, pp. 265-270, 20-22 October 2005.

14 J. Trembly, R. Gemmen and D. Bayless, “The Effect of Coal Syngas Containing HCl on the Performance of Solid Oxide Fuel Cells: Investigations into the Effect of Operational Temperature and HCl Concentration,” *Journal of Power Sources*, vol. 169, p. 347, 2007.

15 F. N. Cayan, Z. Mingjia, S. R. Pakalapati, I. Celik and N. Wu, “Effects of Coal Syngas Contaminants on SOFC Anodes: A Review,” in *Proceedings of the Mini-Symposium “Coal Based Fuel Cell Technology Status, Needs and Future Applications*, Morgantown, WV, 2007.

16 O. Marina, L. Pederson, D. Edwards, C. Coyle, J. Templeton, M. Engelhard and Z. Zhu, “SOFC Operation on Hydrogen and Coal Gas in the Presence of Phosphorous, Arsenic, and Sulfur Impurities,” in *8th Annual SECA Workshop*, San Antonio, TX, 2007.

17 J. Stevenson, O. Marina, L. Pederson, C. Coyle, E. Thomsen and D. Edwards, “SOFC Tolerance Limits for Phosphorous and Arsenic,” in *11th Annual SECA Workshop*, Pittsburgh, PA, 2010.

18 M. Schafer and G. Alptekin, “Testing for the Effects of Syngas Contaminants on WGS and FT Catalysts, Kick Off Presentation,” TDA Research Inc., n.d.

19 R. Schoff and N. Holt, “Gasification Technology Options for SOFC Applications,” in *8th Annual SECA Workshop*, San Antonio, TX, 2007.

20 G. Energy, “Specifications for Fuel Gases for Combustion in Heavy-Duty Gas Turbines, GEI 41040j,” GE Energy, April 2007.

21 P. Bergmann, S. van Paasen, H. Bergmann and H. Boerrigter, “The Novel ‘OLGA’ Technology for Complete Tar Removal from Biomass Producer Gas,” in *Pyrolysis and Gasification of Biomass and Waste Expert Meeting*, Strasbourg, 2002.

22 H. Boerrigter, S. van Paasen, P. Bergmann, J. Könemann, R. Emmen and A. Wijnaands, ““OLGA” Tar Removal Technology: Proof of Concept for Application in Integrated Biomass Gasification Combined Heat and Power Systems,” Patten ECN, 2005.

23 W. Trapp, “Eastman Coal Gasification: 19 Years of Reliable Operation,” in *ICHemE Conference: Gasification: The Clean Choice for Carbon Management,* Noordwijk, 2002.

24 D. Bell, B. Towler and M. Fan, “Chapter 6. Sulfur Recovery,” in *Coal Gasification and its Applications*, Elsevier Inc., 2010, pp. 113-136.

25 R. Bucklin and R. Schendel, “Comparison of Flour Solvent and Selexol Process,” Energy Progress, vol. 4, no. 3, p. 139, 1984.

26 R. Siriwardane, “Durable Regenerable Sorbent Pellets for Removal of Hydrogen Sulfide from Coal Gas”. U.S. Patent 5,703,003, 1997.

27 R. Siriwardane, “Durable Zinc Oxide-Containing Sorbents for Coal Gas Desulfurization”. U.S. Patent 5,494,880, 1996.

28 G. Howe, J. Albritton, D. Denton, R. Gupta and B. Turk, “RTI Warm Syngas Cleanup Operational Testing at Tampa Electric Company’s Polk 1 IGCC Site,” Final Technical Report submitted under DOE/NETL Cooperative Agreement DE-FE0026622, 2018.

29 Nexant, RTI International, and Aerojet Rocketdyne, “Integration of Aerojet Rocketdyne and RTI Advanced Gasification Technologies for Hydrogen-Rich Syngas Production Task 2: Techno-Economic Analysis for IGCC Application,” Technical Report submitted under DOE/NETL Cooperative Agreement DE-FE0012066 RD&D, 2016.

30 V. Gupta, P. Sharma, D. Denton and B. Turk, “Benefits of Integration of Aerojet Rocketdyne and RTI Advanced Gasification Technologies for Hydrogen-Rich Syngas Production Task 5: Techno-Economic Analysis for Coal-to-Liquids via Fischer-Tropsch Synthesis,” Technical Report submitted under DOE/NETL Cooperative Agreement DE-FE0012066, 2016.

31 H. Boerrigter, H. den Uil and H. P. Calis, “Green Diesel from Biomass via Fischer-Tropsch synthesis: New Insights in Gas Cleaning and Process Design,” in *Pyrolysis and Gasification of Biomass and Waste Expert Meeting*, Strasbourg, 2002.

32 NETL, “Quality Guidelines for Energy System Studies: Cost Estimation Methodology for NETL Assessments of Power Plant Performance,” DOE/NETL 2011/1455, April 2011.

7.0 GUIDELINES FOR THE DESIGN OF A POWER GENERATION MODULE

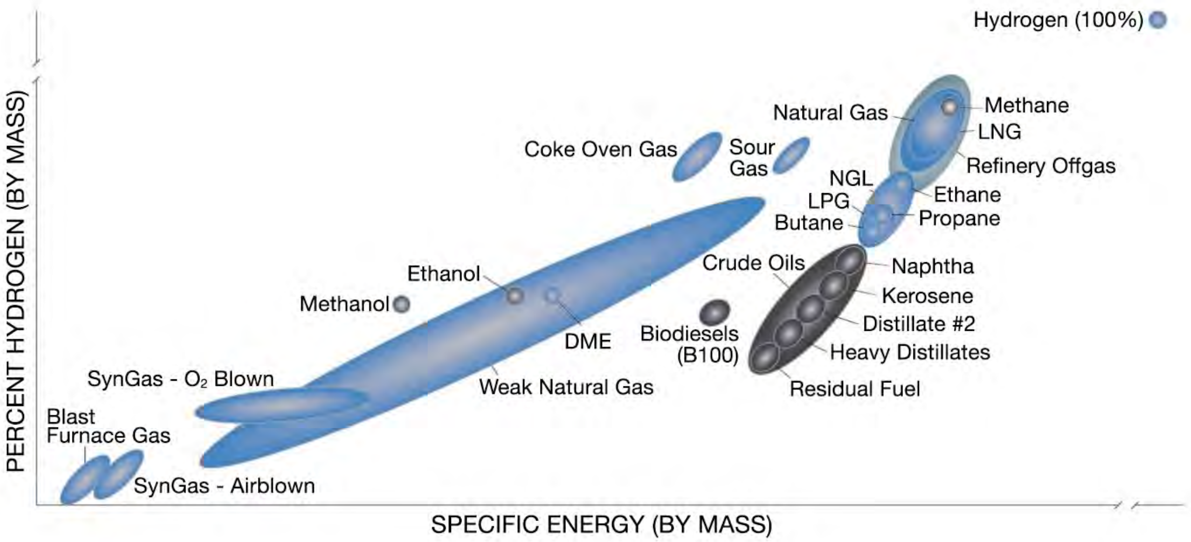
7.1 INTRODUCTION

The attractiveness of gas turbines for gasification-based, combined-cycle power and co-generation plants is derived from their high fuel flexibility, efficiency, and power density. They are proven over a wide range of unconventional fuel applications and well suited to the modularized plant concept. Gas turbines offer a wide span of configurations and integration options that can be tailored to deliver optimal economic and operational benefit. Key options are the flexibility of steam generation for additional power or supply of process heat for chemical production (i.e., co-production/polygeneration) or cogeneration, such as district heating (i.e., combined heating and power [CHP]). The successful application of a power block in a modular plant requires the developer to be informed of the available power and plant integration options. The intent of this chapter is to provide a developer already familiar with natural gas turbine or combined cycle plants with an introduction to the configuration and operational differences of synthesis gas (syngas)-fueled turbines. The goal is to enable an informed tradeoff between performance, cost, reliability, and operational flexibility and their impact on environmental performance and permitting strategy (see Section 7.4).

7.2 SYNGAS FOR POWER

For an integrated gasification combined cycle (IGCC) plant, the combined power block (i.e., gas turbine, steam generator, and steam turbines) constitutes approximately 10 to 15% of the total plant cost.[1] However, the turbine choice determines the size, cost, and configuration requirements of the entirety of the plant. The heating values (in British thermal units [Btu]/standard cubic feet [scf] and Btu/lb) of syngas produced from gasifying carbonaceous fuels are well below those of other widely used fuels[2] (Figure 7-1). Gas turbines have been applied over most gaseous and liquid fuel types. The syngas derived from both air- and oxygen (O_2)-blown gasifiers is located at the bottom left corner of Figure 7-1. This is the domain for gas turbines. In 2016, the global inventory of syngas fueled turbines was approximately 16 gigawatts-thermal (GWth)[3] for plants operating or under construction.

While internal combustion engines (ICEs) are attractive due to their operating flexibility, efficiency, and size, they have limited capability for low heating value gases. The application of ICEs to gaseous fuels is currently limited to natural gas and biogas with a lower volumetric heating value limit of 10,817.3 Btu/lb[4] (weak methane at 1 atm, 70°F). Hydrogen normally present in syngas can cause pre-ignition before full compression. The option of syngas refueling and repowering for boiler-based steam electrical power generation is discussed in Section 7.13.



Used with permission from GE[2]

Figure 7-1. Heating values for liquid and gaseous fuels showing syngas among the lowest

7.2.1 GAS TURBINES

Syngas-capable turbines are available over a wide range of capacities (Table 7-1). Even the largest heavy-duty gas turbines (HDGTs) and their generators (Figure 7-2)[5] can be shop fabricated, assembled, and transported by rail and truck. For gas-turbine, combined-cycle power blocks, steam turbine stages, heat recovery steam generators (HRSGs), and condensers are configured as knock-down/field modular (i.e., site assembly) optimized for transport and efficient use of site manpower resources at the construction site. Manufacturers draw from their long experience to determine the optimum balance of modularization and shop fabrication for major components and equipment. The development of a standard modularized design requires detailed, pre-construction engineering and design to minimize site construction time. Suppliers can also provide experience in site management planning and logistics.

Table 7-1. Representative capacities of commercial gas turbines with syngas capability

MODEL	SUPPLIER	HZ	CLASS	MWE (NG-ISO)
LM2500	GE[2]	60	Aero	24
6B	GE[6]	60	HDGT	43
6FA	GE[6]	60	HDGT	78
M701DA	MHI[7]	50	HDGT	140
7F Syngas	GE[6]	60	HDGT	232
GT26	Ansaldo[8]	50	HDGT	370
SGT-A65	Siemens[9]	60	Aero	58-62
SGT5-2000E	Siemens[10, 11]	60	HDGT	187
SGT6-5000F	Siemens[10]	60	HDGT	260
M701F4	MHI[7]	50	HDGT	324

Used with permission from GE[5]

Figure 7-2. Large gas turbine prepared for shipment; combustors, pre-fabricated manifolds, and control skids will be shipped separately

7.2.2 COMMERCIAL AVAILABILITY

The time span from project start to commercial operation date (COD) impacts project cost. Longer projects incur more financing interest on debt during the capital expenditure period, in addition to material, equipment, and labor cost escalation. Even if listed in brochures, they are not necessarily available as "off-the-shelf" delivery. Project developers should recognize that turbine delivery time is a key driver of schedule. Other schedule elements, such as project permitting, should be underway as early as possible. After a notice-to-proceed (NTP) is issued and orders are placed, the turbine supplier's global supply chain is activated for the materials and manufacture of key components, such as rotor forgings, casing castings, and sophisticated processes required for hot gas path components; many of which will be specific to syngas. Some examples include the substitution of syngas diffusion combustors in place of natural gas dry, low-NO_X (DLN) combustors, turbine nozzles with larger throat areas, longer last-stage buckets and shrouds, and hot gas path components with special environmental coatings for syngas.

Manufacturers claim delivery of natural gas turbines in as little as six months; syngas turbines take longer. As a result of their unique components, materials, and engineering, it is not uncommon for the delivery of a syngas turbine to take 12 to 18 months from order to delivery. One reason is the low order volume of syngas turbines. Each project can also have special requirements that necessitate additional design engineering—some may require combustion testing that also stretches delivery. After mechanical completion, commissioning, systems checkout, first fire and combustion tuning can take up to an additional year. The installation and commissioning time should be incorporated with margin into the project schedule. As a part of a developer's due diligence, the manufacturer should be required to supply a listing of reference plants and operating plant contacts. Operators can provide information on experience and, in many cases, recommendations to address with the manufacturer. Discussions should begin as early as possible with turbine suppliers regarding their delivery timelines, delivery schedules, and the manufacturing slot an order will be placed.

To break the paradigm of every turbine being "special" order, a goal of the modular concept is standardization based on a limited range of syngas turbine models. Plant scalability can be accommodated by multiple turbines providing a benefit in operational flexibility (e.g., lower power turndown ratios with increased reliability). This approach can help to increase order volume, which provides incentive for manufacturers to maintain dedicated resources, build a secure and agile supply chain, spread development cost recovery over a larger number of turbines, and shorten delivery times. This will incentivize manufacturers to include modular plants as a market requiring responsiveness to remain competitive.

7.3 KEY FACTORS IN TURBINE CHOICE

Turbine suppliers are continuously developing new models and upgrades for their operating fleets. Most syngas turbines are HDGTs with industrial single-shaft (i.e., shared turbine and compressor shaft) and commonly configured for combined-cycle plants. General Electric (GE) also offers a syngas-capable aeroderivative (i.e., two-shaft, free-spin power and compressor stage) that can provide quick response—shorter startup and lower turndown capability—than HDGTs. On a simple cycle basis, aeroderivatives (Aeros) also hold an efficiency advantage over HDGTs (approximately 5 to 7 points based on lower heating value [LHV]) and are well suited for peaking service. However, peaking service is not likely to be a common requirement for modular gasification plants. The disadvantages of Aeros include limited choice (only the LM2500 was found to be syngas capable), small size, higher cost (in terms of both capital expenses [CAPEX] in $/kilowatt [kW] and operational expenses [OPEX] in $/kilowatt-hour [kWh]). Therefore, the following discussion will be focused on HDGTs.

The turbine should incorporate only commercially proven components that have been proven in operation with syngas of similar composition and combustion properties. In terms of the U.S. Department of Energy's (DOE) Technology Readiness Level (TRL), the recommended TRL is 9 (*i.e., technology is in in its final form and the actual system[s] is [are] operating over the full range of expected conditions*). Proof should be operation of one completed failure-free maintenance interval (minimum 24,000 to 32,000 total fired hours) and verification with a major inspection.

Turbine selection dictates the size of all the other equipment in the plant. If the turbine fuel consumption is low, the size of the gasifier and air separation unit (ASU) will be proportionately smaller. Likewise, the gas cleanup and sulfur recovery sections of the gasification block will also be smaller to accommodate the reduced volume of syngas. Other turbine characteristics and their influence on other plant components include the following:

- Turbine compressor pressure ratio, performance map, and efficiency.
    - *Pressure ratio sets syngas pressure to the turbine, and gasifier operating pressure is designed to be significantly higher for control safety.*
- Exhaust flow rate and temperature.
    - *Defines the amount of energy that can be economically recovered with a heat recovery steam generator and choice of a bottoming steam turbine cycle.*
- Criteria pollutant emissions (nitrogen oxide [NO_X], carbon monoxide [CO], volatile organic compounds [VOCs], particulate matter [PM]).
    - *Compliance with air quality regulations and facility permitting.*
- Efficiency.
    - *Both economics and compliance for output-based emissions.*

7.3.1 SYNGAS TURBINE IMPACT

The thermodynamic properties of gasification-produced syngas differ from those of natural gas in several key aspects (Table 7-2).[12]

Table 7-2. Representative syngas properties by gasifier type (coal feedstock)[12]

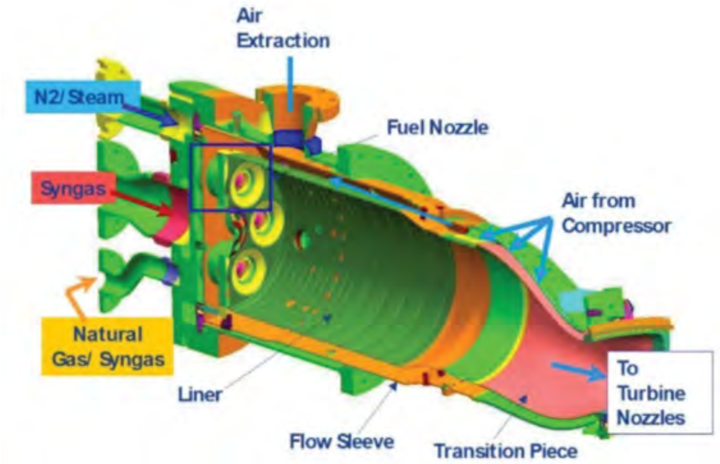
GASIFIER TYPE	% VOL H_2	DENSITY (LB/SCF)	LHV (BTU/LB)	LHV (BTU/SCF)	WOBBE INDEX
O_2/Steam	37.3	0.0496	6,096	302	12.85
Air Blown	18.7	0.0617	2,575	159	6.06
Natural Gas	0.0	0.0423	21,553	910	7.45

7.3.1.1 HYDROGEN

Hydrogen content makes syngas unsuitable for current fuel/air premix DLN and Dry Low Emission (DLE) combustors due to hydrogen's high flame speed (Table 7-3)[13] and broad flammability range (rich and lean fuel-to-air ratios) that can cause flame migration upstream to the combustor head end and into the combustor pre-mixer. (GE has recently introduced a DLN-2.6e combustor incorporating micro-mixer channels for premixed hydrogen [H_2]/NG with capability to 50% H_2vol for the 9HA.02 turbine.[13] Power Systems, Mfg. LLC (PSM) has developed a FlameSheetTM combustor for up to 65% H_2.[14]) Pre-mix combustors are limited to a maximum of 5 to 10% volume H_2. The solution for higher hydrogen is diffusion flame combustion (e.g., GE Multi-Nozzle Quiet Combustors [MNQCs] for HDGTs[15] [Figure 7-3] and singular annular combustors [SACs; Figure 7-4] for Aeros). Syngas-capable diffusion combustors are also available from Siemens/Ansaldo and Mitsubishi Heavy Industries (MHI). Hydrogen requires special operating safety procedures, such as startup and shutdown on natural gas, plus stricter purge requirements for fuel systems, turbines, HRSGs, and exhaust ductwork (further discussion of safety issues is in Section 7.12).

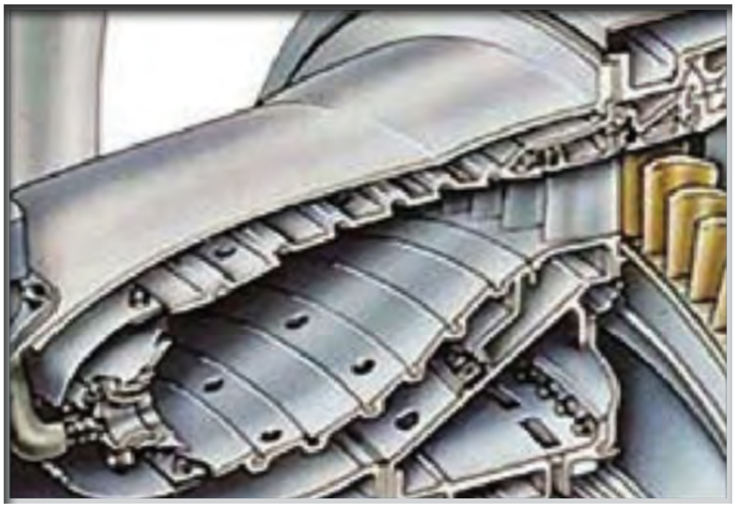
Table 7-3. Laminar flame speeds for various gases -H_2 being ~4X that of other fuel gases[13]

FUEL	FORMULA	LAMINAR FLAME SPEED (CM/SEC) AT STOICHIOMETRIC CONDITIONS
Hydrogen	H_2	170
Methane	CH_4	38.3
Ethane	C_2H_6	40.6
Propane	C_3H_8	42.3
Carbon Monoxide	CO	58.8



Used with permission from GE[16]

Figure 7-3. Diffusion syngas combustor for a heavy-duty gas turbine



Used with permission from GE[2]

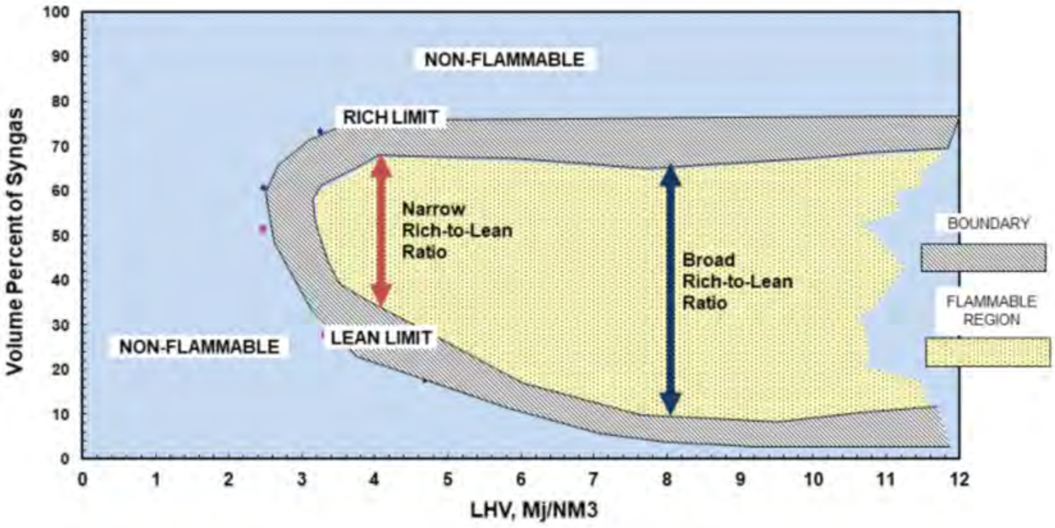
Figure 7-4. Annular syngas diffusion combustor for an aero-derivative turbine

7.3.1.2 HEATING VALUES

Depending on feedstock and gasifier type (e.g., air- or O_2-blown), the syngas volumetric heating value (Btu/scf) can range from 100 to greater than 300 Btu/scf. These values are 20 to 40% of the heating value of natural gas and require combustor ability to both initiate and maintain stable combustion. Figure 7-5 shows the flammability limits for syngas combustion in air.[17] At a heating value of approximately 100 Btu/scf and an equivalence ratio[*] of 1, combustion is quenched and combustor light-off cannot be achieved regardless of equivalence ratio. As the syngas heating value increases, stable combustion is supported over a wider range of equivalence ratios. Due to imperfect mixing within the combustor, localized fuel-to-air ratios can vary. The combustor design must employ comprehensive and interactive modeling, testing, and field verification for the manufacturer to guarantee emissions performance (e.g., CO, NO_X, VOCs) with stable and robust operation. The combustor must also be capable of seamlessly accomodating turndown, fuel transfers, and load following transients. Extensive experience with diffusion combustors has proven the ability to achieve these requirements. It should be noted that hydrogen (i.e., greater than 15%) is beneficial and improves the combustion stability and reduction of CO. Syngas combustion and proven emissions mitigation techniques are discussed in Section 7.9.1.

---

[*] Equivalence ratio is defined here as the ratio of actual fuel-to-air ratio to a fuel's stoichiometric fuel-to-air ratio.



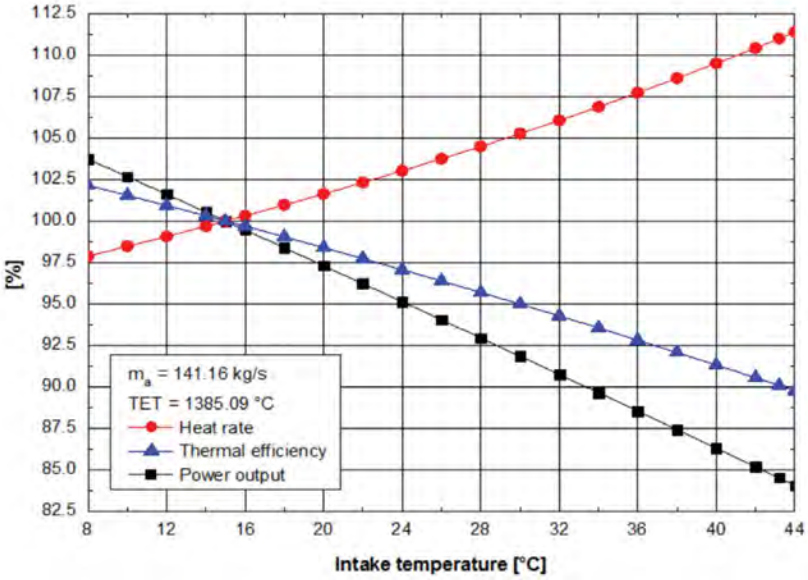
Used with permission from GE[17]

Figure 7-5. Theoretical combustion flammability limits for syngas

7.3.2 OUTPUT ENHANCEMENT

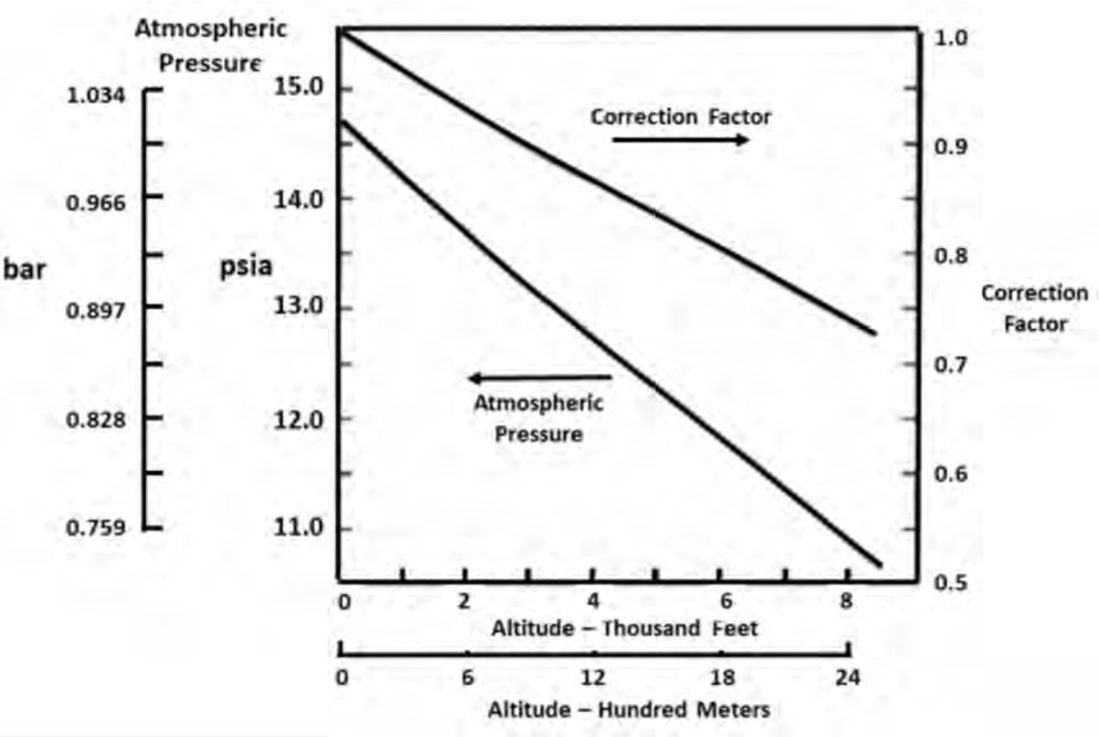
Turbine integration with other major components of an oxygen-blown gasification plant offers a significant output performance advantage over natural gas. Syngas turbines can deliver output at their torque limit to higher ambient air temperature than natural gas turbines. Additional considerations are needed beyond turbine performance at ambient conditions established as standards by the International Organization for Standardization (ISO), which are stated to be 15°C/59°F and 60%RH at sea level. For ambient temperatures below ISO conditions, the output will generally be limited by the turbine shaft torque limit; above ISO conditions, there is a rapid decrease in output due to reduced air density. The torque limit for a machine is designed based on low ambient temperatures (normally 32°F or less) so that there is a large enough factor of safety for normal operating temperatures.

For a plant having power as its revenue source, the ability to maintain high power output at higher ambient air temperatures provides a significant revenue and economic benefit. The underlying physics is the increased turbine mass flow with syngas. A turbine compressor, operating at constant speed and geometry, is in effect a "constant volume flow" machine.[18] The volume flow into the compressor is equal to the average axial velocity multiplied by compressor inlet flow area. At a constant rotational speed, the average inlet gas velocity at any location is essentially constant. The power delivered from turbine to generator (in watts [W]) is torque (in Newton-meters [N-m]) x rotational velocity (in Hertz). Since the mass flow rate through the turbine is equal to the volumetric flow rate (V) multiplied by the gas density, mass flow rate will vary directly with ambient air density (i.e., pressure and temperature at the inlet to the compressor, where pressure and temperature are at ambient, causes the air density to inversely vary with temperature). The performance degradation due to ambient temperature[19] and altitude[20] are shown in Figures 7-6 and 7-7. This degradation can be significant and must be incorporated into the project's economics.



Used with permission from GE[19]

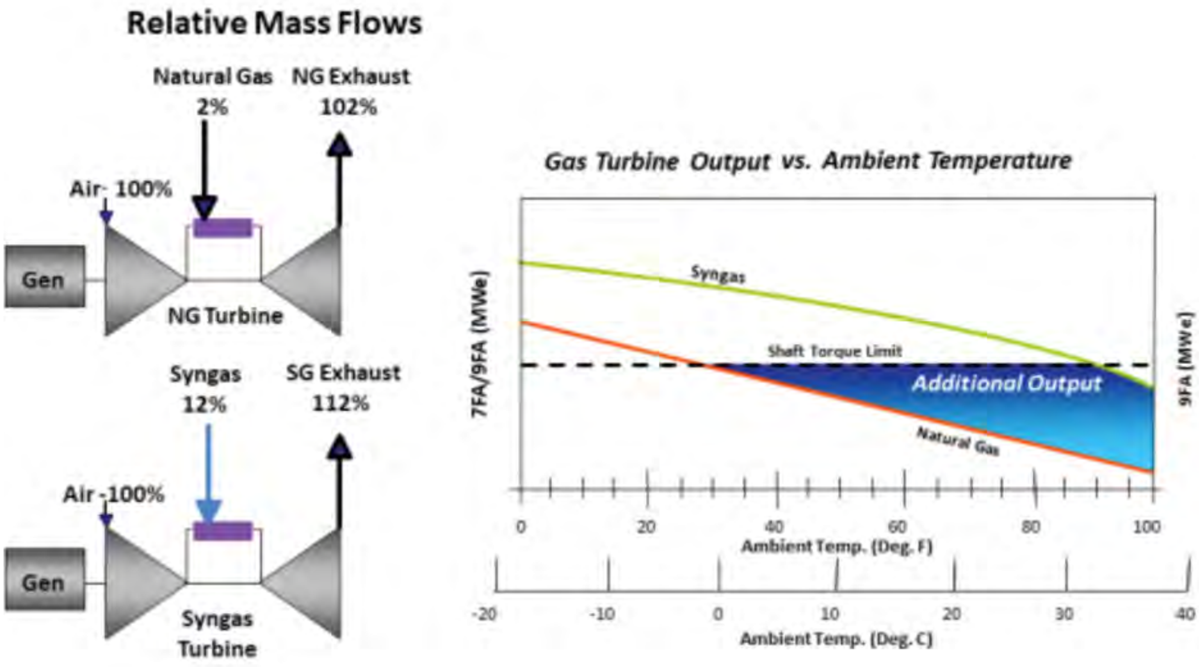
Figure 7-6. Degradation of a turbine output and heat rates with ambient temperature



Used with permission from GE [20]

Figure 7-7. Degradation of turbine output with altitude

The output advantage of syngas is shown in Figure 7-8.[21] With natural gas, the fuel mass flow rate is on the order of 2% of the compressor air mass flow rate. For syngas, the turbine combustor still demands the same total energy input rate (LHV Btu/hour), but with its lower specific heat, the required syngas flow to the compressor will be greater than that of natural gas. The result is an approximately 10% net mass flow increase through the turbine's expander. This extends the "knee" (i.e., the temperature at which the output begins to fall below the torque limit) to higher temperatures. The knee can be extended to even higher temperatures with air and nitrogen integration with the ASU. One caveat is that the precise location of the "knee" may be determined by turbine and other plant limitations, such as cooling capacity for the steam cycle and generators, discharge temperature limits of compressor materials, and bucket and vane aeromechanics. Additional options for output enhancement are discussed in Section 7.9.



Used with permission from GE [21]

Figure 7-8. Maximum turbine output maintained at high ambient temperature

7.3.3 TURBINE FIRING TEMPERATURES

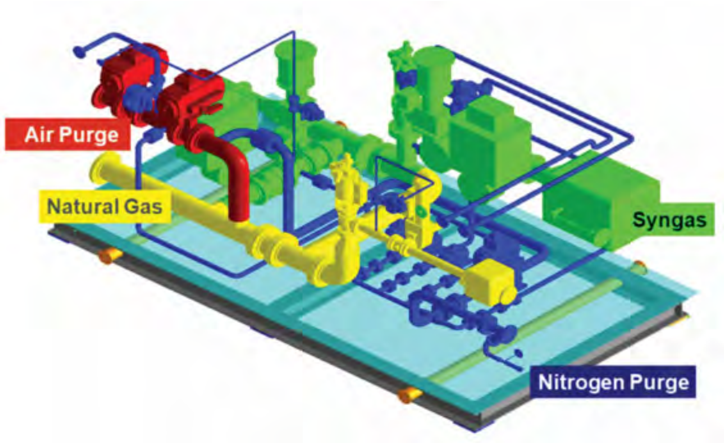
Due to the higher syngas mass input and higher moisture content from the H_2 content, turbine hot gas path components are subject to higher aerodynamic loadings with increased heat transfer, which impacts nozzle and bucket temperatures. If steam is used as a diluent, there will be additional heat transfer enhancement that must be considered in hot gas path thermal management. To preserve component life and maintenance intervals equivalent to natural gas operation, turbine manufacturers commonly reduce firing temperatures[†] by as much as 50°F.

7.3.4 TURBINE MODIFICATIONS FOR SYNGAS

Syngas turbines are modified versions of natural gas turbines. A primary adaption for syngas is the replacement of natural gas combustors with syngas diffusion flame combustors. Syngas combustors can accommodate nitrogen (N_2) or steam diluent injection for NO_X control and dual-fuel capability. Other design modifications include an upgraded main shaft for higher torque; larger 1st stage nozzle throat area; larger 3rd stage buckets; and an exhaust diffuser having aerodynamically and structurally improved struts, higher expansion ratio for increased total pressure recovery, and reduced turbine back pressure.

7.3.5 SYNGAS FUEL SUPPLY AND CONTROL

The combination of the combustor and the turbine fuel control system determines a turbine's ability to respond to changes in electrical load demand (e.g., turndown, ramp up, shutdown) and syngas production from the gasification block. Due to the hydrogen content of syngas, it is necessary that syngas turbines be started and shutdown on an alternative fuel. With a dual-fuel combustor system, this can be either natural gas or distillate oil.[‡] Typically, backup fuel is used through full-speed, no-load (FSNL), grid synchronization, and up to 50 to 70% of full load when transfer to full syngas operation can proceed. This dual-fuel capability is provided by a modular fuel control skid (Figure 7-9).[22] When nitrogen from the ASU is required as a diluent for NO_X reduction, a nitrogen control skid is also required. Compressor pressure ratio determines fuel pressure delivered to the combustor. Fuel and diluent pressure drop across the gas pre-treatment filters; control and transfer valves of the fuel control system; and gas pre-treatment, supply manifolds and combustor pressure drop must be accounted for when determining the minimum syngas and natural gas supply pressures.



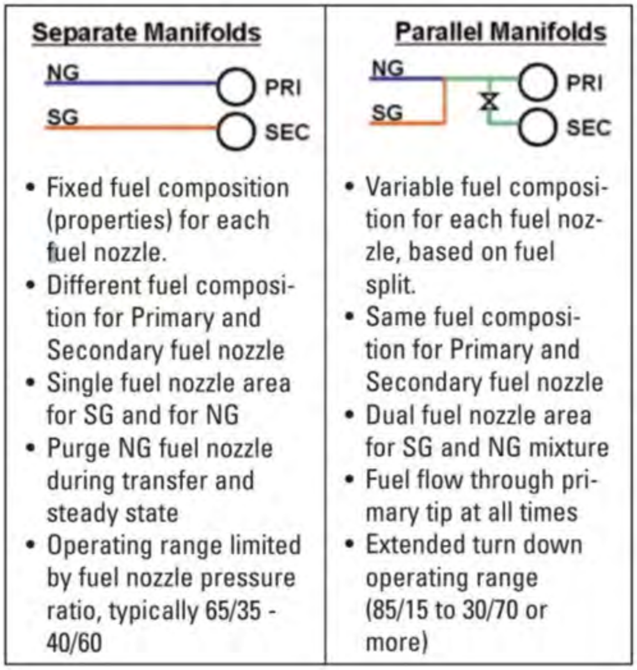
Used with permission from GE[22]

Figure 7-9. Dual-fuel control skid for extended turndown

Dual-fuel systems for syngas turbines are designed to operate over the entire load range, including 100% of either fuel. Depending on the quantity of syngas available, the unit may also be operated in a variety of fuel splits. Fuel systems can be broadly categorized into two classes—separate and parallel manifold (Figure 7-10). (Note that these are gross simplifications of the actual systems, which will require additional valves and manifolds to provide for decisive fuel transfers, purge of fuel lines and nozzle tips, controls, and shutdown procedures). The separate manifold system is limited by fixed nozzle sizing for each fuel, while the parallel manifold system provides for a greater range of operability.

[†] Firing temperature is at the exit of the turbine combustor while turbine inlet temperature (TIT) is at the exit of the first stage nozzles.

[‡] Ultra-low sulfur diesel (Number 2 fuel oil at 0.015 wt% S) is the preferred liquid fuel with special provisions required to accommodate high-sulfur liquid fuels.

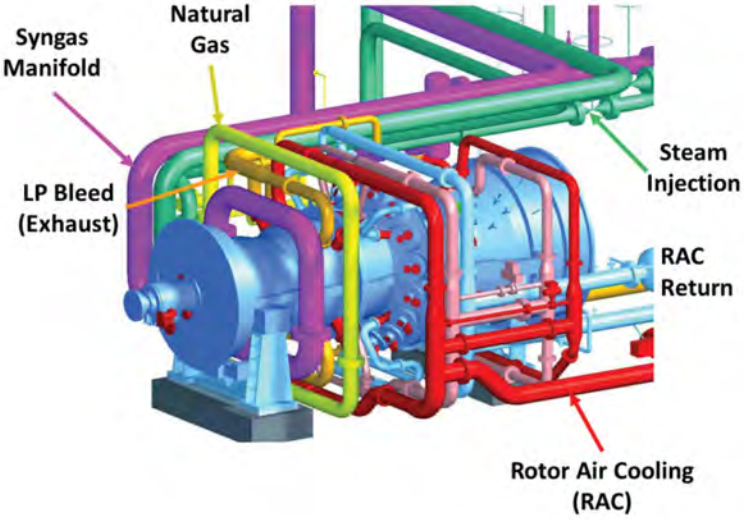


Separate Manifolds	Parallel Manifolds
• Fixed fuel composition (properties) for each fuel nozzle. • Different fuel composition for Primary and Secondary fuel nozzle • Single fuel nozzle area for SG and for NG • Purge NG fuel nozzle during transfer and steady state • Operating range limited by fuel nozzle pressure ratio, typically 65/35 - 40/60	• Variable fuel composition for each fuel nozzle, based on fuel split. • Same fuel composition for Primary and Secondary fuel nozzle • Dual fuel nozzle area for SG and NG mixture • Fuel flow through primary tip at all times • Extended turn down operating range (85/15 to 30/70 or more)

Used with permission from GE [22]

Figure 7-10. Simplified dual-fuel syngas control systems

Among the design considerations specific to syngas and other low-Btu fuels is the sizing of manifolds, piping, purge lines, valves, and control components to process fuel volumetric flows up to 4 to 5 times larger than natural gas. For large gas turbines, commercial fuel valves that are large enough to handle 100% of the fuel flow for the required service may not be available and require parallel valve arrangements. Also, the size and arrangement of manifolds must provide a balanced fuel distribution to the combustors with reasonable pressure drops while continuing to preserve accessibility to the turbine for inspection and maintenance. Diluent lines and, compressor discharge extraction lines (when gas turbine/ASU integration is required) must be also accommodated. Fuel, steam injection, and compressor air extraction manifold arrangement for a Siemens SGT6-5000F syngas turbine is shown in Figure 7-11.[23] The tightness of the arrangement can be compounded by extremely low syngas heating values from air blown gasification (around 120 Btu/scf), high fuel temperature (1,000°F), and the extraction of air supply to the gasifier. Dual-fuel capability allows the continuation of power and steam supply in instances of syngas curtailment and for startup and shutdown procedures (as necessary).



Used with permission from GE [23]

Figure 7-11. Manifold arrangement for a Siemens SGT6-5000F syngas turbine

7.3.6 SYNGAS CONTAMINANTS

The following list of syngas contaminants are of concern for the turbine:

- **Alkali Metals**—Sodium and potassium sulfates formed from combustion can corrode and foul the metals and coatings of combustion liners, transition pieces, and airfoils.
- **Particulates**—Particulates can erode airfoils and thermal barrier coatings and form deposits that plug aspiration cooling holes in nozzles, shrouds, and tips. In severe cases, erosion can cause rotating component liberation.
- **Condensables**—High molecular weight hydrocarbons present as droplets will raise the specific heating value higher than the fuel control system allows. The result can be burnout of combustor nozzles and temperatures exceeding material limits throughout the hot gas path.[§] Droplets can also contribute to first-stage turbine blade erosion.

The specifications for allowable contaminant levels are provided by turbine manufacturers. Those from GE are shown in Table 7-4.[24] Even though the gasification cleanup processes, as detailed in Chapter 6, must achieve these levels, additional provisions installed prior to the fuel control skid are recommended for turbine protection. An example[25] of a gas pre-conditioning system is shown in Figure 7-12.

1. **Preheater**—Required to prevent methane hydrate formation across the downstream pressure control valve and remove potential condensed liquids.
2. **Pressure Regulator**—Regulates the pressure of the gas whenever the supply pressure exceeds the maximum allowable pressure of the gas turbine fuel control system (see Section 7.8.1 on load following.)
3. **Filtration**—Dual coalescing filters—typically capable of removing 99.9% of all particles that are 0.3 microns or larger. Flow to filter canisters can be switched for replacement or purged of liquid and cleaned.
4. **Fuel Gas Superheater**—Typically designed for 50°F of input fuel superheating to eliminate condensed fuel components (or as required in the turbine manufacturer fuel specification).

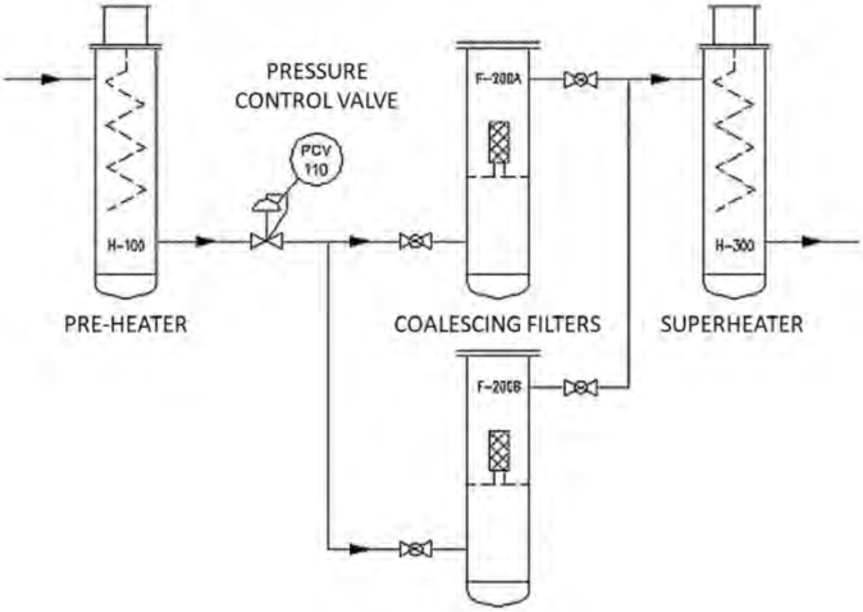
Table 7-4. GE Specifications for turbine syngas contaminant levels[24]

TRACE METALS	TURBINE INLET LIMIT, Xe ppbw		FUEL EQUIVALENT LIMIT, XFe ppmw			
	MACHINE MODEL		MACHINE MODEL			
	MS3000 MS5000 B, E, AND F-CLASS	6C, FB, H-CLASS	MS3000 MS5000 B, E, AND F-CLASS			6C, FB, H-CLASS
			TURBINE INLET FLOW/FUEL FLOW (E/F)			
			50	12	4	
Lead (Pb)	20	20	1.00	0.240	0.080	The limits for Pb, V, Ca, Mg for FB, H-Class machines are identical to the limits for other machines. Consult GE for alkali metal (Na and K) limits.
Vanadium (V)	10	10	0.5	0.120	0.040	
Calcium (Ca)	40	40	2.0	0.480	0.160	
Magnesium (Mg)	40	40	2.0	0.480	0.160	
SODIUM + POTASSIUM						
(Na/K)=28	20	3	1.00	0.24	0.080	
(Na/K)=3	10	3	0.50	0.12	0.040	
(Na/K)=<1	6	3	0.30	0.072	0.024	
PARTICULATES						
Total	600	400	30	7.2	2.4	Consult GE for particulate limits in 6C, FB, H-Class machines.
Above 10 microns	6	4	0.3	0.072	0.024	
Liquids: No liquids allowed; gas must be superheated.						

All piping and hardware downstream of the fuel conditioning skid should be stainless steel to prevent flaking of rust or corrosion entering the fuel control skid. In the process design phase, probabilistic pressure drop analysis is required of all process and fuel treatment components from gasifier to turbine combustors (including pre-fuel filtration and fuel control skid) to determine the gasifier's minimum operating pressure for all operating conditions.

---

§ Turbine control systems will include monitoring arrays of thermocouples at the turbine exhaust plane to detect over-fire conditions or combustor flameouts. Even if damage is prevented, these events can cause undesirable trips and power disruption.



Used with permission from GE[25]

Figure 7-12. Fuel gas pre-conditioning system

7.4 ENVIRONMENTAL

7.4.1 REGULATORY

A Title V Air Emission Permit issued by the U.S. Environmental Protection Agency (EPA) is necessary for a modular gasification power plant. If the plant is in a region with ambient air quality not in compliance with the National Ambient Air Quality Standards (NAAQS) established by the Clean Air Act (CAA) it is designated as a “Non-Attainment Zone.”[26][¶] In this case, stricter requirements are specified for emissions reduction technologies.

7.4.2 CRITERIA POLLUTANTS

Turbine exhaust will contain criteria air pollutants, such as NO_X, CO, ozone, PM, VOCs, sulfur dioxide (SO_2), and sulfuric acid (H_2SO_4). Those that are primarily produced by the turbine are NO_X, CO, PM, and VOCs. Modular gasification plants will have annual Potential-to-Emit (PTE) quantities that require New Source Review (NSR) and Prevention of Significant Degradation (PSD) review. PSD applies to major new sources or major modifications of an existing source located in areas designated as attainment or unclassifiable with NAAQS. The determination of PSD compliance is generally a state government function. It is assumed in this handbook that the plant site is in an attainment area and that PSD requires Best Available Control Technology (BACT) plus compliance with New Source Performance Standards (NSPS). Applicable federal EPA NSPS standards for combined cycle gas turbines are defined in the Code of Federal Regulations (CFR) under 40 CFR Part 60.40Da as revised in 2012.[27] Under this rule, IGCC plants (40 CFR 60[28]) are subject to the Subpart Da rules that apply to electric utility steam generating units and specifically (Part 60.40a) for IGCC plants as follows:[27]

> *“The IGCC electric utility steam generating unit is capable of combusting more than 73 MW (250 MMBtu/hour) heat input of fossil fuel (either alone or in combination with any other fuel) in the combustion turbine engine and associated heat recovery steam generator.”*

Further clarification is provided under Part 60.41Da Definitions:[29]

> *“Integrated gasification combined cycle electric utility steam generating unit or IGCC electric utility steam generating unit means an electric utility combined cycle gas turbine that is designed to burn fuels containing 50% (by heat input) or more solid-derived fuel not meeting the definition of natural gas.”*

For the calculation of thermal input-based emissions (PM and metallic hazardous air pollutants [HAPs]), in 2007, EPA clarified that this be based on heat input to the gas turbine and not the heat input of feedstock to the gasification block (as for combustion coal plants).

EPA rule’s current criteria pollutant standards for combined cycle gas turbine performance are included in Table 7-5. These regulations only apply to the combined cycle gas turbine; they do not apply to the gasification portion of the plant, even though a significant portion of the steam used to produce electricity in the steam turbine can come from syngas coolers. Further guidance is also given in the rule for the calculation of gross output for co-production and CHP. More stringent permit requirements may be applied from local air quality assessment, additional impacts, and public comment.

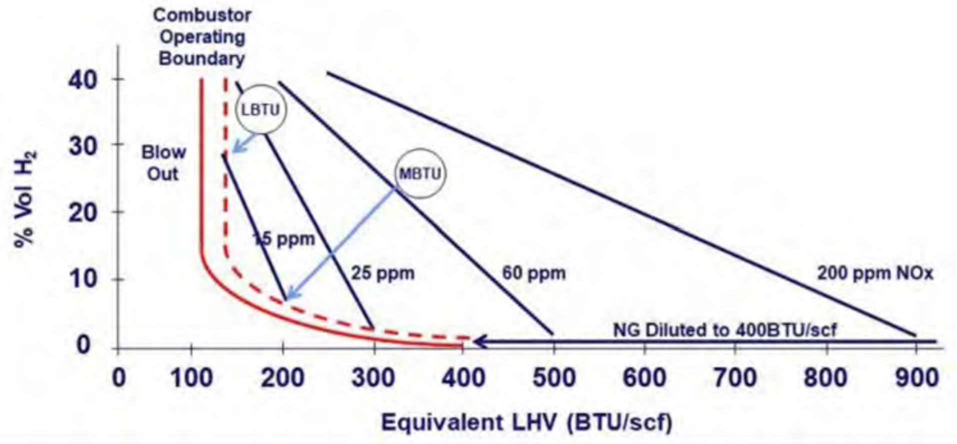
[¶] Details for individual pollutants are available at *https://www.epa.gov/criteria-air-pollutants/naaqs-table.*

Table 7-5. Standards for IGCC combusting more than 250 MMBtu/hour in turbine and HRSG commencing construction on or after May 3, 2011[30]

POLLUTANT	EMISSION LIMIT	BASIS	NOTES
NO_X	88 ng/J (0.70 lb/MWh)	Gross output	3.0-day rolling average
	95 ng/J (0.76 lb/MWh)	Net output	
	… or alternative		
NO_X+CO	140 ng/J (1.1 lb/MWh)	Gross output	
	150 ng/J (1.2 lb/MWh)	Net output	
PM Filterable or	0.04 lb/MMBtu (0.4 lb/MWh)	Gas turbine input	Defined test methods and frequency
Total non-Hg HAP metals	0.4 lb/GWh		

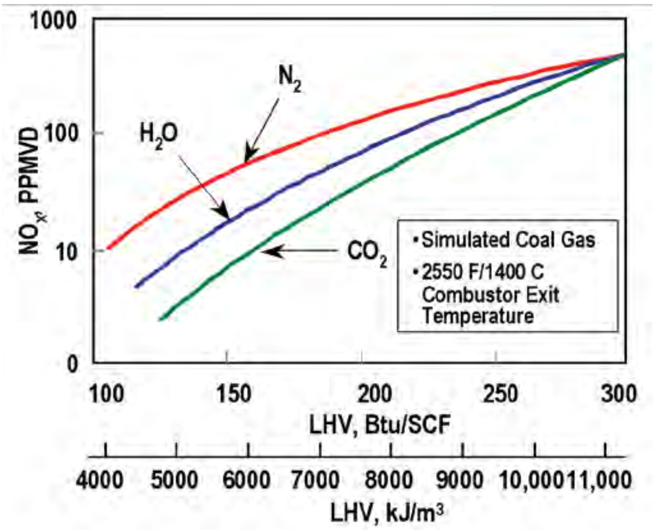
7.4.3 BEST AVAILABLE CONTROL TECHNOLOGY

For projects in attainment regions, diffusion combustors with diluent NO_X control have generally been accepted as BACT and suitable for air permits.[**] The operating map for a diffusion combustor demonstrating the effectiveness of dilution to reduce combustion NO_X is shown in Figure 7-13 for varying hydrogen concentrations.[31] Other diluents that are effective for NO_X reduction are water and carbon dioxide (CO_2) (Figure 7-14).[22] With oxygen-blown gasification having available high-pressure nitrogen, injection also increases turbine power output (see Section 7.9.1).



Used with permission from GE[31]

Figure 7-13. Diffusion combustor performance with hydrogen (stability range increases with nitrogen diluent)

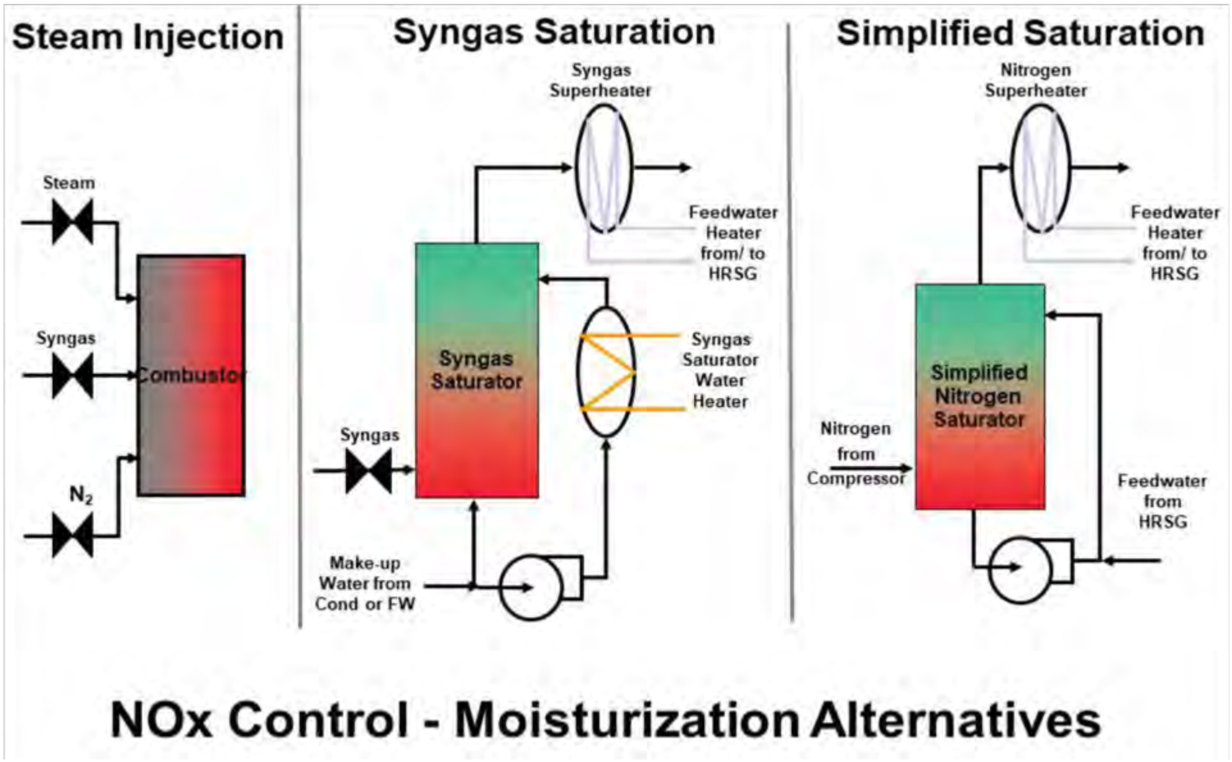


Used with permission from GE [22]

Figure 7-14. Effectiveness of diluents for NOX control

[**] For non-attainment regions, selective catalytic reduction (SCR) may be required for Lowest Achievable Emissions Rate (LAER). For ammonia-based SCR with sulfur bearing fuels, the production of ammonium salts can foul HRSG heat transfer areas, requiring care in design for cleaning.

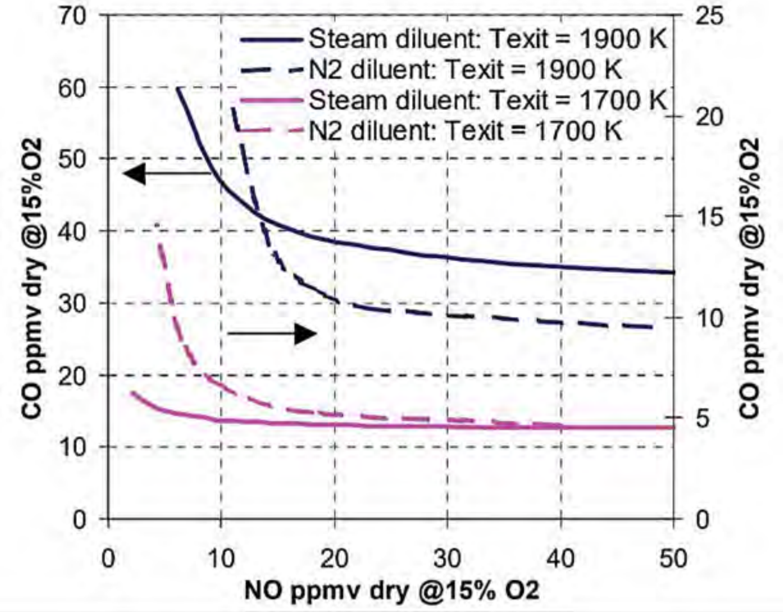
For air-blown gasification where nitrogen for diluent is not available, steam injection and/or fuel moisturization using a process gas saturator (Figure 7-15) are also effective in reducing NO_X. The steam can be injected directly into the combustor or the fuel can be moisturized. The choice will depend on project-specific factors, including the sensible heat available in the fuel, syngas properties, and the availability of steam at suitable superheat and pressure.



Used with permission from GE[22]

Figure 7-15. NOX reduction with moisturization

Carbon monoxide is not specifically defined as a HAP by the EPA, but it can be combined with NO_X for compliance (Table 7-5). To improve turbine performance, turbine manufacturers continue to increase turbine firing temperatures and expander pressure ratios. Higher temperatures increase the kinetics of NO_X formation. The challenge to the combustor designer is that flame temperature reduction comes with a reduction of combustion efficiency and a corresponding increase in unburned hydrocarbons (i.e., VOCs) and CO. This tradeoff is shown in Figure 7-16[32] for diffusion combustion of syngas. (Note that there is a critical point beyond which increasing diluent to reduce NO_X causes CO emissions to exponentially rise.) In practice, turbine control systems are tuned to maintain constant NO_X and CO compliance. More recently, adaptive controls using learning algorithms and feedback of combustor dynamics, real-time fuel property data, and stack monitors can maximize turbine performance while maintaining compliance.



Used with permission from GE[32]

Figure 7-16. Turbine inlet temperature tradeoff between NOX and CO

Choosing diluent feed rate and firing temperature to achieve 15 parts per million by volume (ppmv) NO_X and 25 ppmv CO for normal operation with syngas in an F-class combined cycle power block would result in approximately 0.54 lb/megawatts-hour (MWh) NO_X and 0.33lb/MWh CO (based on power output) for a combined NO_X and CO emissions rate of 0.87 lb/MWh (estimated for a plant heat rate of 8,500 Btu/kWh higher heating value [HHV].) Either alternative (NO_X or NO_X+CO) meets EPA's criteria pollutant standards given in Table 7-2. Since NO_X and CO are output based, raising efficiency is also an effective approach for compliance.

7.4.4 STARTUP/SHUTDOWN EMISSIONS

The startup of the gas turbine precedes syngas production and warmup of the HRSG for steam generation, during which neither nitrogen nor steam is available as a diluent. When the combustion turbine reaches FSNL, it is synchronized with the electrical grid and the main breaker is closed. Shortly after the gas turbine is synchronized it is loaded to a minimum—either to meet plant demand, including the ASU compressors, or to qualify as spinning reserve load. After warmup of the HRSG, steam can be used as diluent for emissions control and switched to nitrogen after startup of the ASU. During this initial phase of startup, NO_X emissions will be greater than that for normal operation. For the facility permit application, an estimate is required of total annual emissions on a rolling (tons/year) basis that includes startups, shutdowns, and trips. This requires a prediction of the number of annual startups, their length, and their contribution to annual mass emissions. The emissions for startup and shutdown events must be evaluated as part of a detailed modeling analysis and based on data and guarantees provided by the turbine manufacturer.

7.5 OTHER PERMITS

The responsibility for obtaining permits varies by permit type as agreed upon between an owner and the contractor building the project under an Engineering, Procurement, and Construction (EPC) Agreement. Generally, air and water permits are subcontracted in turn to environmental engineering groups with the necessary experience and specialized knowledge of current codes and requirements.

Federal

- Federal Energy Regulatory Commission (FERC)—designation as a Qualifying Facility (QF) or Electric Wholesale Generator (EWG) certification.
- Federal Aviation Administration (FAA)—oversight on stack and flare tower heights.
- U.S. Army Corps of Engineers (USACE)—any water intake structures from rivers considered to be navigable waterways; the definition of a navigable waterway includes almost any stream, creek, or river.

State and Local

- Building Permit.
- Certificate of Public Convenience and Necessity.
- Oil and or liquified natural gas (LNG) tank storage permits for dual-fuel capable turbines.
- Building and stack height variances.
- Construction variances for setbacks and easements.
- Certificate of Occupancy for building structures.

7.6 COMBINED CYCLE – CHOICE, CONFIGURATIONS, AND PERFORMANCE

The economics of gas turbines for power and process depends on the effective use of the exhaust energy—generally 60 to 70% of input fuel energy. In addition to reducing output-based emission requirements, total plant cost reduction ($/kilowatt-electric [kWe]) provides compelling justification for adopting a combined cycle configuration. The addition of a steam turbine bottoming cycle increases net power output by roughly 50%. Even though the total installed cost of the HRSG, steam turbine, condensers, water cooling, and cooling water treatment (such as deaeration or phosphate removal to prevent damage to the cooling system) can be more than the turbine cost,[33] the additional power will reduce the total plant per MW cost by 40 to 50%. A bottoming cycle can also be tapped to provide steam for other processes. Polygeneration or CHP applications significantly increase the overall input heat utilization factor for the plant output basis emissions criteria. In the absence of a bottoming cycle, there is a significant amount of energy otherwise wasted due to the high quantity (i.e., turbines use a lot of excess air) and high temperature of exhaust gases. Compared to simple cycles, a combined cycle will provide about a 10% reduction in heat rate for reductions in fuel cost and stack emissions.

7.6.1 STEAM TURBINES

Combined cycle plants are generally multi-shaft (i.e., gas and steam turbines having separate shafts and generators). As described in Section 7.4.4, this allows the gas turbine to proceed through cold startup and grid synchronization while steam generation from the HRSG and condenser and cooling water circuits ramp up. The startup of the steam turbine will lag that of the gas turbine (see Section 7.8).

The choice of type and configuration (e.g., number, size, and steam requirements) for the steam turbine generator (STG) depends on the steam quantity and maximum temperature obtainable from the HRSG, plus accommodation of heat recovery steam from the gasifier. Extraction may be required from the steam turbine to the gasifier, including regeneration of acid gas removal (AGR) solvent or the gas turbine for NO_X control and/or power output enhancement. Generally, a reheat, condensing steam turbine is chosen that includes an integral high-pressure (HP)/intermediate-pressure (IP) opposed flow section and, depending on the output capacity of the gas turbine, a double flow low-pressure (LP) section with last stage buckets suitable for low condenser pressures. However, if it is intended that the gasification plant provide process (cogeneration) or space heating (CHP) steam, the STG LP sections may be reduced in size or eliminated. A reheat design assures high thermal efficiency and excellent reliability based on a large power experience base. The steam turbine will generally be coupled to a hydrogen-cooled generator, as is the gas turbine(s).

7.6.2 HEAT RECOVERY STEAM GENERATORS

Steam and gas cycle matching is accomplished at the HRSG. HRSGs are provided by several qualified manufacturers. Modular HRSGs consist of four major components: the evaporator, superheater, economizer, and water preheater sections. These are arranged to meet the operating requirements of the steam bottoming cycle and process or heating extraction. When a gasifier with heat recovery options is included, the radiant and/or convective syngas coolers can be used in lieu of the HRSG's HP/IP evaporator(s) to vaporize a portion of the incoming saturated liquid water from the HP/IP economizer section(s). The total superheated steam from the HP stage of the HRSG is fed to the HP stage of the steam turbine to generate power. A wide range of HRSG configurations are available to achieve optimal integration with the steam turbine and process steam demands. To provide the most economical and functional design, it is necessary to evaluate the following parameters:[34]

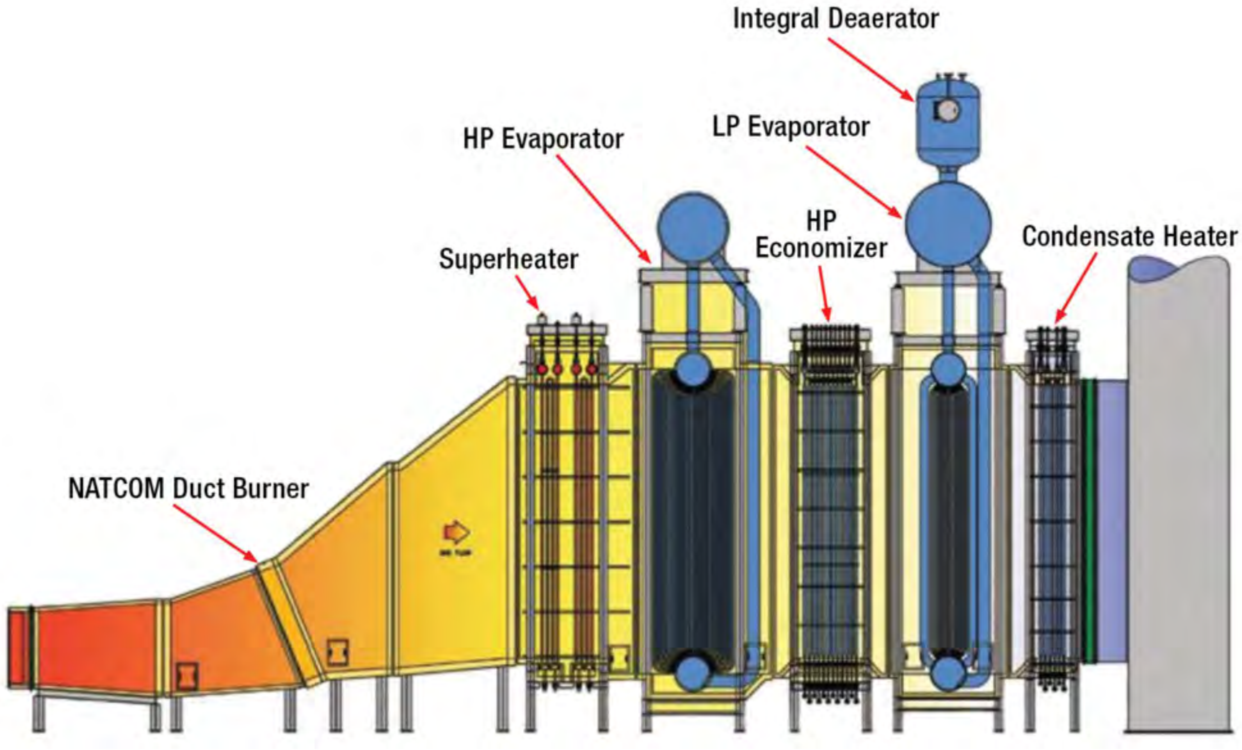
- Allowable gas turbine back pressure.
- Emissions control.
- Steam pressures and temperatures.
- Number of HRSG sections (e.g., superheater, reheat, and economizer).
- Overall pinch point (i.e., the minimum temperature difference between the gas and steam at either entrance/exit location of the HRSG).
- Steam turbine inlet temperatures and design mass flow rates.
- Superheater and economizer approach temperatures.
- Stack temperature.

Each of these parameters and their impact on performance and cost are provided in Table 7-6.

Table 7-6. HRSG design parameters and impact on the power block

KEY HRSG DESIGN FACTORS		
PARAMETER	**DESCRIPTION**	**IMPACT**
Back Pressure	Function of the HRSG cross-sectional area and number of stages	Higher back pressure reduces HRSG cost, but degrades gas turbine performance.
Emissions Control	Selective Catalytic Reduction (SCR) of NO_X	Conversion requires high temperatures. Catalyst and Ammonia/UREA injection locations within the HRSG are determined by catalyst temperature requirements.
Number of HRSG Sections	Determined by the choice of steam turbine and required SCR operating temperatures	Increased efficiency, but HRSG cost will increase with addition of ST stages.
Steam Pressures and Temperatures	Determined by the steam turbine admission requirements	Determines required stages and heat exchanger sizes and cost.
Pinch Point	Minimum temperature difference between exhaust gas and HRSG heat exchanger sections	Low pinch temperatures improve heat recovery, but have a significant impact on HRSG size and cost.
Economizer Approach	Temperature difference between feedwater to the HRSG and its exit gas temperature	Low economizer tube temperatures will cause condensation and corrosion.
HRSG Outlet Temperature	Exhaust temperature reporting to the stack	Minimum HRSG exit temperatures are required to ensure buoyancy and stack plume dispersion.

The HRSG is generally a horizontal gas flow, natural circulation, drum-type multi-pressure design (Figure 7-17) with separate heat transfer stages that are matched to the quantity and temperature characteristics of the combustion turbine exhaust gas. Depending on the turbine model and uprate status,[35] typical turbine exhaust temperatures can range from 850 to 1,150°F. For higher efficiency, the HRSG should be designed for reheating steam from the HP turbine and be able to incorporate steam and water generated by gasifier syngas coolers. When the steam from the HRSG is for a 3-stage steam turbine, the HRSG will have HP, IP, and LP streams with economizers, platens and separation drums, and superheaters for each to produce superheated steam at different temperatures and pressures. In addition to generating and superheating steam, the HRSG performs reheat duty for the cold/hot reheat steam for the steam turbine, accepts saturated steam from the gasifier, provides condensate and feedwater heating for the gasifier radiant cooler, and deaerates the condensate.



Permission pending from Cleaver-Brooks[36]

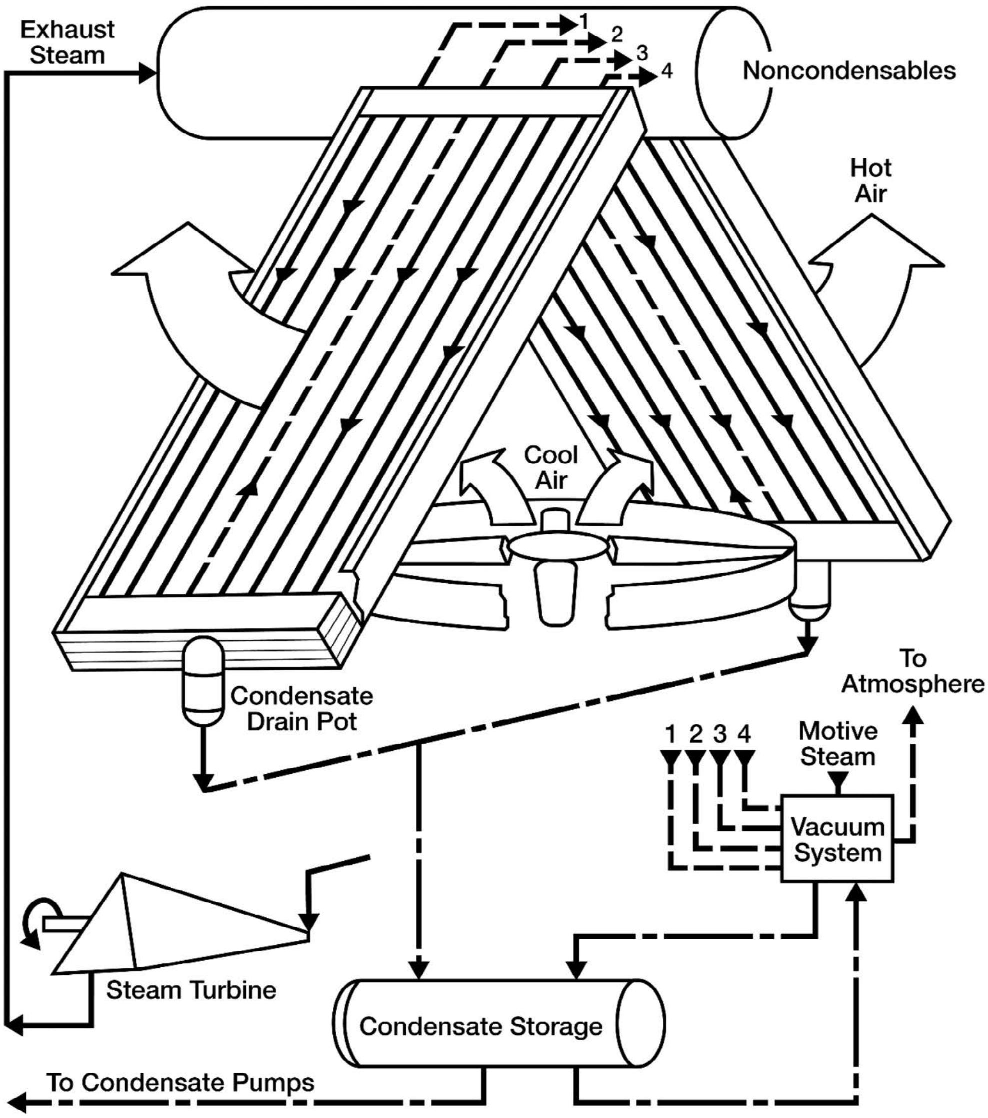
Figure 7-17. Configuration of a modular 3-pressure HRSG

For CHP and/or cogeneration applications, a portion of the steam from any or all the various pressure streams may be dedicated to supplying process steam for some end-use, such as space heating. Additional space should be reserved for potential selective catalytic reduction (SCR) installation. Sulfur (as hydrogen sulfide [H_2S]) in syngas will be converted to sulfur trioxide (SO_3) in the turbine combustor and react with ammonia for the SCR to form ammonium sulfate and bisulfite. This necessitates special design of the low-temperature sections to mitigate ammonium sulfate deposits and allow for ease of cleaning by washing or other means with provisions for collecting the resultant waste material.

7.6.3 CONDENSERS

The steam cycle's condenser receives exhaust from the low-pressure steam turbine stage and condenses it to liquid to be to be pumped at high pressure for return to the economizer section of the HRSG. The key functions of the condenser are to provide a low back pressure to the steam turbine to maximize power output/thermal efficiency, deaerate the condensate to minimize corrosion, and collect all condensate drains. Steam condensers have generally been of a shell and tube design with the turbine steam exhaust on the shell side and counter-flow cooling water on the tube side. This requires either a water source, such as a cooling pond, or the addition of either a cooling tower (e.g., evaporative natural draft/mechanical/forced draft) or direct air cooling. When water is scarce or not allowable due to regulatory water restrictions, air-cooled steam condensers (Figure 7-18) have proved effective. Air-cooled condensing is attractive for several reasons:

- Makeup water is not required, thereby permitting plant siting without large cooling water supply.
  This is important to the flexibility of modular gasification plants.
- Maintenance is less expensive.
- Water treatment chemicals are not required, and there are no problems associated with blowdown water disposal.
- Fogging, misting, and local icing resulting from wet cooling tower plumes are not encountered.



Used with permission from Babcock & Wilcox[34]

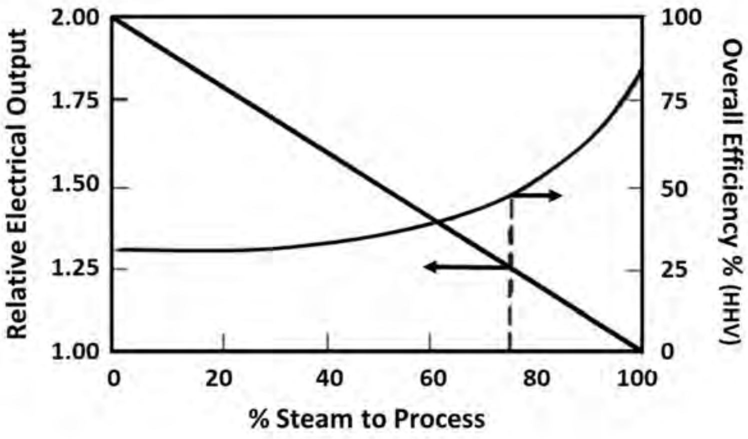
Figure 7-18. Single-stage, direct-forced, air-cooled steam condenser

A disadvantage of air-cooled systems is that the minimum temperature and corresponding steam condenser pressure are not as low as wet cooling power systems, which lowers STG efficiency. Air-cooled condensers also require a larger footprint and greater separation from other process equipment compared to wet shell-and-tube condensers.

Air-cooled condensers may be mechanical or natural draft. There are two types of available systems: direct and indirect using an intermediate water coolant loop. In the direct system, the turbine exhaust steam is cooled by the air, effectively combining the conventional condenser and cooling tower into one device. The system thermodynamic advantage is that only one heat exchanger temperature drop is required versus two in the indirect system. A disadvantage is the need for large ducts to carry the steam from the low-pressure turbine exhaust to the steam condenser. In the indirect system, a surface or spray steam absorber absorbs heat from the steam and a secondary water loop transfers the energy to a separate air cooler, which rejects the heat to the atmosphere. Expensive large diameter ducts are replaced by the water piping and an additional heat exchanger. A disadvantage is an additional thermal loss from the condenser pressure drop.

7.7 COGENERATION

For cogeneration with steam export to process or CHP, the accounting of energy results in an overall efficiency significantly higher than combined cycles that only produce power.[37] This is advantageous in the calculation of plant output-based criteria emissions. (Current EPA regulations (40 CFR Part D) uses 75% of cogeneration energy export for output-based emission rates.[38]) The effect on overall heat utilization efficiency (treating the HRSG as a boiler) of increasing the LP (150 pounds per square inch gauge [psig]) steam export that would otherwise be used in the LP turbine is depicted in Figure 7-19.



Basis:

1) Steam conditions 1450 psig, 950 F (101 bars, 510 C), 150 psig (11.4 bars) process, 2 1/2" (63.5mm) HgA condenser pressure
2) Three stages of feedwater heating
3) Boiler efficiency–85% HHV

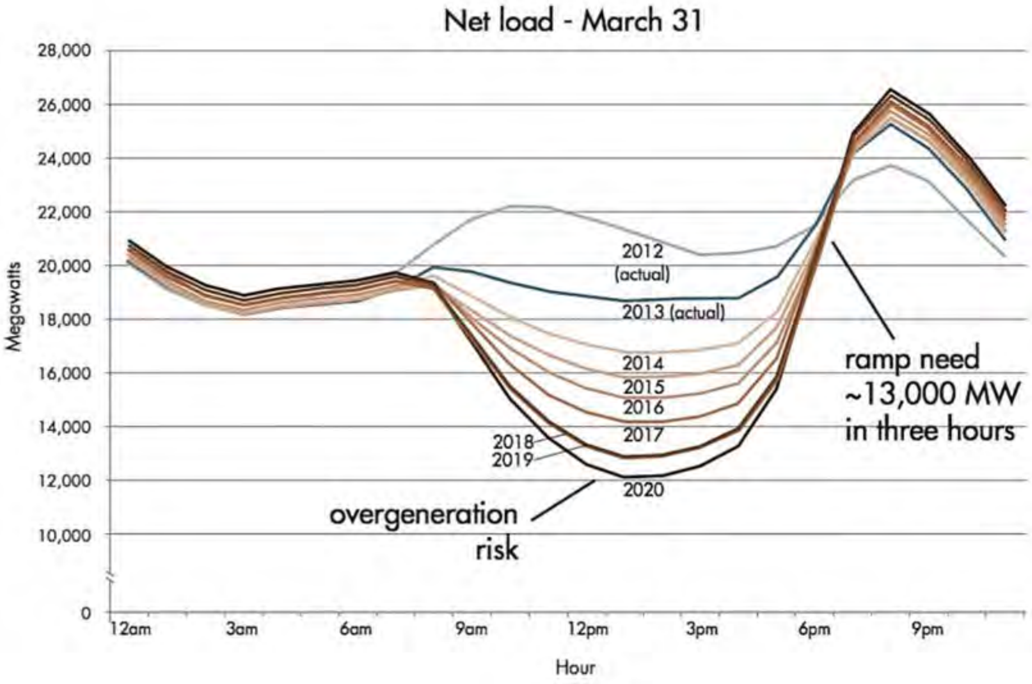
Used with permission from GE[37]

Figure 7-19. Heat utilization effectiveness using steam for cogeneration

As more LP steam is exported, the electrical output ratio (relative to the electrical output at zero steam-to-process [i.e., with power provided by the HP and IP turbines only]) decreases from a maximum of approximately 2.0 to 1.0 if all LP steam is delivered to process. With all LP steam exported, the relative power (1.0) is that generated by the HP and IP STG stages. However, the overall efficiency/performance increases from 35% with no export to 85% at full export. A GE MS7001EA gas turbine generator in a combined cycle providing 150 pounds per square inch gauge (psig; 10.3 bar) process steam can yield an overall energy effectiveness of nearly 75% on an HHV fuel input basis.

7.8 LOAD FOLLOWING

The increasing penetration of renewables into grids has become a challenge to Independent System Operators to meet hour-to-hour demand due to the variability of renewable supplies. New units will be expected to cycle from their first day of operation. For example, the hour-to-hour power demand is characterized by the "Duck" curve[39, 40] (Figure 7-20).



Used with permission from CAISO[39]

Figure 7-20. Daily "Duck Curve" load curve for California showing severe evening ramp requiring reserve power [40]

The increasing amount of renewable solar power generation in some locations has resulted in an ever-decreasing amount of electricity demand during the early afternoon hours, which results in renewable over-generation. This trend also leads to an increasingly steep ramping period during the late afternoon and early evening that renewables are unable to meet. To achieve this higher ramp rate, gas turbines are advantageous due to their agility, fast start, and rapid ramping capability. Due to low natural gas prices, the Energy Information Agency (EIA) estimates that the U.S. electric power sector has added, or plans to add, 11.4 GW worth of new gas turbine combined cycle power plants in 2019 and 2020.[41] From cold start, Aeros can reach full load in 10 minutes (10%/min ramp) and recent rapid response combined cycle plants can also ramp from hot start to 66% load in 10 minutes (6.6%/min) and to 100% load in 30 minutes (1.7%/min as governed by steam generation lag). This responsiveness allows gas turbine plants to participate in bidding for "real-time" (minutes) forecasts of demand and other system conditions. Immediate response requires that these plants be operated as "spinning reserve" (i.e., operating at low load while synchronized to the grid). Gasification power or cogeneration plants are challenged to participate in this market due to the responsiveness of the gasification plant and capital recovery that demands stable revenue. A gasification plant can still provide "day-ahead" capacity, which requires turndown and even nightly shutdown. Alternatively, a reserve of natural gas can be kept onsite to boost ramp rates while the gasification plant catches up (since the gas turbines used will have natural gas capability anyway).

7.8.1 CONTROL STRATEGIES

For gasification power plants, there are three main control strategies:

- **Gasifier Lead**—In response to a demand, output is controlled by adjusting the feed rate to the gasifier. The turbine control system responds by regulating the fuel flow to the gas turbine with the inlet fuel valve, which in turn regulates the gasifier pressure. The benefit is that the gasifier can be operated within recommended operating conditions for efficient carbon conversion and extended refractory life. Generally, this load-following control strategy is not employed when rapid load changes are anticipated.
- **Turbine Lead**—Plant output is controlled by varying the fuel flow to the gas turbine. Syngas pressure to maintain turbine fuel supply is then controlled by manipulating the feed rates into the gasifier. This allows for faster load changes by drawing down the syngas capacity in the syngas cleanup units. Also, when the turbine is operating in dual-fuel mode, transfer to higher backup fuel can close the deficit in syngas supply to the turbine. The disadvantage is the time lag in which the gasification and air supply plants need to ramp up syngas production. There also may be large swings in the AGR system's performance that degrades gas cleanup effectiveness and gasifier carbon conversion efficiency.
- **Coordinated Mode**—Changes in plant load are initiated by simultaneously adjusting the fuel flow to the gas turbine and the gasifier feed rate to maintain the required pressure. A feed-forward control signal is provided to the gasifier to adjust the fuel gas production rate in anticipation of a change in fuel plant pressure. Also, a feedback signal to the turbine fuel control limits the rate of change of syngas flow rate to avoid an excessive change in syngas pressure.

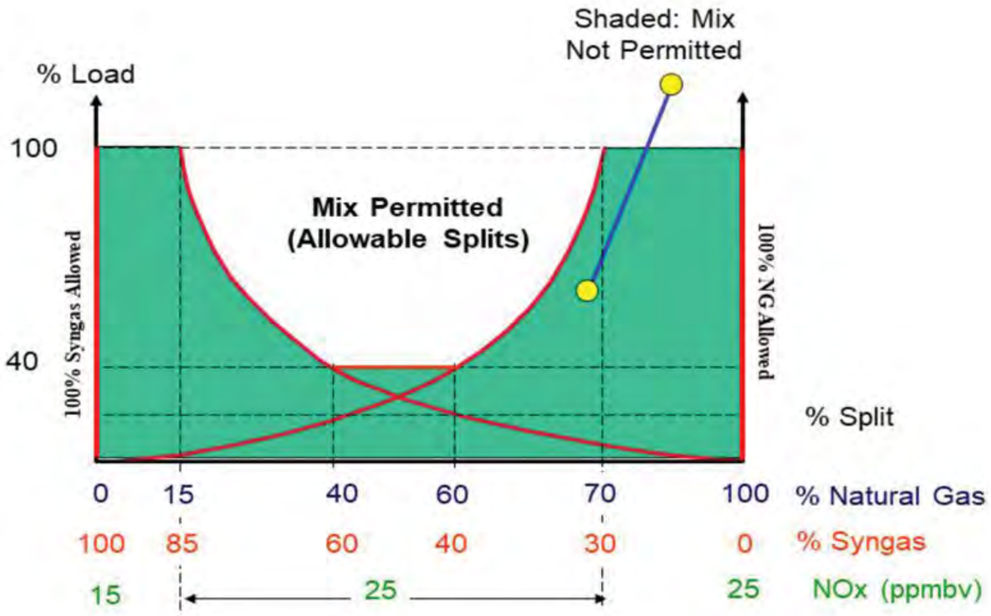
7.8.2 RAMP RATES

Due to the turbine's dual-fuel capability, power generation can begin while the gasifier block proceeds through its startup. Ramp rates are defined in terms of %load/min for turndown (load-shedding) or ramp up. These are important parameters in evaluating the capability of a plant to load follow. For rapid load increases, the turbine can generally consume more gas than can be immediately produced by the gasification plant. The same is true during load-shedding. Data from operating IGCC plants show ramp rates of 1.5 to 3%/min. The Negishi IGCC plant (Japan) has reported[42, 43] ramp up and turndown ramp rates of 3%/min at 75 to 100% load. The Cool Water IGCC plant (California, USA) conducted load shedding tests[44] from 100% to 50% load and back. The overall rate of change was 2.5%/minute for ramp up and 3.0%/min for downturn. Although these rates are slower than natural gas combined cycle (NGCC) plants, certain steps can be implemented during the design phase to improve the ramp rates for fuel supply.

- In anticipation of load changes, operate the turbine in dual-fuel mode and be ready to use backup fuel for load ramp up and decrease natural gas as syngas is available.
- Adopt similar features and control strategies from rapid start and fast response NGCC plants.
- Use a high-fidelity total plant simulator to optimize control design and control strategies in preparation for anticipated load following requirements.
- Include the ASU's ability to ramp the delivery of oxygen and nitrogen up or down as a key specification selection requirement and consider the storage of onsite liquid oxygen (LOX) to accelerate gasifier block startup.

7.8.3 DUAL-FUEL OPERATION

The turbine must use backup fuel during startup and shutdown. This capability can be expanded to continuously co-fire over a wide range to provide operational flexibility for load changes and fuel transfers. Allowable fuel splits and exclusion zones will depend on the fuel control skid's configuration, as well as control valve choice for the combustor (as necessary). A map of allowable fuel splits and output is shown in Figure 7-21 for a parallel manifold—dual-fuel system with natural gas as the secondary fuel.[22] Either 100% syngas or 100% natural gas firing is allowable.



Used with permission from GE[22]

Figure 7-21. The “Bathtub Curve” of allowable fuel split ranges for a dual-fuel control system

At full output, syngas/natural gas splits from 30 to 85% are allowable. The minimum load turndown is 40% along with a syngas range restricted from 40 to 60%. At low loads, the syngas/natural gas split is defined by the minimum allowable control valve pressure drops and combustor nozzle pressure ratios for the protection of the combustor from flame attachment. Similar operating maps will apply for the syngas turbines from other turbine manufacturers.

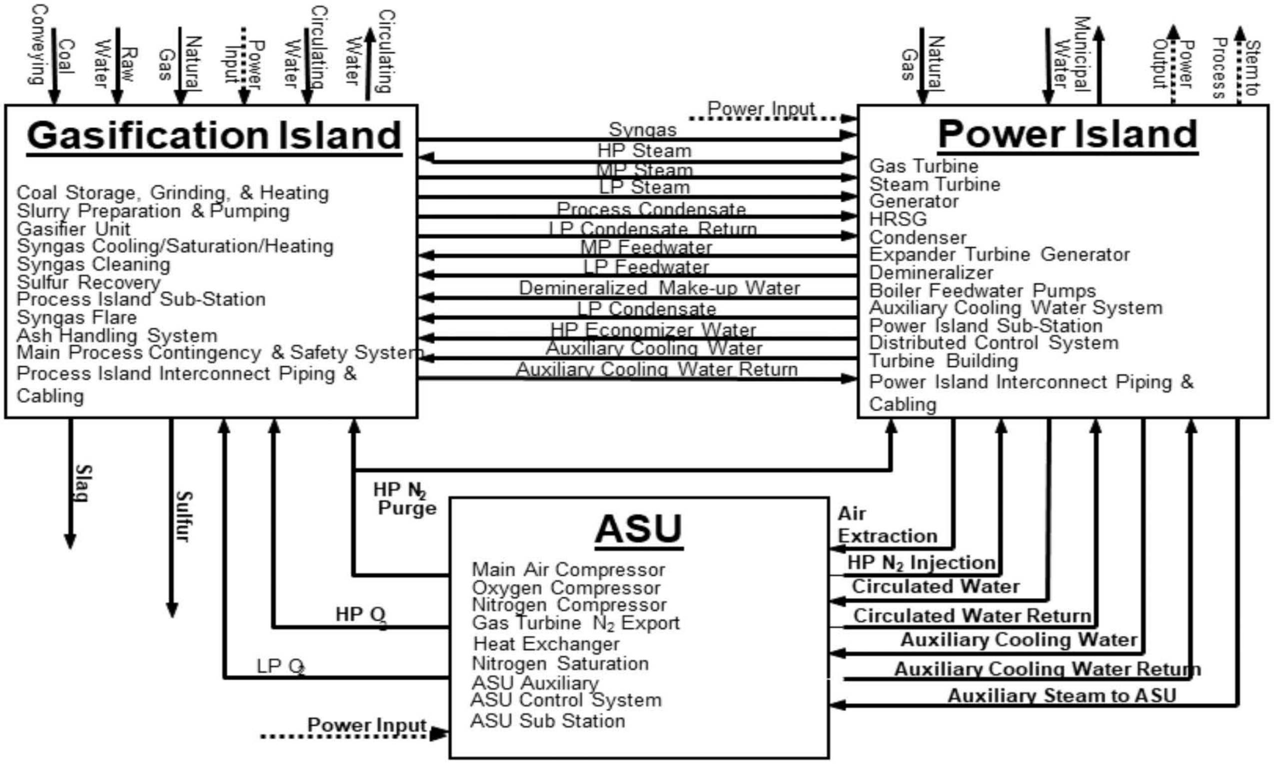
7.9 TURBINE/PLANT INTEGRATION

The level of integration of a modular plant’s components can be simple or complex. The primary goal of the engineering design process is to achieve the project’s financial performance requirements of cost (i.e., fixed and variable), schedule, thermodynamic performance (i.e., output, efficiency, and availability), smooth permitting, and operational flexibility. For example, a project’s capital investment strategy may have an over-the-fence division of responsibility for the delivery of oxygen and nitrogen. Another example is an adjoining refinery with the delivery of refinery residuals for gasification feed or supplementary refinery waste gas as monetary exchange for steam and electrical power.

7.9.1 INTEGRATION REQUIREMENTS AND OPTIONS

With the internal and external project requirements, an integration study can proceed through a value-added assessment of each of the options. The designer must choose each of the IGCC components to fit together and seamlessly operate during all modes of operation—startup, shutdown, and full- and part-load operation under hot and cold ambient conditions. Lumping components together and applying a master control system might not meet goals. This section discusses integration benefits that can inform decisions (i.e., what integration options are necessary, and what additional options can be implemented to improve performance, operability, and availability).

A block flow diagram of an integrated gasification power plant including flows in and out of the gasification, oxidant, and combined cycle blocks is shown in Figure 7-22.[17] In addition to the integration of turbines within the power block, further integration between the three major plant blocks is necessary.

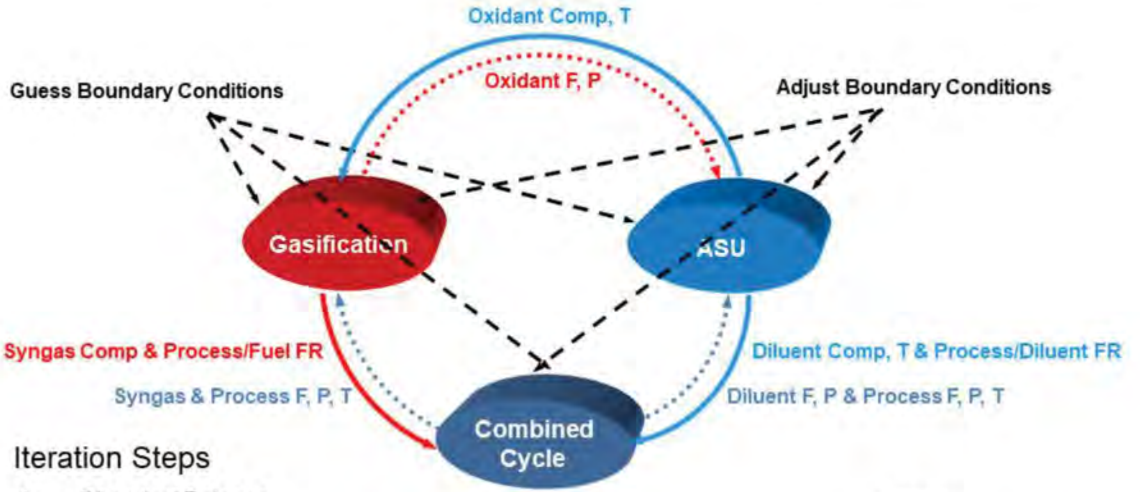


Used with permission from GE[17]

Figure 7-22. Block flow diagram for an integrated gasification power plant

The iterative process to close mass and thermal balance is shown in Figure 7-23.[45] Suitable tools for integration analysis are ASPEN Plus™ or HYSYS™, in conjunction with a combined cycle modeling tool, such as GTPro™. A plant heat and mass balance is also required to achieve closure of the steam and water balance (this is important at locations with limited water supply or when Zero Liquid Discharge [ZLD] is required). Both power and gasification islands have water treatment facilities (e.g., grey and black water from the gasifier, demineralization/deaeration, conditioning of condensate makeup supply for the power block). The water and steam transfers between the major plant blocks (i.e., power, ASU, and gasifier) include:

- Export of gasifier steam generation (IP and/or LP) to the steam turbine steam cycle.
- Export of HP and/or IP steam from the gas turbine HRSG to the gasifier for temperature modulation and, if required, syngas shift for CO_2 capture.
- Export of HP and LP feedwater and demineralized water from the turbine island to the gasifier.
- Export of high-pressure nitrogen from the ASU to the gas turbine for NO_X reduction and output enhancement and shutdown purge.
- Export and return of auxiliary cooling water between the turbine island, ASU, and gasifier.
- Power from the turbine island for plant auxiliaries, such as pumps, compressors, and fans. For a gasification plant, gasification-related auxiliaries will be approximately 20% of the gas turbine and steam turbine gross output.



Iteration Steps

1. Chemical Balance.
2. Mass and Energy Balance.
3. Pressure and Power Balance.
4. Repeat steps 1 – 3 until convergence and performance CTQ's are met.
5. Reliability and Economic Analysis.
6. Repeat steps 1 – 5 until CTQ's are fulfilled.

Keys:

Comp = Composition (Mol%)
F = Mass Flow Rate
P = Pressure
T = Temperature
FR = Mass Flow Ratio

Used with permission from GE[45]

Figure 7-23. Iterative process for closure of primary heat and mass balances

The ASU consumes approximately 15% of the gross plant output; to reduce consumption compressed air can be supplied to the ASU from the gas turbine (air-side integration [ASI]). Most HDGTs are configured for compressor air extraction to provide hot air back to the turbine air inlet to prevent icing. For ASI, GE and Siemens extract air from the shell of the syngas combustors (with a limitation of approximately 20% of the total turbine compressor air flow). The possible air integration levels with the performance impact for using turbine extraction with an ASU Main Air Compressor (MAC) are shown in Figure 7-24.[46] An evaluation of whether to use air extraction, or how much air extraction is desired, should include consideration of the following advantages and disadvantages:

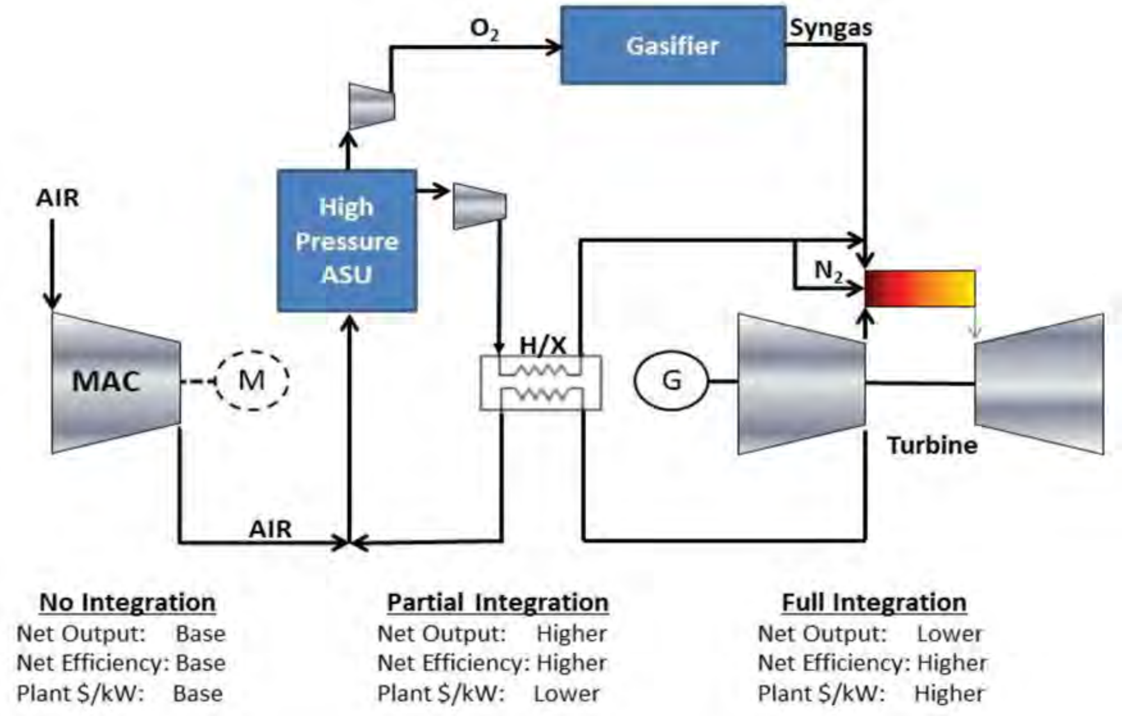
Advantages

- Reduced size, cost, and power consumption of the ASU MAC and potentially its elimination.
- Reduced parasitic load for increased plant efficiency.
- Increased plant output (determined by the level of ASU integration [Figure 7-24]).

Disadvantages

- Complexity of coordination between turbine and ASU during startups and load changes.
- Manifolding at turbine (difficult accessibility) and ducting of hot, high-pressure air from the turbine to the ASU island.

For turbine compressor pressure ratios less than 16 and integrated with a high-pressure ASU, an air boost compressor will be required.



Used with permission from GE[46]

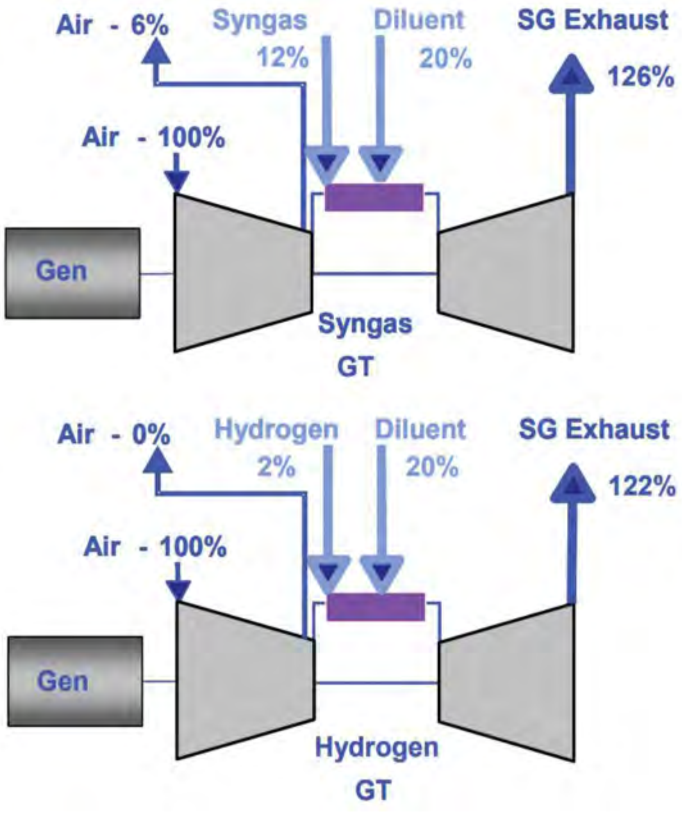
Figure 7-24. Air-side integration (ASI) between ASU and turbine

The degree of ASI in terms of ASU capacity can range from partial up to full when 100% of the ASU capacity is supplied by the gas turbine and the ASU MAC can be eliminated. Both full and partial air integration have been employed at European IGCC plants[47]; partial integration[48] has been employed at the Edwardsport IGCC plant (Indiana, USA). However, the European plant having full integration necessitated several control modifications over a period of years to address reliability issues. Due to this experience, it is recommended that a dynamic model of the total plant with its power block be developed as part of the design phase to ensure that all operational and transient modes can be seamlessly achieved without upsets, trips, or harm to the plant.

7.9.2 PERFORMANCE IMPACT OF AIR-SIDE INTEGRATION

Due to its lower mass heating value, syngas accounts for 12% of the total turbine mass flow rate compared with 2% with natural gas, and nitrogen diluent accounts for an additional 20% (compare Figure 7-25[49] with Figure 7-8). After accounting for 6% air extraction (approximately 25% of ASU demand), the result is a net 24% additional mass flow through the turbine compared to a natural gas turbine. For an IGCC carbon capture and storage (CCS) plant where the H_2 concentration in the syngas is higher, the fuel flow is more like natural gas (1 to 2%). For this case with zero air extraction, the result is a net 20% additional mass flow through the turbine as compared to natural gas. Air extraction could be advantageous for hydrogen due to the diluent.

At temperatures below the “knee” (where the output/ambient temperature curve falls below the shaft torque limit [Figure 7-26]), compressor inlet guide vanes (IGVs) are closed to maintain the torque below the shaft limit. In this area of IGV control, air extraction can be taken without reducing net power output. The extraction helps to mitigate other turbine mechanical limits and reduce ASU power requirements. As the knee is approached, compressor IGVs are opened to full (i.e., 100%) at and beyond the knee. With the combination of air extraction, higher syngas mass flow and diluent injection, the knee can be extended from near-iso conditions to as high as 80 to 95°F ambient.



Used with permission from GE[49]

Figure 7-25. Impact of air-side integration on turbine output

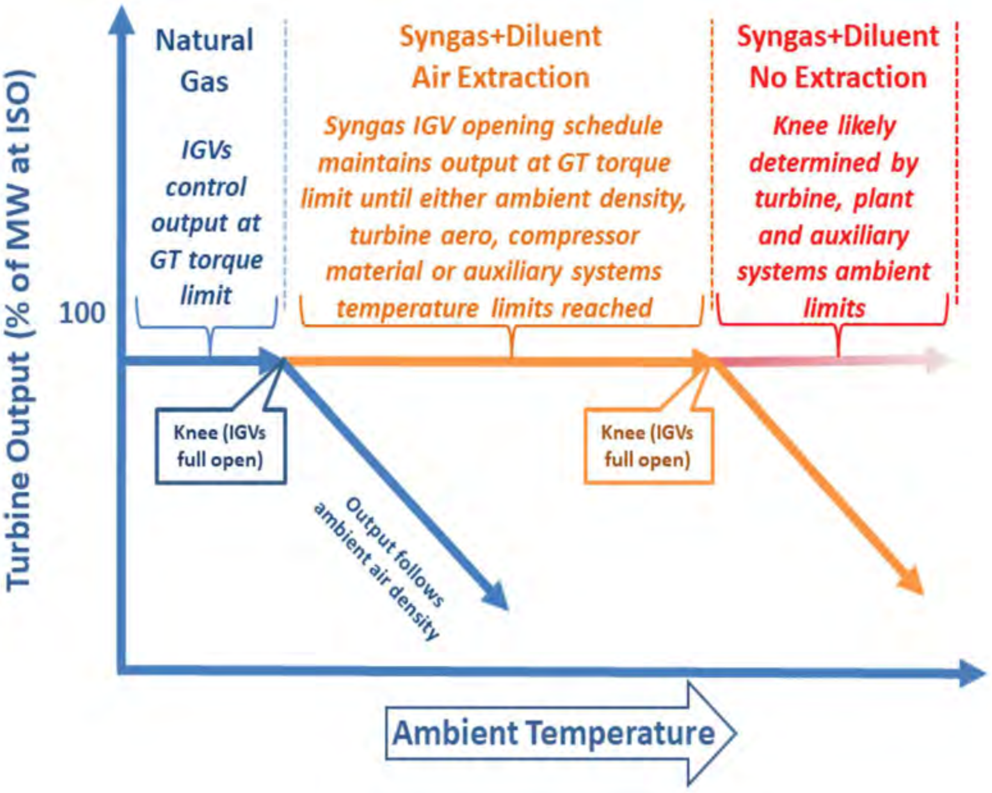


Figure 7-26. IGV control with ambient temperature

In addition to maintaining shaft torque below its limit, achieving the desired net power output requires an evaluation of other potential output limiting factors, such as:

- Is there adequate generator and STG cooling capacity at high ambient temperature?
- Are adequate compressor surge and stall margins maintained over the full operating range?
- Are latter stage compressor material strength properties adequate for maintaining the full load design pressure ratio at high ambient temperatures?
- Are there aero-mechanical limitations of the turbine (e.g., stage Mach number)?

These considerations and the identification of other potential limitations require that the plant designer consult with turbine and equipment suppliers through plant design optimization and the development of plant equipment specifications.

7.10 CARBON DIOXIDE CAPTURE

Carbon dioxide capture can be implemented within the gasification island (i.e., pre-combustion) or as an addition to the power block (i.e., post-combustion). Various options are listed with impact and mitigation to performance and implementation components in Table 7-7.

Syngas turbines operate at high air-to-fuel ratios and CO_2 exhaust concentrations of approximately 6 to 8% vol at atmospheric pressure. Within the gasification plant, the majority of the carbon will wind up as CO and its conversion to CO_2 (exothermic) and subsequent removal can be accomplished (endothermic regeneration) at high pressure (greater than 20 bar) and CO_2 concentration (approximately 20% vol) as part of the gasification plant's AGR train. The higher CO_2 concentration, combined with the high pressure of syngas, results in an approximately 50/1 mass transfer advantage for pre-combustion CO_2 capture compared to post-combustion CO_2 capture. The high oxygen content in the flue gas also accelerates the degradation of amines commonly used for post-combustion CO_2 capture.

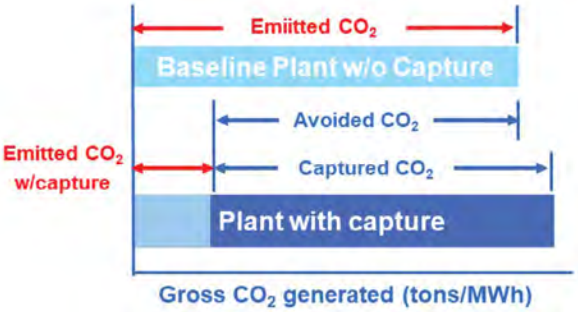
Table 7-7. Comparison of CO_2 capture options for a gasification power plant

CO_2 CAPTURE OPTIONS FOR A MODULAR GASIFICATION PLANT				
CASE	DESCRIPTION	POWER BLOCK IMPACT	MITIGATION	COMMENTS
PRE-COMBUSTION CAPTURE				
1. CO_2 capture within gasification syngas treatment train	Catalytic CO-shift Scrubbers with sorbent having high CO_2 selectivity Potential refrigerated sorbent	High H_2 syngas to turbine Reduced syngas mass flow and output Increased steam extraction for CO_2 sorbent regeneration Combustor and fuel skid modifications	Premixing of diluent and syngas to combustor H_2 limits Additional diluent to maintain output and reduce NO_X	Mixture capture processes Lowest avoided CO_2 cost Proven turbine capability (up to 100% H_2)
POST-COMBUSTION CAPTURE				
2. CO_2 capture from syngas turbine exhaust gas	Addition of CO_2 scrubber after HRSG (e.g., monoethanolamine (MEA) and compressors for CO_2 export for CCS	High capital cost – may exceed that of the power plan Reduced output and efficiency Higher exhaust back-pressure Lower gas turbine pressure ratio and efficiency LP Extraction for solvent regeneration and smaller steam turbine LP Exhaust gas reheat to maintain stack buoyancy	Redesign and re-optimization of bottoming cycle CCS optimized steam turbine	High cost has disincentivized commercial implementation
3. Exhaust CO_2 capture with partial recirculation of exhaust gas to turbine inlet	Same as Case 2, but with a fraction of the exhaust gas from the HRSG returned with fresh air to the gas turbine compressor inlet	Reduced capital cost and energy requirement of the absorber Increases gas turbine compressor air inlet temperature and reduces output similar to hot day	Cooling of EGR extraction air	Limited commercial experience
ADVANCED LOW CO_2 CYCLES				
4. Oxy-fuel combustion	Pure O_2 for syngas combustion in place of air combustion Eliminates CO_2 AGR in gasification plant	No air nitrogen eliminates NO_X production Extreme firing temperature requires new and advanced turbine hot-gas-path (HGP) design, materials, and combustor	Additional development	Needs evaluation for further development

Table 7-7. Comparison of CO_2 capture options for a gasification power plant

CO_2 CAPTURE OPTIONS FOR A MODULAR GASIFICATION PLANT				
CASE	DESCRIPTION	POWER BLOCK IMPACT	MITIGATION	COMMENTS
5. Supercritical CO_2 cycles (i.e., Allam-Fetvedt Cycle)	Semi-closed cycle oxy-fuel combustion with supercritical CO_2 working fluid	Turbine suitable for supercritical CO_2 adopted from ultrasupercritical steam technology Oxy-fuel combustion eliminates NO_X allowing for simple diffusion combustor Recuperative cycle eliminates bottoming steam cycle No stack – only output is CCS-ready supercritical CO_2	Pilot validation of concept with natural gas and front-end engineering design (FEED) studies in support of commercial implementation completed Pre-FEED for coal syngas underway	Current focus is development of commercial natural gas projects Syngas development underway

Pre-combustion capture provides a significant advantage over post-combustion capture in terms of capital, operating costs, and parasitic power demand. (Plans by Statoil-Hydro to employ post-combustion capture to natural gas combined cycle turbines at the Mongstadt refinery to be used for enhanced oil recovery were abandoned due to high cost and permitting difficulty due to Amine carryover from the scrubbers.[50]) There are limited studies that have direct comparative assumptions and design parameters (e.g., fuel costs, capacity factor, configurations, project financial assumptions, construction and materials cost indices, inflation, and discount rates). Cost and performance studies are outdated between the date they are performed and when they are published; however, a useful parameter to evaluate these options for the developer's decision of whether to employ CO_2 capture is the avoided cost of CO_2 capture that estimates the levelized cost of electricity (LCOE) impact of adding CO_2 capture to the baseline plant without CO_2 capture (Figure 7-27).[51, 52]



$$\text{Avoided Cost (\$/ton)} = \frac{COE_{cc}\,(\$/MWh) - COE_{base}(\$/MWh)}{CO_{2\ base}(tons/MWh) - CO_{2\ cc}\,(tons/MWh)}$$

$$\text{Capture Cost (\$/ton)} = \text{Avoided Cost (\$/ton)} \times \frac{\text{Avoided tons/MWh}}{\text{Captured tons/MWh}}$$

Used with permission from GE[53]

Figure 7-27. Defining avoided cost of captured CO_2

Another interpretation is the total tax on carbon emissions needed to justify implementing CO_2 capture (should such a policy be implemented where the avoided cost of CO_2 is a key metric). Representative avoided CO_2 capture costs[52] (using NGCC and supercritical pulverized coal [SCPC] plants' avoided cost as a proxy for a syngas fueled combined cycle) are given in Table 7-8.

Table 7-8. Comparison of avoided cost for post and pre-combustion capture[52]

AVOIDED CO_2 CAPTURE COST*						
POWER TECHNOLOGY	CAPTURE TECHNOLOGY	FUEL	CO_2 EMISSIONS NO CAPTURE (KG/MWH)	CO_2 EMISSIONS WITH CAPTURE (KG/MWH)	CAPTURE LEVEL %	CO_2 AVOIDED COST ($/TON)
IGCC	Pre-Combustion	Bit Coal	793	115	85%	$43
SCPC	Post-Combustion	Bit Coal	804	109	86%	$59
NGCC	Post-Combustion	Natural Gas	370	55	85%	$80

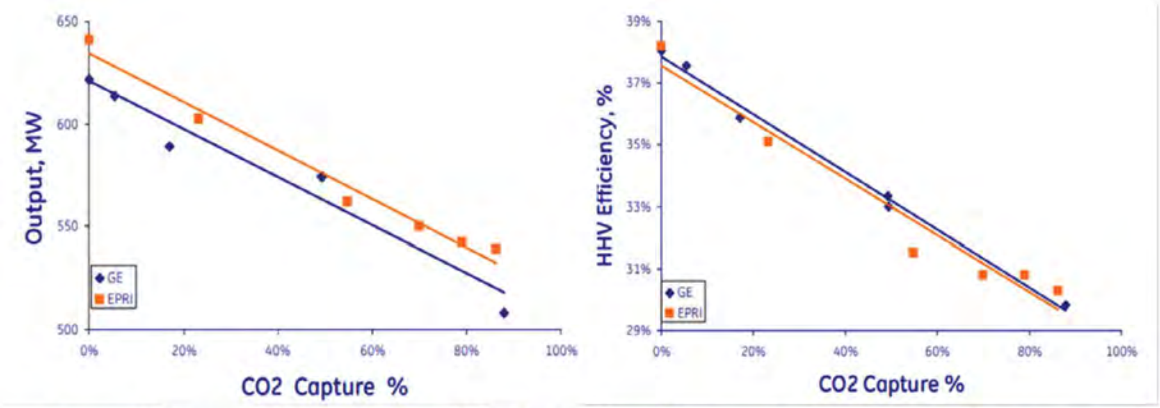
**Average of studies for plants in Organization for Economic Cooperation and Development (OECD) countries.*

7.10.1 LEVEL OF CARBON DIOXIDE CAPTURE

The implementation of CO_2 capture requires a decision on how much carbon should be captured. For power plants, two criteria have historically been used to determine the required level of CO_2 capture:

- **NGCC Equivalence**—Requires the reduction of CO_2 from zero capture (approximately 1,800 lb/MWh for a normal IGCC plant) to 800 lb/MWh (NGCC equivalent). This can be achieved at approximately 50 to 65% capture, depending on the plant's baseline efficiency.[53] To wit, the Kemper county IGCC plant, for the few months it was able to operate, had 67% capture.
- **Full (90%) Capture**—Results in approximately 200 lb/MWh—approximately 25% of that of NGCC without CO_2 capture.

At NGCC equivalence, syngas hydrogen content will be approximately 65% vol, which meets the specification limits of those from GE and Siemens for their F-class syngas combustors (for example). With the transition to a net-zero carbon energy economy, NGCC equivalence is no longer considered a valid criterion, and full capture is essentially required for every plant that does not use 100% hydrogen. If biomass is used as the primary fuel, net-zero life-cycle CO_2 emissions can be achieved with less than full capture due to the fact that biomass is carbon-neutral. However, at 90% capture, the hydrogen concentration will exceed 90% vol. For hydrogen exceeding 65%, pre-dilution of the syngas is required. Diffusion combustors use separate injection of syngas, air, and diluent at the combustor head end into the combustion zone. Due to NO_X degradation and combustor dynamics, the maximum hydrogen concentration at the fuel injection nozzles is limited to 65%. When syngas hydrogen concentration exceeds 65%, pre-mixing of the syngas with diluent is required to reduce its concentration to 65% to ensure suitability with the combustor. Additional diluent for NO_X reduction or output enhancement can still be injected at the combustor as needed. Turbines with diffusion combustors have been successfully applied to syngas blends containing greater than 90% hydrogen. There are output and efficiency advantages for choosing less than full capture; for example, the predicted performance degradation with increasing capture level is shown in Figure 7-28 for the addition of CO_2 capture to GE's coal IGCC reference plant design.[54] These efficiency/performance penalties can be mitigated or avoided by choosing some form of biomass as the primary fuel (since biomass is essentially carbon-neutral).



Sources: GE Internal Study, 2007 and EPRI Design Considerations for CO2 Capture: Engineering and Economic Assessment of IGCC Coal Power Plants for near-term Deployment, 2008

Used with permission from GE[55]

Figure 7-28. IGCC performance degradation with increasing CO_2 capture level

7.10.2 OFFSETTING THE COST OF CARBON DIOXIDE CAPTURE

Based on the project's avoided CO_2 cost, a project developer may be able to make an informed decision for CO_2 capture based on the following:

- 45Q federal tax code.
- Reduced cost for the purchase of offsets within regional cap and trade markets.
- Utilization of captured CO_2 for chemical production.
- Sale to a CO_2 pipeline operator for enhanced oil recovery (EOR).

7.10.2.1 U.S. INTERNAL REVENUE CODE SECTION 45Q FEDERAL TAX CODE

Per U.S. Internal Revenue Code Section 45Q enacted in 2009, CO_2 capture projects that meet eligibility requirements and store CO_2 in geologic formations according to federally established requirements, or that otherwise apply it to beneficial and secure use in products, are eligible for a 45Q credit. In addition to generated power, eligible products must result in a net reduction in emissions on a life cycle basis. As amended in 2019 by H.R. 2486, 45Q tax credits were increased to \$50 per short ton of CO_2 for storage and \$35 for EOR.

7.10.2.2 CARBON CAP AND TRADE MARKETS

The two regional carbon trading markets in the United States are the Regional Greenhouse Gas Initiative (RGGI) in New England and the Mid-Atlantic and the California Cap-and-Trade Program (CCTP). These trading programs require that major CO_2 emitting facilities acquire CO_2 allowances to offset emissions that exceed state compliance levels. Based on recent auctions, allowance prices have ranged from \$2.00/short ton to \$5.00/short ton.

7.10.2.3 CARBON DIOXIDE UTILIZATION

Carbon dioxide is used in numerous applications, such as fertilizer production,[55] EOR, carbonation of beverages, freezing and cooling food products, and as an inert gas in firefighting.[56] Among these, EOR is the largest consumer.[57] The price of CO_2 that can be obtained for EOR is dependent on the accessibility of a CO_2 pipeline and the price of oil. Total CO_2 costs (both purchase price and recycle costs) can amount to 25 to 50% of the cost per barrel of oil produced. Due to the volatility of oil price, the price that oil well developers are willing to pay is also highly volatile, making firm estimates of price and project revenue challenging.

7.11 ELECTRIC GRID INTERCONNECT

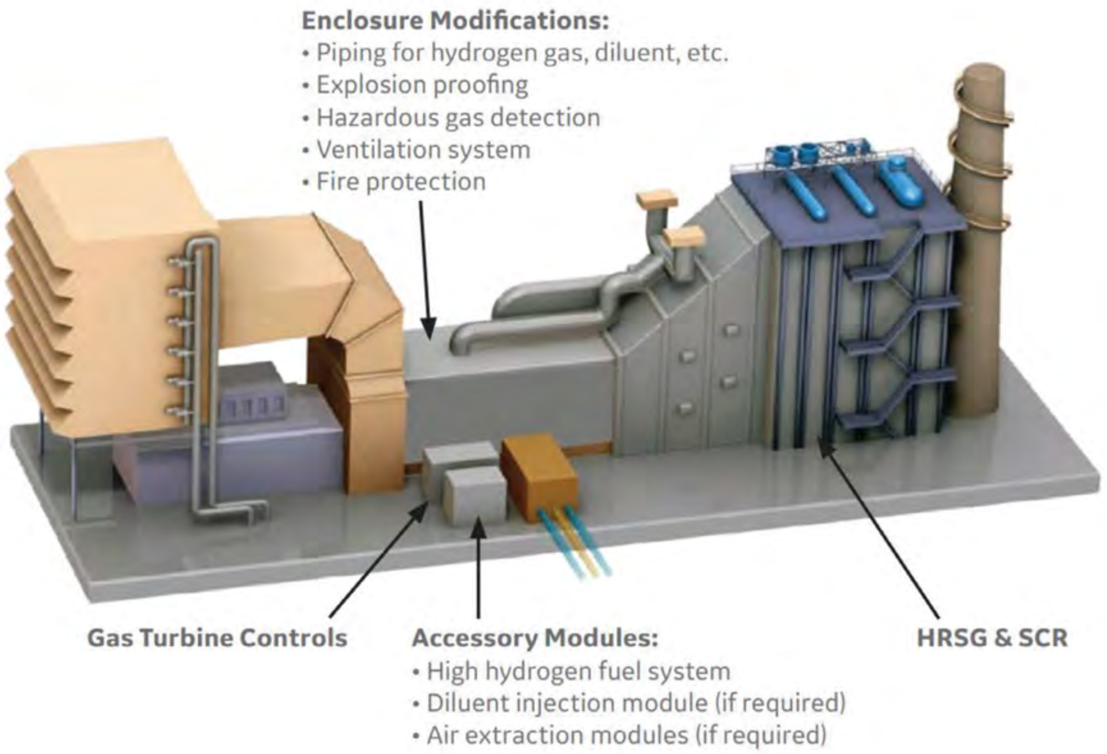
The interconnection provides access to the power markets by way of the transmission grid (see Chapter 8). Securing an acceptable electrical interconnection to the local utility can either allow or indefinitely delay power plant development. A well-designed interconnection is necessary to meet code and safety requirements. The typical data required to apply for connection to a grid includes:

- Estimated reactive capability curves (leading/lagging MegaVars versus Megawatts).
- Generator output as a function of generator cold gas temperature.
- Generator output as a function of liquid temperature.
- Estimated saturation and synchronous impedance curves (per unit armature voltage and current) as a function of field current.
- Estimated VEE curves (per unit kV-Amperes versus Field Current [AMPS]).
- Startup power draw and whether the plant has black start capability.

7.12 SAFETY

Safety is a paramount requirement. This summary highlights additional power block safety considerations associated with syngas and should not be interpreted as a comprehensive listing of power-block associated safety requirements.

For the turbine power block, the syngas control skid enclosure and the gas turbine compartment are enclosed spaces and sources where personnel can be exposed to hazardous gases (Figure 7-29). Hydrogen requires additional safety measures due to its non-luminous flame and high combustibility. The potential hazards associated with syngas are summarized in Table 7-9.



Used with permission from GE[2]

Figure 7-29. Power block modifications for syngas

Table 7-9. Safety measures for syngas

SYNGAS COMPONENT	SAFETY RELATED CHARACTERISTIC	SAFETY CONCERN	WHERE	MITIGATION
Hydrogen	Non-luminous flame	Not visually detectable Personnel compartment entry	Turbine enclosure Fuel Skid compartment	Additional UV and Hazgas Detectors
	Broad flammability range (auto-ignition temperature ~1,000°F/540°C)	Explosion	Turbine enclosure, fuel skid compartment	Compartment explosion proofing Increased compartment ventilation
		Turbine trips Combustor flameouts	Turbine casing, HRSG and exhaust ducting	Combustor flame detection Rapid fuel shutoff Lengthened evacuation
	Low specific volumetric energy density	Larger gas supply manifolds	Turbine enclosure	Increased purge capacity and reduced exchange times
	Low MW	Leakage across seals and gaskets	Piping and manifolding	Welded construction wherever possible
CO	Buoyancy	Trips, combustor flameouts and hideout at high points	HRSG	Increased length of air purge
	Toxicity	Personnel exposure	Turbine enclosure, fuel skid compartment	CO detectors Increased compartment ventilation

7.12.1 OCCUPATIONAL SAFETY AND HEALTH ADMINISTRATION AND THE HAZARD AND OPERABILITY PROCESS

The power block is subject to the Occupational Safety and Health Administration's (OSHA) Project Safety Management (PSM) requirements[58] for the prevention of and response to the accidental release of hazardous substances within a process facility that can threaten workers. The power block should be applicable to hazard and operability (HAZOP) review in conjunction with the total plant HAZOP analysis. The HAZOP review process identifies potential safety and operability issues as early in the design process as possible. Usually, the first HAZOP analysis is performed when front-end engineering and design (FEED) and Piping and Instrumentation Diagrams (P&IDs) are available and the design can be modified without major cost. At this point, the design is defined well enough to allow meaningful answers to the questions from the HAZOP process. However, the review is considered preliminary in nature, and a second detailed HAZOP study is performed prior to the release of Approved for Construction P&IDs. The HAZOP analysis of turbine supplier drawings is performed as they become available and continues throughout the plant construction period. However, design changes may be costly by this point.

7.13 REFUELING AND REPOWERING OF BOILERS

The refueling (i.e., fuel substitution) of pulverized coal (PC) power plants could provide a relatively low capital-cost strategy where a modular gasification plant could provide syngas as a replacement for the PC boiler's coal combustion. The cost could be significantly reduced through the reuse of the steam turbines with improved environmental performance. Additional challenges are the limited space at existing sites and the competing option of refueling by conversion to natural gas. Low natural gas prices—in the United States—have made the latter the preferred choice for both refueling and new plants (at the time of publication).

Repowering (i.e., power block replacement) entails the complete replacement of a PC boiler with a gasification plant and syngas turbines and the retrofitting of the steam turbines with a HRSG to complete a syngas combined cycle. An example of this is the Wabash River Coal Repowering Project[59] (USA), which installed GE 7FA.01 gas turbines as the primary power block with HRSG configuration matched to the existing steam turbines. In this case, existing site and water use permits were helpful. The current decommissioning of many PC plants will bring many steam turbines and associated auxiliary equipment to the "grey" market at low prices. Although steam conditions may be imperfect for optimal performance, developers should consider the availability of this equipment to substantially reduce capital costs for modular gasification power projects.

7.14 SUMMARY

The intent of this chapter is to inform developers of modular gasification projects of the options and associated factors relevant to the choice of a power block. The recommendations are derived from a long history of successful commercial application of gas turbines for efficient and clean utilization of gasification-derived syngas. A project developer has many options for configuration and optimization to achieve project financial requirements. To inform the decision process, the developer must have a definitive listing and ranking of criteria, especially their allowable ranges. Modular gasification plants enable the economical use of local sources of both biofuels and municipal waste—whether as primary or secondary feedstock. When these biofuels/wastes are combined with CO_2 capture, a net-negative carbon footprint is possible.

Endnotes

1 NETL (May 2007) “Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity Final Report”, DOE-NETL-2007/1281.

2 GE Energy (January 2019) “Gas to Power – the Art of the Possible”, GE Energy GEA34108.

3 C. Higman (October 19, 2016) “State of the Gasification Industry: Worldwide Gasification and Syngas Databases”, Gasification & Syngas Technologies Conference, Vancouver.

4 Wartsila Inc (2017) “Gas and Multi-Fuel Power Plants”, Brochure.

5 GE Power Systems (1999) “Integrated Gasification Combined Cycle”, GE Power Systems, GEA-13021.

6 GE Power (February 2016) “Heavy-Duty Gas Turbine Flange-to-Flange Solution”, GEA30379B, *https://www.ge.com/content/dam/gepower-* pgdp/global/en_US/documents/service/gas%20 turbine%20services/flange-to-flange-fact-sheet.pdf.

7 Mitsubishi Heavy Industries Group (n.d.) “History of Mitsubishi Power’ Gas Turbines”, Mitsubishi Power, Ltd., *https://www.mhps.com/products/gasturbines/history/index.html*.

8 Ansaldo Energia (n.d.) “GT26 The Best Value”, *https://www.ansaldoenergia.com/PublishingImages/GT26/GT26.pdf*.

9 Nils Lindstrand (August 2, 2019) “This Swedish scientist works towards fulfilling Siemens Energy’s 2030 hydrogen pledge”, Siemens Energy, *https://www.siemens-energy.com/global/en/news/magazine/2019/hydrogen-capable-gas-* turbine.html.

10 Siemens (2017) SGT5-2000E heavy-duty gas turbine (50 Hz), *https://www.siemens*- *energy.com/global/en/offerings/power-generation/gas-turbines/sgt5-2000e.html*.

11 Siemens (2017) SGT6-5000F heavy-duty gas turbine (60 Hz), *https://www.siemens*- *energy.com/global/en/offerings/power-generation/gas-turbines/sgt6-5000f.html*.

12 Higman C. and M. van der Burgt (September 18, 2003) “Gasification” Elsevier, Gulf Professional Publishing.

13 J. Goldmeer (February 2019) “Power to Gas: Hydrogen for Power Generation”, GE Power, GE-33861.

14 P. Stuttaford et al. (2016) “FlameSheetTM Combustor Engine and Rig Validation for Operational and Fuel Flexibility with Low Emissions”, ASME Turbo Expo 2016, Paper GT2016-56696, June 13-17, 2016.

15 H.E. Miller (March 3, 1988) “Development of the GE Quiet Combustor and Other Design Changes to Benefit Quality”, GE Power, GER-3551, Presentation to American Cogeneration Association in San Francisco, California.

16 Robert Jones and Jeffrey Goldmeer (2009) “Meeting the Decarbonized fuel challenge with gas turbines”, CoalGen Europe, September 1-4, 2009.

17 N.Z. Shilling (March 23, 2003) “Gas Turbines for Low CO_2 Power Production”, Conference on Cycles for Low Carbon Dioxide, Cranfield University.

18 GE Power Systems (October 2000) “Economic and Technical Considerations for Combined Cycle Performance- Enhancement Options”, GE Power Systems State-of-the-Art Paper # GER-4200.

19 A.P.P. Santos, C. Andrade, and E. Zaparoli (2012) “Comparison of different gas turbine inlet air cooling methods”, World Acad. Sci. Eng. Technol. 61:40-45.

20 F.J. Brooks (October 2000) “GE Gas Turbine Performance Characteristics”, GE Power, GER3567H.

21 R.M. Jones, J.M. Bisseaud, and M. Moliere (2007) “Power Generation Markets Driving Fuel-Flexible Solutions”, Power-Gen Europe 2007, Madrid, Spain.

22 N.Z. Shilling and R.M. Jones (2003) “The Impact of Fuel Flexible Gas Turbine Controls on Integrated Gasification Combined Cycle Performance”, GT2003-38791, Proceedings of ASME Turbo Expo 2003 Power for Land, Sea and Air June 16–19, 2003, Atlanta, Georgia.

23 H. Morehead (2005) “Siemens Technology Improvements Enhance IGCC Plant Economics”; Gasification Technologies Conference, San Francisco CA, October 12, 2005.

24 GE Energy (January 2002) “Specification for Fuel Gases for Combustion in Heavy-Duty Gas Turbines”, GE Energy GEI41040G. *https://citeseerx.ist.psu.edu/viewdoc/download?doi=10.1.1.143.6392&rep=rep1&type=pdf*.

25 EML Manufacturing, Fuel Gas Conditioning Skid for Gas Turbines”, *http://www.emlmanufacturing.com/Fuel_Gas_Conditioning.htm#Fuel_Gas_Skid_For_Gas_Turbines*, EML Manufacturing, LLC.

26 EPA, 40 CFR Part 50 - National Primary and Secondary Ambient Air Quality Standards. Amended 1994.

27 EPA, 40 CFR Part 60.40Da - Applicability and designation of affected facility. Amended February 16, 2012.

28 EPA, 40 CFR Part 60 - STANDARDS OF PERFORMANCE FOR NEW STATIONARY SOURCES. 1971.

29 EPA, 40 CFR Part 60.41Da - Definitions. Amended April 24, 2013.

30 EPA, 40 CFR Part 60.44Da - Standards for nitrogen oxides (NO_2). Amended February 16, 2012.

31 N.Z. Shilling and R.M. Jones (2003) “The Response of Gas Turbines in a CO_2 Constrained Environment”, Proceedings 2003 Gasification Conference, San Francisco, October 2003.

32 V. Iyer, J. Haynes, P. May, A. Anand (2005) “Evaluation of Emissions Performance of Existing Combustion Technologies: Paper GT2005-68513 Proceedings of GT2005 ASME Turbo Expo 2005: Power for Land, Sea and Air June 6-9, 2005, Reno- Tahoe, Nevada.

33 T. Schultz (April 7, 2016) “Select NETL Turbines-Focused Systems Analysis”, NETL-PUB-20427.

34 The Babcock & Wilcox Company (1992) Steam – Its Generation and Use, Chapter 31, 40th edition, North Carolina: Babcock & Wilcox.

35 T. Ginter, T. Bouvay (2006) “Uprate Options for the MS7001 Heavy Duty Gas Turbine”, ger3808c, GE Energy.

36 Cleaver-Brooks Inc (2019) HEAT RECOVERY STEAM GENERATOR, When Reliability Matters, CB-8586 2019, *http://cleaverbrooks.com/docs/brochures/CB-8586-HRSG.pdf*.

37 R.W. Fisk and R.L. VanHousen (August 1996) “Cogeneration Application Considerations”, GE Power Systems GE Power Generation GER3430F.

38 EPA, 40 CFR Subpart Da - Standards of Performance for Electric Utility Steam Generating Units. June 13, 2007.

39 D. Roberts (March 28, 2020) “The Missing Puzzle Piece for Getting to 100% Clean Power”, VOX, *https://www.vox.com/energy-and-environment/2020/3/28/21195056/renewable-energy-100-percent-clean-* electricity-power-to-gas-methane.

40 B. Jones-Albert (October 12, 2017) “Confronting the Duck Curve: How to Address Over-Generation of Solar Energy”, DOE Office of Energy Efficiency and Renewables.

41 EIA (January 14, 2020) “Short Term Energy Outlook”, EIA.

42 M. Yamaguchi (2004) “First Year of Operation of Operational Experience with the Negishi IGCC”, Gasification Technologies Conference, Washington D.C., October 3-6, 2004.

43 H. Tenaka (2004) “The Control System Applied in Negishi IGCC”, Gasification Technologies Conference, Washington D.C., October 3-6, 2004.

44 EPRI (1990) “Coolwater Coal Gasification Final Report”, EPRI Report GS-6806.

45 N.Z. Shilling (January 2009) “Clean Coal and Carbon Capture”, Presented at U.C. Berkeley DNV Technical Seminar.

46 N.Z. Shilling (2014) “Gas Turbine Trends in Fuel Flexible Applications”, GE Power, Presented at Coal Gen 2014, Nashville, TN.

47 F. Hannemann, B. Koestlin, G. Zimmermann, H. Morehead, F. García Peña (2003) “Pushing Forward IGCC Technology at Siemens”, Presented at 2003 Gasification Technologies Conference, San Francisco, California, October 12-15, 2003.

48 L. Duke (2010) “GE’s IGCC Solution Update and Considerations for Carbon Capture”, Presented at Coal Gen Conference, Pittsburgh PA, August 24, 2010.

49 K. Payrhuber, R.M. Jones, and M.H. Scholz (June 2008) “Gas Turbine Flexibility with Carbon Constrained Fuels,” ASME Paper GT2008-50556, Proceedings of GT2008 ASME Turbo Expo, Berlin Germany.

50 BBC News (20 September 2013) “Norway abandons Mongstad carbon capture plans”, *https://www.bbc.com/news/world-europe-24183443*.

51 M. Finkenrath (2011) “Cost and Performance of Carbon Dioxide Capture from Power Generation” Working Paper, IEA.

52 N.Z. Shilling (2008) “New Thinking and Old Standby Solutions for Coal: GE Power, Presentation at 2008 NAAG Presidential Initiative on Energy, May 7, 2008, Cour da Lane, Idaho.

53 R. Schonewald (February 10, 2009) “Turbine and System Developments for Syngas/Hydrogen Gas Turbine Technology”, presentation at International Colloquium on Environmentally Preferred Advanced Power Generation.

54 K. White (October 2008) “GE’s Carbon Island”, Proceedings of the Gasification Council Annual Conference, Washington DC.

55 D. Heaven, J. Mak, D. Kubek, M. Clark, C. Sharp (2004) “Synthesis Gas Purification in Gasification to Ammonia/Urea Plants”, Gasification Technologies Conference, Washington DC, October 4-6, 2004.

56 IPCC (2005) IPCC Special Report on Carbon Dioxide Capture and Storage. Prepared by Working Group III of the Intergovernmental Panel on Climate Change, B. Metz, O. Davidson, H.C. de Coninck, M. Loos, and L. A. Meyer, eds. Cambridge, United Kingdom and New York, NY: Cambridge University Press.

57 NETL (March 2010) “Carbon Dioxide Enhanced Oil Recovery – Untapped Domestic Energy Supply and Long Term Carbon Storage Solution”, DOE/NETL.

58 OSHA, 29 CFR Part 1910.119 - Process safety management of highly hazardous chemicals. April 15, 2019.

59 NETL (2000) “Wabash River Coal Gasification Repowering Project Final Technical Report”, DOE/NETL Morgantown, West Virginia Cooperative Agreement DE-FC21-92MC29310.

8.0 GUIDELINES FOR THE DESIGN OF AN ELECTRIC GRID CONNECTION MODULE

8.1 INTRODUCTION

A gasification plant combined with gas-fired and/or fuel cell generators produces electricity that is exported to the electric grid. A project developer backed by investors can build this facility and sell electricity for revenue that recovers the investment costs and provides a rate of return. However, the facility must have an electric grid interconnection to do so, hereafter referred to as a "grid connection module." This chapter addresses how to permit and build substations necessary for this interconnection. The revenue potential from selling power is best estimated with an understanding of the electric grid (e.g., generation, transmission, and distribution) and markets. For the project developer, the key is to maximize return on investment (ROI), simply defined as the ratio of the benefit (or return) of an investment divided by the cost of the investment. Maximizing ROI is done by diminishing the variables impacting the benefit (e.g., making suboptimal decisions, reduced energy demand, and taxes) or reducing the cost by making efficient investments. The planning, design, and construction of the power plant and its interconnection to the electric grid have the most impact on ROI. Understanding the electric grid will also help the project developer to define the investment costs. This chapter addresses how electricity is produced, sold, and consumed, followed by the investment in the electric grid connection module.

The electricity produced by gasification-based and other generation facilities is sold to customers, such as homes and businesses, over the electric grid. Power plants connected to the electric grid participate in electricity markets. An overview of the components of the electric grid and how the grid functions will help project developers and investors understand how they can profit from a gasifier-fueled generation power plant. These concepts will address both benefits (revenue) and costs (capital, operating, and maintenance).

8.1.1 INTRODUCTION TO THE ELECTRIC GRID: GENERATION, TRANSMISSION, AND DISTRIBUTION

The electric power industry is arranged into three separate functional units: generation, transmission, and distribution. Understanding these functional units is critical when planning to connect any generation system to the electric grid.

8.1.1.1 GENERATION

Generation is the act of producing electrical energy from an energy resource. When a turbine converts wind or moving water to electricity, or a photovoltaic panel converts sunlight to electricity, that turbine or panel is a generator[1] and can be part of a generation facility. Large power plants, often operated by electric utilities, generate huge amounts of electrical power, usually hundreds to thousands of megawatts (MW) and cumulatively huge amounts of energy (typically measured in megawatt-hours [MWh]).

Nameplate capacity (also known as the rated capacity, nominal capacity, installed capacity, or maximum effect) is the intended full-load sustained output of a facility, such as a power plant or electric generator. Capacity is the maximum amount of power a plant can generate on a consistent basis and is usually significantly below the nameplate (advertised) capacity, because the plant has to service internal loads (e.g., control systems, fuel and exhaust systems, and other devices) so that the plant can operate correctly.

Non-utility entities that own and operate the means to produce electricity (e.g., renewable or traditional generation) can generate electricity and sell it back to the electric grid. These are called independent power producers (IPPs), which play an important role in many states according to the state regulations and policies of the public utility commission (PUC). An IPP or non-utility generator (NUG) is a non-public entity that owns facilities to generate electric power for sale to utilities and end users. IPPs invest in generation technologies and recover their costs from the sale of the electricity.[2]

Utilities recover costs for providing electric service to retail customers through a combination of rate components that comprise customers' monthly electric bills. Rates and rate designs are set by state regulators and vary by jurisdiction, utility, and customer class. In addition to the fundamental tenet of setting fair and reasonable rates, rate design balances economic efficiency, equity and fairness, customer satisfaction, utility revenue stability, and customer price and bill stability. The rate of return for utilities goes to their investors; for IPPs, any excess profits go to the investors behind the IPP.

8.1.1.2 TRANSMISSION

Overhead transmission lines[3] on large towers use alternating ("back and forth") current (AC), because of historical limitations of direct current (DC). The so-called "War of the Currents" showed that AC was superior for long distance transmission, because DC is not easily converted to higher or lower voltages and higher voltages are necessary for transmission with low losses.[4] Transmission is the process of moving bulk amounts of electricity over these high-voltage, typically AC lines (115 to 765 kilovolts [kV]). AC-DC conversions are used for very long distances and there are also high-voltage DC lines with voltages of 100 to 1,200 kV (e.g., 450 kV for HVDC for Quebec-New England and 500 kV for the Pacific DC Intertie).

Once generated, electricity needs to be transported to consumers (sometimes over great distances) by interconnecting the generator to the existing, complex system of power lines and associated equipment that move electricity: the electric grid. A connection to the electric grid is an investment that is compensated by subsequent electricity sales in a dedicated market.

Electricity is bought, sold, and traded in wholesale and retail markets, which operate similarly to wholesale and retail markets for other products. The purchase and sale of electricity to resellers (entities that purchase goods or services with the intention to resell them to another entity) is done in the wholesale market, while the sale of electricity to consumers is done in the retail market.[5]

Organized wholesale electricity markets were created to address ever-increasing electricity prices and encourage innovation through free-enterprise competition. Both capacity and energy delivered can be compensated. Capacity represents a commitment of resources to deliver when needed, particularly in the case of an electric grid emergency.[6]

8.1.1.3 DISTRIBUTION

High-voltage transmission lines transmit bulk power to substations, where the electricity is converted to a lower voltage. That low-voltage electricity is then distributed to homes and businesses on separate, lower-voltage distribution lines (generally less than or equal to 35 kV).

Historically, most electricity has been produced at large facilities, such as coal, hydroelectric, or nuclear power plants. These facilities produce vast amounts of power and require a large electric grid capacity to transport that power to customers. A distribution pole with the equipment labeled is shown in Figure 8-1. This is representative of large, electric grid elements. The insulator at the top of the pole prevents arcing from the 4 kV (for example) power line to the pole. A protective fuse is shown along with a lightning arrestor ("rod"). By contrast, smaller generators can directly connect to low voltage distribution lines: electricity is distributed locally, without having to be transferred to higher voltage transmission lines. This is referred to as "distributed generation."

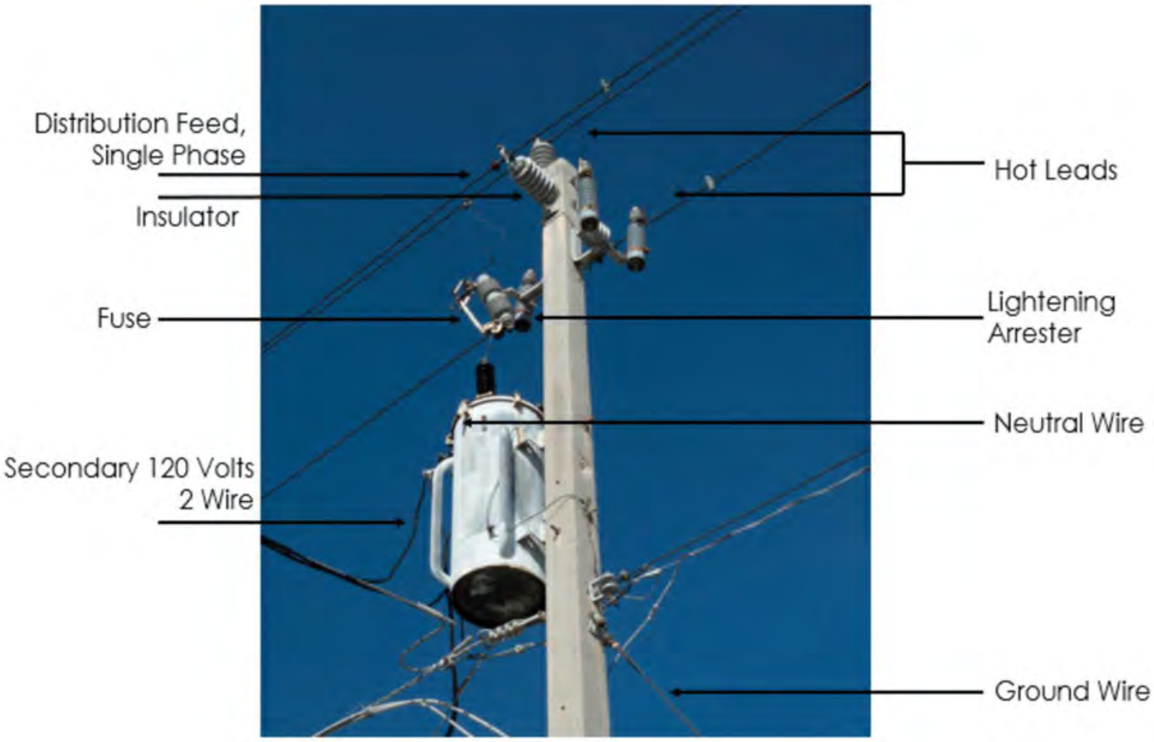


Figure 8-1. Distribution pole with equipment

8.1.1.4 POWER QUALITY

Utilities have minimum requirements for power quality due to contractual agreements with their customers.[7] A gasifier plant can be a source of high-quality power, compensated by the utility or by local customers. A high-quality power supply should be nearly constant amplitude (voltage, power) and the electric load (consumer) can vary in amplitude of power use, but the frequency and phase should be nearly locked to the supply. Terms often associated with power quality are "volts" and "vars" (volt-amps reactive), which are based on a phase difference between current and voltage[8] often due to motors and other devices with large magnetic fields that require compensation from the electric grid.[9] Volt-Var optimization can lead to high-quality electricity useful for distribution customers.[10] High-quality power is required by industrial and commercial customers.[11] Data centers or offices with computers, servers, storage will pay a premium for smooth power (i.e., power without voltage irregularities or harmonic disturbances).

8.1.2 DEVELOPMENT TYPE

To control investment costs, it is important to understand the three main classes of project development:

1. **Onsite Use Facility**—installing and owning a small facility for onsite energy use.
2. **Commercial-Scale with Offsite Export**—commercial-scale (locally- or community-owned) energy project with excess power sold for offsite use.
3. **Third Party**—working with a third-party project developer to develop commercial-scale systems with offsite delivery.

The customer-side voltage would likely be 4 kV, 13 kV, or 34.5 kV and above, respectively, for these cases, with customer side transformers stepping down the voltage as needed. Only the second and third cases require consultation with the utility for an electric grid connection (see Section 8.1.3). However, in the case of dual use with onsite use and export of power, the procedure is similar to the second and third cases. These three cases are discussed in the following subsections.

8.1.2.1 ONSITE USE FACILITY

A landowner or business can purchase and own a relatively small generator or facility designed to supply one's own energy needs. The generator might be off the electric grid and installed exclusively for onsite use or might be grid connected. Gasification would be an option for larger facilities, but is impractical for any site with maximum power needs less than 500 kilowatts (kW), equivalent to approximately 100 residential homes in the United States.[12]

8.1.2.2 COMMERCIAL-SCALE WITH OFFSITE EXPORT

Commercial facilities are designed to produce more electricity than is needed onsite, with the excess electricity sold to utilities and/or customers for profit. The sale of generated electricity requires a power purchase agreement (PPA) between the project owner and the utility. In the United States, PPAs are typically subject to regulation by the Federal Energy Regulatory Commission (FERC). FERC determines which facilities are applicable for PPAs under the Energy Policy Act of 2005 (EPACT 2005). A PPA is a legal contract between an electricity generator/provider and a power purchaser (typically a utility or large power buyer in an electricity market). Contractual terms may last anywhere from five to 20 years, during which time the power purchaser buys energy, and sometimes also capacity and/or ancillary services (e.g., power quality, including frequency control, spinning reserves, and operating reserves) from the electricity generator. Such agreements play a key role in the financing of independently owned (i.e., not owned by a utility) electricity generating assets. The seller under the PPA is typically an IPP.

A landowner developing a commercial-scale project must make careful decisions about how to structure the business side of the project, including which type of business entity will best fit the project's investment and ownership structure and how the project owners will comply with securities laws, filing and reporting requirements, and other legal obligations.

8.1.2.3 THIRD PARTY

Third-party development is a common form of commercial-scale ownership. This option usually requires comparatively little effort on the part of the landowner. A third-party project developer will likely coordinate the interconnection application and process along with most other aspects of project development.

Landowners can sell or lease their land to an entity willing to construct and operate an energy facility. The landowner communicates the development rights to a project developer, most likely using an option, lease, easement, or some combination of these. The project developer would then likely develop, build, and operate the generation project, with the landowner providing the land on which the project sits and receiving rental income or a small percentage of revenue generated by the facility.

This option for development entails less risk to the landowner than direct ownership and often requires no initial capital investment by the landowner. Contracting with a project developer requires the landowner to carefully negotiate legal agreements to ensure fair compensation and a fair allocation of the rights, responsibilities, and risks associated with development. Negotiating this kind of agreement requires significantly less effort than independently developing an entire project.

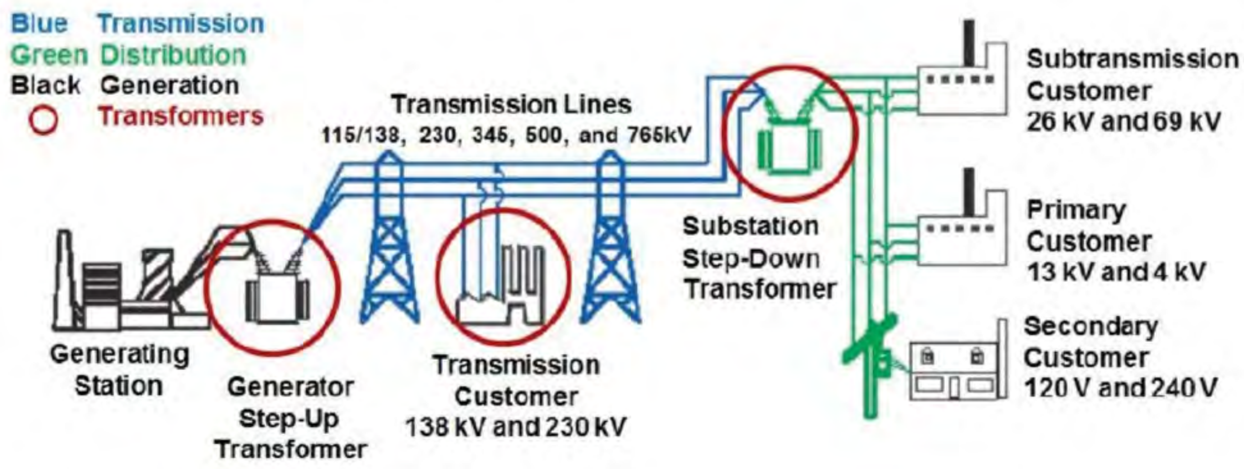
8.1.2.4 GASIFIER-FUELED DISTRIBUTED GENERATION

An emerging type of generation is gasifier-fueled power plants exporting power to the electric grid and potentially serving local loads. When searching for potential sites, it is beneficial to identify sites that are closer to existing transmission lines. If a site has existing electrical infrastructure, it is likely close to a significant sized substation. The availability of these and other sites with an existing interconnection can be obtained with commercially available databases.

Some local loads are connected with primary customer connections of 13 kV or 4 kV (Figure 8-2, which have higher voltage than the secondary customer connections that serve residential areas. Local authorities can help with exploring these connection opportunities.

It is important for the project developer to be familiar with the concept of a microgrid. A microgrid can be defined as "a group of interconnected loads and distributed energy resources (DERs) within clearly defined electrical boundaries that acts as a single controllable entity with respect to the grid." A microgrid can connect and disconnect from the electric grid to enable operation in either grid-connected or islanded (disconnected) mode.[13] The microgrid must survive intentional or unintentional disconnection ("islanding") from the main electric grid, meaning it still powers all the loads within its drawn boundary. This criterion adds complexity to the engineering and increases costs. The discussion in this chapter assumes the system has a graceful power down or "safe mode."

Figure 1. Electric Power Grid Representation



Source: DOE, 2006; see footnote 13. Modified based on industry review.

Figure 8-2. Role of transformers in connecting generators to the electric grid; three phases indicated[14]

8.1.3 INTERCONNECTION

From a network point of view, an interconnection is the physical linking of an entity's assets with equipment or facilities that do not belong to the entity. In the electricity context, an interconnection physically connects a power plant to the electric grid and consists of a transformer and hardware to synchronize the signal to the electric grid. The physical location of the interconnection should be surrounded by a fence and may look like a substation seen in many communities. The power plant discussed in this chapter will have one or more gasifiers, gas-fueled generators, and some other generators and control systems for the balance of plant. It will be able to power loads on a separate distribution circuit and disconnect from the main electric grid for short periods of time (i.e., hours). The plant will also be modular and called a "Modular Plant" ("ModPlant"), because it is not a fully featured microgrid where generation must equal load during islanding events. A predictable design for the ModPlant implies a repeatable, simplified design for the interconnection, yielding a modular grid interconnection (grid connection module). The ModPlant can become a microgrid with additional loads and control considerations.

The transformer is a "step-up" type, meaning it raises the AC voltage for longer range transmission. A schematic of the generating station and the generator's step-up transformer between the power plant and the electric grid is shown in Figure 8-2.[15] The step-up transformer is generally located in a substation, which is a high-voltage electric system facility. The substation is used to switch generators, equipment, and circuits or lines in and out of a system. It is also used to change the AC voltage or change AC to DC or DC to AC. Some substations are small with little more than a transformer and associated switches, while others are very large with several transformers and dozens of switches and other equipment.[16] The transformer interconnecting the ModPlant to the electric grid will likely be the former (small), but there will likely be a full substation within a few miles of the plant.

The utility is interested in protecting its electric grid with standard protective equipment and the design and implementation of the generating station during the interconnection process. The interconnecting transformer is electrically connected to generators and loads via a power plant bus (Figure 8-3). If this bus gathers power at a voltage of 4 kV from its generators, then a transformer (Figure 8-4) is needed to change the AC voltage from 4 to 69 kV, and a switch is needed to enforce power flow. The plans can emphasize power export onto the electric grid or add service of loads on a separate distribution circuit. This affects the contingencies during an outage on the utility grid, because the distribution grid provides a place to "sink" power.

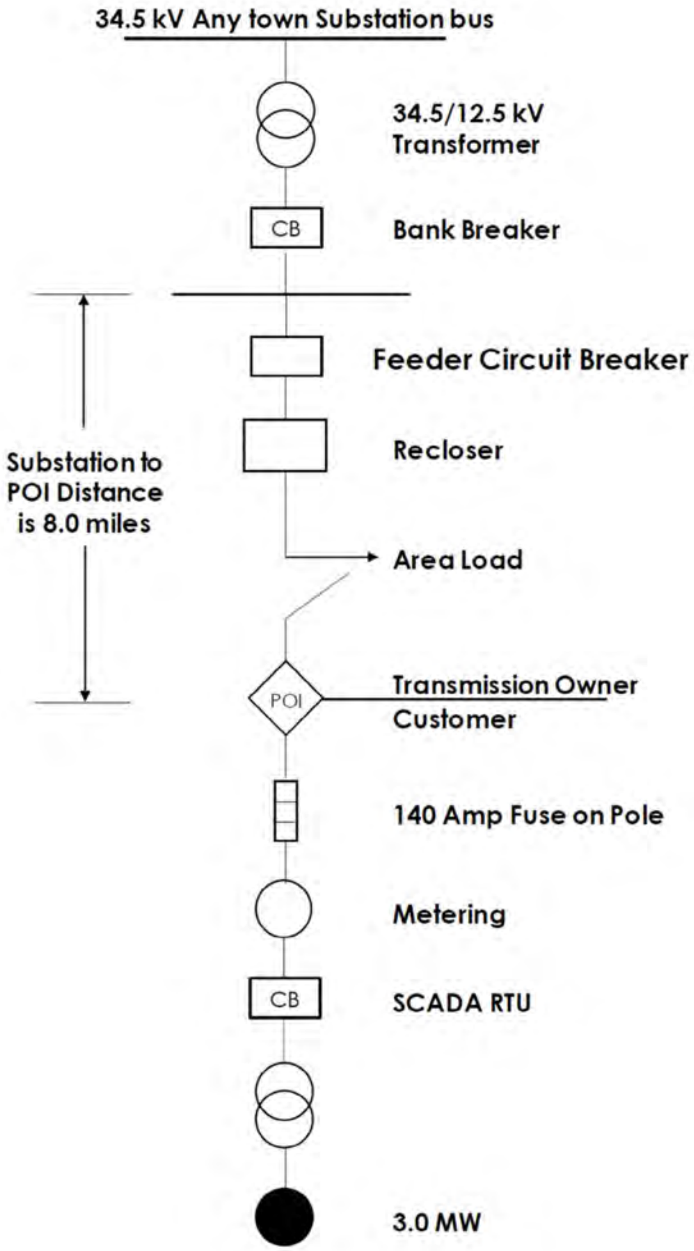


Figure 8-3. 3.0-MW generation single-line diagram

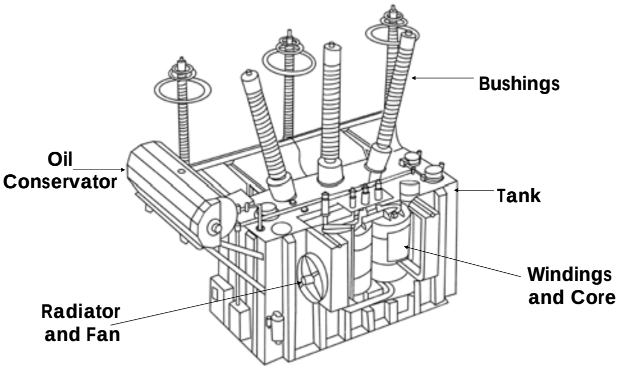


Figure 8-4. Substation showing transformer appropriate for connection to subtransmission (left); detailed diagram of transformer (right)

8.1.4 POINT OF COMMON COUPLING AND POINT OF INTERCONNECTION

The point of common coupling (PCC) is where the power plant electrical system "handshakes" the electric grid. It is the point where the generator facility is connected to the shared portion, or potentially shared portion, of the utility system. The Institute of Electrical and Electronics Engineers (IEEE) 1547 standard establishes this point as the location where voltage and harmonic limits are measured and applied. The PCC is the physical point where a portion of the electric grid under separate administration can disconnect from or reconnect to a portion of the larger electric grid. The term Point of interconnection (POI) is less technical and refers to where the portion of the larger electric grid can connect with or disconnect from a portion of the grid under separate administration.[17] The POIs represent a business demarcation between the utility and the third-party. The PCC represents a physical switching device of the ModPlant that isolates it from the circuit segment (i.e., electric grid) and is controlled by the ModPlant owner (i.e., the project developer). Using a POI is more meaningful to the project developer, because the location of the interconnection is determined by the land boundary.

A transformer and a switch (plus other protective devices) increase the voltage ("step-up") for export to the higher voltage (likely subtransmission) line at the PCC/POI. The connection is AC at 60 hertz (Hz) (i.e., cycles per second). The generation frequency and phase must be synchronized with the external electric grid via the interconnection. The power is based on current and voltage. There are several technical variables to know for the connection, including:

- Generator Nameplate Rating (kW and kilovolt-Amps [kVA]).
- Generator Voltage.
- Type of Service (Wye or Delta).

It is important to know the maximum amount of power that will be exported at any time (units are measured in kW or MW); however, voltage and current are normally slightly out of phase rather than in phase, because of magnetic loads (motors) and other so-called inductances, so the rating must also state kVA.

The planned generator voltage is provided so that the correct transformer model can be purchased. The voltage is based on the potential difference between two terminals, much like that observed with a household battery. The utility (grid) side of the transformer (the high voltage side) is called the primary side, while the generator side is referred to as the secondary side (OSHA). The transformer will be three-phase, because it is most efficient for high-voltage applications, as opposed to the single-phase service provided on the distribution lines to homes. The Wye and Delta connections to do this are named for their geometry and illustrated in Figure 8-5 (shown together for convenience; see GeneralPac[18] for a video). For Wye (the "Y" shape) connections, the terminals are tied together (at the same voltage) on their negative side. For Delta connections, the terminals are connected in series in a triangle configuration so that the points A, B, and C are all at different voltages. When purchasing a transformer, the unit may require three transformers (one for each phase) and they may not be in one enclosure.

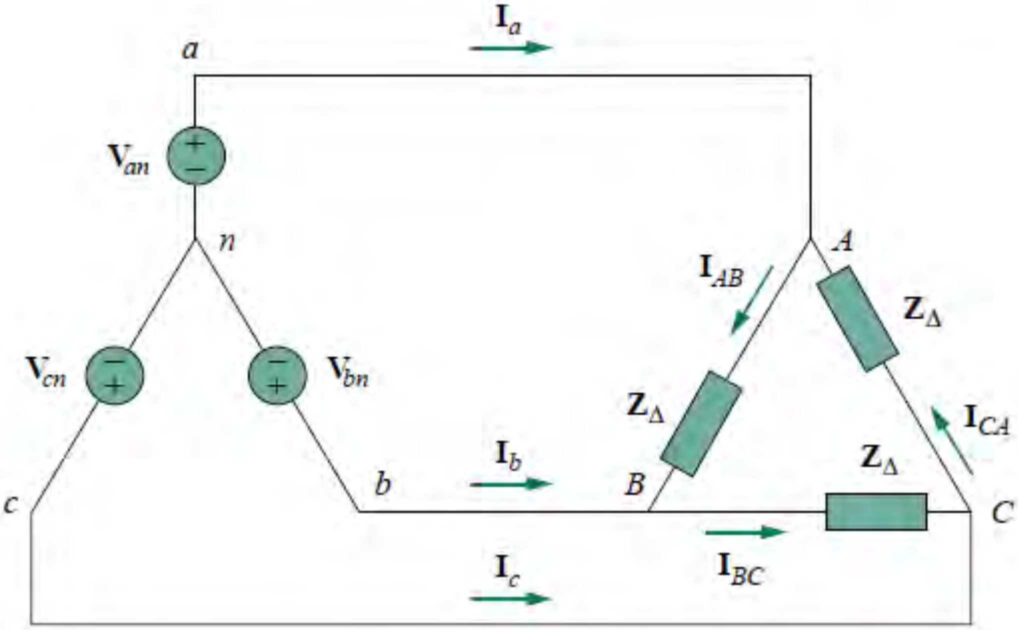


Figure 8-5. Wye (left) and Delta (right) connections

8.1.5 KNOW THE CUSTOMER: THE UTILITY PERSPECTIVE

Utilities are the leading players in the electric industry and collectively responsible for most electricity generation, transmission, and distribution to the public. Site location will determine which utility the project developer works with to interconnect. Utilities are unique legal entities and considered natural monopolies. The nature of their business makes it more efficient for only one utility to operate at a time in any given geographic area. In regulated markets, the government permits the monopoly and instead protects the public interest in access to reasonably priced electricity by heavily regulating electric utilities and monitoring their functions.[19] In deregulated markets, generation and electricity retail are often open for competition. Typically, in deregulated markets it is only the "wires" (i.e., the transmission and distribution grids) that are regulated. The project developer may need to become familiar with the rules of the state/region where they intend to build the ModPlant.

A generating facility sometimes operates off-grid, meaning the load is never powered from the electric grid, and is occasionally done with data centers and industrial loads. A cost-benefit analysis often favors interconnection to the electric grid and the project developer must advise the utility of its plans in this case.

It is conceivable to interconnect to the local electric utility's transmission or distribution system and sell to another customer, such as another utility or large consumer. These complications are reduced through markets that guide the sale of electric power.

8.1.5.1 UTILITY INDUSTRY OVERVIEW

To understand the electric grid, recall that good service and low prices are typically due to competition in a free market. Until the early 1970s, this was inaccurate for the electricity industry in the United States as prices fell due to better technologies (like gas turbines) and increased demand that led to better economies of scale.[20] The energy shocks of the 1970s led to increased prices due to fundamental defects in the industry, including:

- Growing technical complexity.
- Over investment due partly to the nature of incentives to utilities in regulated markets.
- Bureaucracy of the state regulatory process.
- Risk averse leadership, particularly regarding whether an energy investment is both used and useful (i.e., something individuals will buy and something individuals want).
- Low research and development (R&D) investments.

The transition from a regulated electric industry toward deregulation is discussed in the following subsections.

8.1.5.1.1 ECONOMIES OF SCOPE AND SCALE

Historically, electric service is provided by vertically integrated utilities, meaning they offer generation, transmission, and distribution service. This assumes an economy of scope, meaning one firm can provide multiple services more efficiently than separate firms providing each type of service. This historical development hid an essential truth about capital costs. It costs twice as much to build out two systems to move electricity, but the amount of generation is roughly proportional to the demand, not the number of wires (i.e., transmission and distribution is a function of the location and size of generation), but generation is not dependent on the wires.

Utilities have also served relatively large numbers of people to take advantage of "economies of scale," meaning that the average and marginal unit costs of production ($/kW) decline as the output of firms increases. Larger firms are assumed to be more efficient than smaller firms in the electricity business. The historical observation is that electricity provision shows economies of scale—that utilities are "natural monopolies." For example, it does not make economic sense to have duplicate electric grids in the same area, because of the increase in capital costs without providing additional electric service. An analogy is communications, where there is one cable company and one local phone company in a geographic area, but there are multiple ways to obtain cellular or internet service. The tradeoff is regulation for the removal of competition.

A natural monopoly must be regulated to avoid exploitation of customers. In the early 1900s, a consensus emerged that vertically integrated companies should be granted monopoly status within a geographic area in exchange for regulation that obliged them to serve consumers at prices and terms that were regulated by the respective states where these companies operated and essentially guaranteed rates of return to attract capital. This became known as the "utility consensus." Economies of scale were the justification for granting electric utilities franchise monopolies. In past decades, the assumption of economies of scale has been questioned in the generation and retail businesses, but the "wires" segments of the supply chain (i.e., transmission and distribution) exhibit economies of scale and are tightly regulated.[21]

8.1.5.1.2 DEREGULATION

Deregulation started in the 1970s when electricity prices started to rise and inefficiencies became evident. Deregulation introduced incentives for competition to the generation and retail (sales) areas and is applicable in 17 states (at the time of publication). This underestimates the true extent of deregulation, because most large utilities participate in electricity markets, where power and capacity are purchased without explicit reference to the generation source.

Deregulation can also be thought of as restructuring. Most of the time, transmission is regulated by federal authorities, not the states. A well-known breakdown in oversight happened in the Enron scandal in the early 2000s. Markets are indispensable to the function of the electric grid. Utilities had to divest or retire generation assets, maintaining only the "wires business." The power of markets is seen in maps of independent system operators or Regional Transmission Organizations (RTOs).[22] There are seven RTOs in the United States and their management is done by computers, which themselves matured in the later decades of the 20th century. Most of the western United States is uncovered, as is the southeastern United States. This is due to the existence of rural cooperatives and vertically integrated utilities. These utilities tap into other markets for power, except for utilities in the Electric Reliability Council of Texas, which does not have an easy connection to the rest of the United States.

The definition of economy of scale differs according to the public interest in a region. To that end, there are three main types of electric utilities, and the type will impact the rules for interconnection and operations for the project developer. Types of electric utilities include:

- Investor-owned utilities (IOUs).
- Municipal utilities.
- Cooperatives.

8.1.5.1.3 INVESTOR-OWNED UTILITIES

The most common IOUs in the United States are private, for-profit enterprises with stock-based ownership, such as Duke Power Company and Southern Company. These utilities finance new projects through the sale of debt and equity. They are naturally motivated to maximize profits, pay returns to investors, and encourage further investment. They are also the most heavily regulated utilities in the industry. The PUC regulates the prices that IOUs pay for the electricity they purchase from IPPs. This applies to renewable energy and other resources.

8.1.5.1.4 MUNICIPAL UTILITIES

Municipal utilities are created as functions of town, city, and county governments. For example, Kansas (USA) has 118 municipal electrical utilities, which provide distribution services.[23] Few states fully regulate municipal utilities. Most rely on elected officials as "owners" who are publicly accountable for the utility's operations. This ensures that customer rates remain reasonable and that environmental values are considered.

Although municipal utilities are not subject to comprehensive regulation, some general energy-specific laws do apply to municipal utilities. The state's renewable energy standards and the requirement for tariffs encouraging locally owned and onsite generation can apply to municipal utilities of sufficient size. For interconnections, there may be an ambiguity about whether to connect to the municipal utility lines or the regional utility. In Georgia (USA), the electric municipalities and electric cooperatives have agreements with Georgia Power (part of Southern Company) in sharing the transmission grid and jointly owned units in Georgia.

8.1.5.1.5 COOPERATIVES

Electric cooperatives are private, non-profit, consumer-owned utilities controlled by their member-owners through an elected board of directors. Electric cooperatives serve 50% of rural consumers in the United States, largely due to federal legislation. The government subsidizes loans to promote electric development in places where too expensive for IOUs to do business. Most states do not regulate the rates they can charge, because electric cooperatives are non-profit entities directly accountable to their member-owners. The state's renewable energy standard and tariff requirements, which encourage locally owned and onsite generation, may apply.

8.1.5.1.6 LEGAL BASIS FOR INTERCONNECTION

For the gasifier power plant, which is also distributed generation, the export of power at subtransmission voltages or higher is required. The legal framework for this is the Public Utility Regulatory Policies Act (PURPA, Pub. L. 95–617, 92 Stat. 3117, enacted November 9, 1978)—a U.S. Act passed as part of the National Energy Act. This was extended by EPACT 2005, which updates the rules for interconnection. PURPA and later legislation made it possible for IPPs to profitably take part in electrical energy markets.

EPACT 2005 adds five new federal standards to PURPA Section 111(d):

- (11) Net Metering (see EPACT 2005 Sec. 1251 for details).
- (12) Fuel Sources (see EPACT 2005 Sec. 1251 for details).
- (13) Fossil Fuel Generation Efficiency (see EPACT 2005 Sec. 1251 for details).
- (14) Time-Based Metering and Communications (see EPACT 2005 Sec. 1252 for details).
- (15) Interconnection (see EPACT 2005 Sec. 1254 for details).

The key statement is that interconnection is available upon request: "Each electric utility shall make available, upon request, interconnection service to any electric consumer that the electric utility serves." Thus, based on PURPA, it is not a question of whether a generator can interconnect, but under what conditions it will do so.

Some states also have deregulated their electricity markets, particularly in retail (point of sale) and generation. Energy policy allows for the interconnection of small (less than 10 MW) and midsize (10 to 50 MW) generators whether the markets are regulated or deregulated. The most important concept is the public interest; PURPA was the first act to establish this concept and the Energy Act of 1992 and EPACT 2005 have expanded upon it (see Appendix C).

8.2 DEVELOPMENT: SITING AND SIZING THE PLANT

Project developers should consider all the site design/engineering issues that relate to the interconnection. Beginning with site selection, the following interconnection-related questions should be reviewed and considered:

- Is the site accessible to maintenance trucks?
- How many acres are available for development?
- Is there existing natural gas infrastructure? (This relates to fuel supply, but also in the case of the ModPlant, gasification.)
- What is the current electric grid infrastructure in the area? Is there a subtransmission line within sight that could be interconnected? (If there is an onsite substation, the site is potentially more valuable.)

The project developer is responsible for the costs of building new distribution or subtransmission lines or any upgrades necessary to increase transmission capacity or improve system protection. It is financially prohibitive to design and build long distribution lines, transmission lines, or other upgrades. All that needs to be built or upgraded for a favorable site is a substation and a few lines of modest length.

The project developer is responsible for obtaining all permits necessary to build any new lines (not the utility). When working through permitting for project construction, the project team should consider that permits may also be needed for transmission work.

Maintenance access is best addressed early in the process, as the available area for the plant infrastructure will be reduced by the area needed for at least one road and a parking area. Once these questions have been addressed, the project developer can estimate capital costs, operations and maintenance (O&M) costs, taxes, interest, and other costs. In combination with the anticipated revenue and assumptions about the "time value of money," these estimates draw on the field of engineering economics, which is used to evaluate a large capital-intensive investment. (For more information, consult the definitions[24] and the informative series of slides[25] that discuss the use of an assumed discount rate to calculate the time value of money in the context of costing the development of infrastructures like power plants. In investments, most of the costs are incurred upfront and revenues occur over several years. Formulas with the assumed discount rate provide insight into the ROI.)

8.2.1 SITE CONSIDERATIONS – THE LOCAL VIEW

A desirable site is flat for easier transportation (e.g., road or rail) and must have sufficient land for the development. Besides the plant, there should be space for the transformer(s) with safety buffers (e.g., a small substation consistent with Underwriters Laboratory [UL] 1741 standards). The acreage needed will scale with the transformer size/rating. For example, if the transformer steps the voltage up to 34.5 kV, it will require less buffer and protection than if it steps up to 69 kV. In addition to the transformers for interconnection, there also needs to be room for parking, perimeter fencing, a control center, the gasifiers and fuel supply, generators, a gas cleanup module, and a small gas mixing station.

One approach is to leverage the abundance of industrial sites currently connected to subtransmission lines of 69 kV or 34.5 kV. From such power lines, the export of power to the extra high voltage (EHV) system is already assured under normal electric grid operating conditions. An example of an industrial site is a retired coal plant, which already has an electrical interconnection. Other sites include former pulp and paper plants, railway yards, and steelmaking plants.

Environmental conditions should also be considered with respect to local regulations and nearby communities. For example, noise management is important in the design phase, as 60 Hz noise and its overtones (e.g., 120 Hz) can aggravate plant personnel and communities neighboring the plant.[26] Sometimes, the plant must be isolated from local bedrock to avoid sound transmission underground.[27]

The following subsection addresses resource planning, which builds confidence that utilities can serve future loads. The project developer must ensure the planned generation facility meets a planned or identified need. A top priority in the planning process is an increased probability of revenue in the future. The project developer's resource planning is also informed by understanding the utility's point of view.

8.2.2 FINDING A GOOD LOCATION: INTEGRATED RESOURCE PLANNING

The utility plans 30 years out through a process called integrated resource planning (IRP). Plant retirements potentially open the opportunity for other plants to supply the electric grid. Utilities must meet peak load or risk brownouts and rotating blackouts, and they address this risk with a planning reserve margin to measure the amount of generation capacity available to meet expected demand within the planning horizon (e.g., 30 years).[28] The utility will determine the future peak and energy demand forecast for 30 years, planned retirements and planned additions in capacity, and the gap in peak and planning reserve margin. These determinations can clarify whether there is a need for new generation by the utility and the size of the generator (i.e., power, in MW or milli-volt amps [MVA]).

An example of resource planning is the PJM Interconnection LLC ("PJM," an RTO spanning multiple utilities—upon adding Maryland; the company was established in 1927 and reorganized in 1956 as Pennsylvania-Jersey-Maryland Interconnection) capacity market, called the Reliability Pricing Model, which ensures long-term electric grid reliability by procuring the appropriate amount of power needed to meet predicted energy demand three years in the future. Under the "pay-for-performance" model, power plants must deliver on demand during system emergencies or owe a significant payment for non-performance. This is a kind of insurance policy—for a small, additional cost (payment to plants that perform well)—enables consumers to have greater protection from power interruptions and price spikes during weather extremes. By matching power supply with future demand, PJM's capacity market creates long-term price signals to attract investments to ensure adequate power supplies.[6] These price signals make investing in new generation potentially lucrative. PJM's market uses wholesale competition (an example of market deregulation).

The details of interconnection, standards, protocols, electrical protection, and handling electric and gas infrastructure are addressed in Sections 8.3, 8.4, and 8.5.

8.3 INTERCONNECTION PROCEDURE

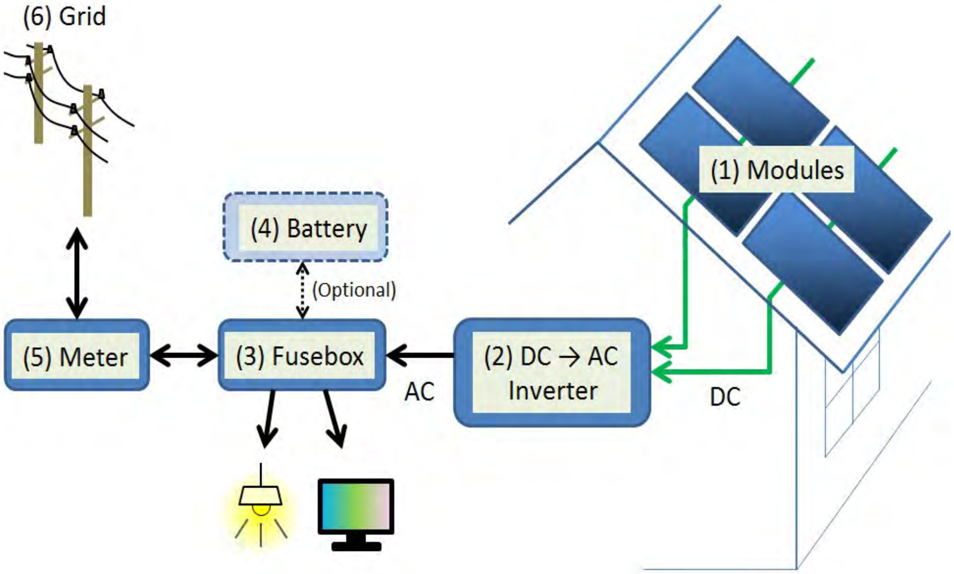
For new electric grid connections, major utilities provide guidelines for the interconnection based on the size of the connection (IEEE standard 1547). The voltage, frequency, and phase must match that of the utility subtransmission or transmission line, which can be achieved with a transformer, switch, and other equipment to synchronize the signal to the electric grid. The utility may provide advice for the plant setup, which includes the integration of multiple generators, fuel supplies, and energy storage. The vendor's job is to ensure that the right communications protocols are used for the controls (see Section 8.3.3).

Once the plant complexity is understood, the project developer will negotiate the details of the interconnection with the utility. The interconnection will consist of a simple substation with a transformer and protective devices, such as relays, reclosers, and fuses in the substation and on surrounding lines. Standard control involves a Supervisory Control and Data Acquisition (SCADA) system with a remote terminal unit (RTU) or programmable logic controller

(PLC) connected to its interconnection breaker. An RTU is a microprocessor-controlled electronic device that interfaces objects in the physical world to a distributed control system (DCS) or SCADA system by transmitting telemetry (literally "remote measure") data to a master system and using messages from the master supervisory system to control connected objects. A PLC has more general capabilities and is superseding the RTU in new developments.[29]

Since the transformers require AC current (if DC generation, like fuel cells or solar power, is installed), an inverter is necessary to make an AC signal at near 60 Hz with appropriate phase. Grid-tie inverters match phase with the utility-supplied sine wave, a process analogous to the synchronization of rotating generators to the electric grid, except that the DC signal is converted to AC first. Additionally, the inverters are designed to shut down automatically upon loss of the connection for power transfer to the transformer.

The preliminary design process consists of drawing so-called "one-line diagrams," which indicate the directions that a single-phase of electricity would flow through circuits. This shows the logic of three-phase electricity that would flow through the interconnection, destined for industrial, commercial, and residential loads. A fuel cell installation will look similar to a solar installation with the solar panels replaced by a fuel cell, because fuel cells and solar panels both generate DC (Figure 8-6).[30] Solar generation can play a role in the ModPlant or could be connected to the service station circuit, but the principle applies that any DC generation source requires an inverter to convert the power to AC at electric grid frequency.



Used with permission from Fraas and Partain[30]

Figure 8-6. Solar power generator with connection to electric grid

8.3.1 INTERCONNECTION PLANNING

The electric grid interconnection process is determined by the rules of the utility, but the process can be managed by planning and tracking documents, including the following steps:

1. Project definition.
2. Interconnection application.
3. Application submittal.
4. Application review.
5. Interconnection agreement.

The project definition is essential to minimizing unexpected costs. Even certification of ownership or site control is essential, because the utility will not review the interconnection application without this information. The capacity, meaning power to export (in kW/kVA or MW/MVA), is essential to knowing which procedure must be followed. The project developer should know the ideal location to connect to the electric grid (the POI/PCC) to determine which utility the plant interconnects. More detailed planning involves whether the power will be exported (in the absence of large loads, yes) and what type (induction, synchronous) and model of generator will be used. The generators in the scope of this chapter are all gas-powered. The consulting engineers will develop a one-line diagram of the project for the application. Figure 8-3 (right side) is an annotated one-line diagram for a solar installation. Most engineering software will be able to create diagrams in standard format.

Once the project is defined, the next step is applying for interconnection. To obtain an interconnection request application, the project developer will contact the utility company that owns the distribution line, transmission line, or substation at the proposed POI. Most large utilities provide downloadable applications (Xcel Energy, for example[31]) on their websites and usually have an interconnection office or individual to contact. These forms can be complex and some questions can only be answered by a consulting engineer. Typical questions include the following:

- What is the project size (MW) and variability around that amount?
- What type of generator will be connected? (induction, synchronous)/model/size)
- Where will the connection to the electric grid be made? (determine PCC/POI)
- To which utility will the interconnection be made? (location dependent)
- What rules apply to the project? (usually FERC, could be AR 521 or similar for projects less than 10 MW)
- Will some or all the generated power be exported to the electric grid?

Tiers are used by regulators to group distributed generators by project size. For example, Tiers 1 to 3 are for small generators, typically ending at 1 to 2 MW. In Florida, for example, Tier 3 is less than 2 MW.[32] Tier 4 maxes out at 10 MW. The anticipated power output of a ModPlant is greater than 10 MW, but some ModPlants may have output less than 10 MW and qualify for special treatment. Any generator with output less than 20 MW qualifies for the FERC study process (see Section 8.4.1).

8.3.2 STANDARDS FOR DISTRIBUTED GENERATION: UL 1741 AND IEEE 1547

Each of the procedures reference technical standards throughout. Standards are established industry requirements for system design and operation. UL 1741 and IEEE 1547 are the two standards relevant for distributed generation equipment. UL 1741 is the "Standard for Inverters, Converters, Controllers, and Interconnection System Equipment for Use with Distributed Energy Resources." Generator choices and power plant layout will be partly driven by these standards. If a well-proven generator is chosen and a competent engineer assists in the design, the project developer may not need to learn the details of these standards. The use of new or experimental generators may require extensive testing to verify adherence to these standards. UL 1741, developed by UL, requires interconnection equipment to meet construction constraints, protect against risks of injury to persons, prescribe rating and marking, and set specific distributed resource tests for various technologies.

IEEE 1547, the "Standard for Interconnecting Distributed Resources with Electric Power Systems," is the standard for connecting distributed resources to the electric grid (see Appendix C). FERC has approved the PJM Small Generator Interconnection Standards (two sets: one for 0 to 10 MW and one for 10 to 20 MW) for Interconnection that incorporate IEEE 1547. Note that the units can be MW or MVA for these purposes. IEEE 1547 describes the design technical and testing requirements, which guarantee that operation of the electric system is not degraded in normal operating conditions. Newer or experimental generators may require extensive testing for potential power quality issues and have anti-islanding capabilities. Islanding occurs when the distributed generation is running, but the POI is isolated in an abnormal way from all or part of the nearby electric grid. The control system must locate loads, begin a shutdown, or transition to minimal power output ("spinning reserve"), potentially shunting power to ground as a last resort. Shunting uses a temporary short-circuit to reroute power from where it can damage components to a place, usually ground, where it will not. This is the principle behind the familiar ground-fault current interrupter circuits found in new, electrically upgraded homes.

8.3.3 SYSTEM INTEGRATION AND CONTROL

System integration is defined in engineering as the process of incorporating component sub-systems into one system to work toward achieving one or more objectives, which, in this case, is exporting power to the electric grid and local loads. An example is connecting fuel systems to the generators, which will require pumps, piping, valves, and simple controls. Modern systems rely on communications for continuous operation, which relates to the physical connections and communication protocols. There are two relevant levels of communications integration: connecting two communication systems with disparate protocols and connecting equipment from different suppliers or different equipment from the same supplier with the same protocol.

New equipment adds to the complexity of a plant. Metcalfe's Law is the heuristic for estimating the value of increasing the complexity of a network and applies to the cost of integration. Metcalfe's Law states that "the value of a network increases quadratically—proportionately to the square of the number of its nodes—while costs would, at most, grow linearly."[33] It is evident that while multiple generators, battery storage, and control systems increase the flexibility and the potential value of the plant, they also potentially increase costs in a complex fashion, whether in construction, integration, or maintenance.

The two relevant protocols are the legacy Distributed Network Protocol 3 (DNP3) and the modern International Electrotechnical Commission (IEC) 61850. In a substation, the main challenge is connecting newer, more capable IEC 61850 automation systems to legacy DNP3 systems. For a new installation, this complexity can be skipped by using only the more modern IEC 61850 protocol. For gas turbines, the major supplier can provide control systems compliant with IEC 61850. As fuel cells and battery energy storage systems (BESSs) mature, similar compliance to newer protocols is expected. Note that loads may communicate using DNP3, so interfaces to IEC 61850 may be needed.

However, the content and structure of the data must be determined and implemented even with the same protocol. Common use cases and implementations are available. One option for simplifying system integration is the Open Field Message Bus (OpenFMB) interoperability framework,[17] which can facilitate the connection of multiple gasifiers, turbines, fuel cells, energy storage devices, and other DER. This system integration framework enables more interoperability of different control and protection subsystems, and may reduce the cost overhead (i.e., more upfront planning and design effort could pay off in cost savings, which will increase ROI).

The master controller must integrate all the individual pieces, because each major piece of equipment needs to be protected. Engineers from different vendors will need to ensure that each subsystem "plugs and plays" with the whole. The process can be simplified by incorporating best practices during the project design phase.

Depending on location, the protective equipment will be installed in an equipment cabinet for physical security purposes and safety. Cybersecurity is always an issue and access control, user authentication, intrusion detection, and responses to successful attacks must be considered. Whatever integration framework is chosen, it is essential for scaleup (to many gasifier-fueled power plants with similar design) to document the process of implementing the power plant so that it can be replicated. The design must be modular (i.e., built of plug-and-play components) with best practices for protecting the transformer, circuits, and gas handling systems.

Documenting the integration process is key to making the solution modular, scalable, and able to be replicated in different locations. Scalability and replicability are the two key features of the electric grid connection module.

8.3.3.1 SYNCHRONIZATION AND CONNECTION

The connection consists of a transformer and hardware to synchronize the signal to the electric grid. The most probable transformer will be a step-up transformer from 4 to either 69 kV or 34.5 kV (Figure 8-2). DC connections require the use of an inverter to produce an AC signal of 60 Hz with appropriate phase. Grid-tie inverters match the phase with that of the utility-supplied AC signal. Additionally, the inverters are designed to shut down automatically upon loss of electrical connection to loads and/or the interconnection, depending on the control system determination.

8.3.3.2 COMMUNICATIONS AND PROTOCOLS

The control system is an "overlay" of the physical power plant. For example, relays on the generation/storage grid control the electricity and transmit information to an operational technology (OT) network. Customization can be done by site engineers either with a graphical interface or by command line programming, according to the relevant protocols.

8.4 THE MODPLANT: SIZING AND BUILDING THE INTERCONNECTION

The final step in plant and interconnection planning is to determine the size/ratings for all the equipment, from generators to transformers. The amount of gas produced by the gasifiers determines the size of the gas turbines, microturbines, or fuel cells that will use the gas for fuel after impurity cleanup. The resulting electric power can serve local loads on the distribution system and provide power to the electric grid via the interconnection. The first interconnection will likely be to subtransmissions, with voltages of 34.5 kV or 69 kV. Table 8-1 shows transmission voltages, types, and usual applications.

Table 8-1. Typical applications for power transmission[16]

TRANSMISSION VOLTAGE	TRANSMISSION TYPE	APPLICATION	VOLTAGE LEVEL
240 V	Distribution	Home/Office	Station Service
4 kV or 10 kV	Distribution	Neighborhood	Feeder
34.5 kV	Subtransmission	Industrial	High Voltage
69 kV	Subtransmission	Industrial	High Voltage
115 kV	Transmission	Intermediate Distance	High Voltage
138 kV	Transmission	Intermediate Distance	High Voltage
161 kV	Transmission	Intermediate Distance	High Voltage
230 kV	Transmission	Intermediate Distance	High Voltage
345 kV	Transmission	Long Distance	Extra High Voltage
500 kV	Transmission	Long Distance	Extra High Voltage
765 kV	Transmission	Long Distance	Extra High Voltage

Gasification is capable of more than providing fuel for the ModPlant. Excess gas can be diverted to natural gas pipelines according to the relevant gas interconnection guidelines. This flexibility reduces the constraint on the ModPlant, allowing it to be more efficient and reliable. Please reference the following case study for an example of how the technical aspect of interconnection planning proceeds, which is a costly portion of the planning process for determining key technical parameters.

Case Study: Connecting to the Electric Grid in Xcel Energy Territory

Xcel documents their interconnection process online.[34] *A 10-megawatt (MW) connection or less avoids the California Public Utilities Commission (CPUC) 3900 Rules derived from the original Public Utility Regulatory Policies Act (PURPA), which made interconnection of modest-sized generators feasible. After the cost of the transformer, the main issue is the cost associated with putting protection on the electric grid side of the point of interconnection (POI)/point of common coupling (PCC). This cost is paid for by the interconnecting project developer/owner and must be included in the engineering economics calculation to determine if the proposed system is economically viable. The process for a 10-MW connection (or less) makes the interconnection a smaller part of the overall cost, but a 20-MW connection or* *larger may be more cost efficient if sufficient generation capacity is available. The protective devices required include reclosers, relays, and fuses on both the utility and Modular Plant (ModPlant) side of the interconnecting substation. The relays are computerized because of the smart grid, leading to finer control and better dispatch of the generated electricity. The equipment is commercial off-the-shelf (COTS) and the designs are reusable given the nature of the plant (gasifiers + generators + energy storage + controls) and the constraint on maximum power (approximately 20 MW) based on utility interconnection costs. A current industrial interconnection may also help with cost efficiency, though upgrades may be necessary.*

Based on a survey of commercially available transformers, with typical specifications provided by Stevens Engineering,[35] *one potential scenario is a 4-kilovolt (kV) generation system exporting power to a 69-kV line at up to 24 MW with two transformer units, each of which contain one transformer for each of the three phases generated. A physical quantity limiting power transfer in this situation is the current of 3,000 amperes (A) necessary to travel into the transformer windings on the generator side; hence two side-by-side transformer units can be used in parallel. Specifications for sound (i.e., noise) levels, fluid types/properties, and other aspects are available for each transformer.*

This process can be generalized with minor adjustments for any utility using a study process. This process (described in the next section) allows the project team to enhance communication with the utility, potentially accelerating the interconnection process.

8.4.1 FEDERAL ENERGY REGULATORY COMMISSION STUDY PROCESS

The FERC study process is used for projects up to 20 MW, which FERC refers to as “small generator interconnection.” This process is part of an overall project management process, including the design, planning, site and circuit development, procurement, and permits. The interconnection for power is distinct from a PPA. The latter is a financial contract, whereas the former concerns physical access to the electric grid. The PPA often determines the primary revenue stream for the project and is negotiated between the project and utility where power is sold.

In addition to purchase rate, a key element to the agreement is the start date. If the system is not fully interconnected to the electric grid by the agreed upon start date, the project developer will likely be responsible for penalties associated with not delivering. This is a surprisingly common issue projects face, and the project developer should be aware of the dates and obligations in all relevant contracts and how they relate before construction begins.

Although the project developer will communicate with representatives at the interconnecting and purchasing utility, these individuals will be different people in different departments and may be at different companies (i.e., if the project developer chooses to wheel power to sell to a utility other than the interconnecting utility). Project developers should not assume that other parties are communicating with each other about the project. It is the responsibility of the project developer to communicate and connect all the pieces.

There is more information about PPAs and interconnections in an online Interconnection Guidebook.[36] An example of a detailed interconnection study of a 12-kV, 3-MW solar interconnection is included in a “Generation Interconnection System Impact Study Report.” The report lists equipment upgrades on the distribution system and other technical specifications for the project. It was likely prepared by an interconnection applicant from a template provided by the utility and with some consulting engineering or utility assistance. The single-line diagram shown in Figure 8-3 (right) includes a large gap (8 miles) between the generation unit and the substation, which will cause some excess loss of power.

8.4.2 DESIGN

Along with documents, visuals (e.g., figures, schematics, diagrams) are essential for understanding the ModPlant and how it will interface with the electric grid. A one-line diagram shows the structure and behavior of the generation plant for one phase, even though the plant can be generating three phases of power. The electric grid interconnection is represented by two overlapping circles in Figure 8-3 and labeled “transformer,” because it is the key piece of equipment. The left-hand side of the figure shows the overall design of the 4-kV loop with the generators. There will be three feeders/lines/connectors on the ModPlant bus. One has one or more turbines, the second has a fuel cell, and the third has battery energy storage. The master controller connects to all three and the POI/PCC.

A 4-Pole Synchronous Gas Turbine comes with transformers in-built to step up its output voltage to up to a value of 15 kV. Depending on the design and choice of turbine, the voltages of the fuel cell and battery energy storage will also have to be increased in preparation for the step up to 34.5 kV or 69 kV.

Auxiliary power is used to power the master and vendor-based controllers. This “station service” of 240 volts (V) can be provided by a combination of solar power, energy storage, and/or a connection to a local distribution system or a secondary circuit powered by the plant bus with a 4kV-to-240V step-down transformer. If the solar power installation is sufficient, it can also export power onto the main electric grid via the distribution system (optional).

8.4.3 PROTECTION

During the interconnection negotiation with the utility, the utility will insist on the protection of its assets as a condition of the interconnection process and at cost to the owners of the planned facility. The negotiation with the utility and providing accurate information about objectives and designs are essential for developing a stable and reliable connection to the electric grid.

The ModPlant consists of a set of gasifiers, plus generators, energy storage, and associated control systems. It is its own ecosystem, but in its simplest form, cannot island (i.e., disconnect from subtransmission and distribution grids) without the generators going offline or into spinning reserve (i.e., online and able to quickly increase output to meet changes in demand or restoration of the interconnection).[37] The system could island by adding loads, making the system more complicated from a control perspective. An additional connection to the distribution system at a lower voltage could provide additional flexibility and income streams.

The key electrical protection issue is how to protect the electric grid from the gasifier-fueled ModPlant, and how to protect the ModPlant assets from an outage on the electric grid or one tripped in the plant itself. The lowest cost option is for the ModPlant to interface with the electric grid with power moving only from the plant onto the electric grid for export to customers and a separate connection to the distribution grid to receive power for operation of the lights and control systems in case of the need for maintenance or a black start (i.e., from power off to fully synchronized with the electric grid). This separation of station service from the generation circuit (i.e., plant bus) is already standard at power plants. The station service is at 240 V and this can be provided by a parallel connection to the subtransmission or to a local distribution line (4 to 10 kV) with a familiar cylindrical pole-mounted transformer or a neighborhood transformer.

Each connection must have standard electrical protection. These protections do not need to be more complicated than what currently exists in neighborhoods, but all connections and loads must be accounted for in engineering plans.

8.5 PERFORMANCE METRICS

The ModPlant performance is measured against the metrics of reliability, availability, efficiency, safety, and resilience. The benefits of the interconnected power plant are increased by smooth running machinery and threatened by external and internal factors. A well-integrated system will protect itself and augment the quality of the exported electricity.

8.5.1 RESILIENT RESPONSE TO OUTAGES

Outages are important and must be accounted for in the design. They behave like shockwaves, in that an outage on the electric grid immediately forces the ModPlant bus to reroute power away from the POI. Not long after, the gasifiers will need to shut down or reroute their power. Outages of ModPlant assets require phased equipment shutdowns or winding down to minimal energy states and shunting some energy to ground as a last resort.

There are two types of plant outages: planned and unplanned. Planned outages due to maintenance follow a set, agreed-upon procedure and schedule. If the gasifiers shut down, then the generators must shut down if there is no connection to the local natural gas grid. But, if there is such a connection, the plant will make more profit by having increased uptime (i.e., higher capacity factor).

Unplanned outages require quick thinking and action. If a generator is down, some or all gasifier output needs to be diverted to the local natural gas system or local storage. After a generation plant outage, the plant needs to be ramped up and resynchronized to the electric grid.

Most plant outages are unplanned. For such a utility outage, the power plant will rapidly become an island. In this islanded state, any power produced needs to be stored onsite in a battery or other medium. When the storage medium reaches a critical level of charge (e.g., 80 to 90% for batteries), the generators need to shut down if the electric grid interconnection has not reopened. Any gas produced by the gasifiers will need to be cleaned up and either stored in tanks or diverted to the external natural gas system. Running generators at a minimal speed and shunting power to ground is another option (if it is only temporary).

A battery energy storage system (BESS) can make the powering up faster and less costly. In a restart, the generators/engines and alternator communicate with one another to reach generator operating parameters. When paralleling an oncoming generator, engine speed (generator frequency) is controlled by the engine electronic control module (ECM). The main control system interfaces with the generators and their control systems in parallel.

8.5.2 PROCESS CONTROL REQUIREMENTS

The plant control systems maximize plant performance against the five metrics of reliability, availability, efficiency, safety, and resilience. A proper control strategy includes an analytical system to interpret the various sensor measurements. Although much of the control is automated, plant personnel should have experience with the controls and operational procedures.

The generators, electrical balance of plant, and the interconnecting substation have their own control systems. Planning, communication, and engineering of the controls are required, because products will come from multiple vendors, and the complexity comes from four functional systems: the generation plant circuit, the interconnection, the service station circuit, and the gasifier plant and piping (fuel plant). The bulk of planning is completed by the consulting electrical and mechanical engineers in consultation with the chemical engineers when selecting and installing the gasifier.

Gasifier operation requires technical skill informed by research on gasifiers and other chemical systems. After installation, a chemical engineer should be available on a contract/emergency basis.

8.5.2.1 GAS AND ELECTRICITY FUNCTIONAL AND PHYSICAL SEPARATION INCREASE RELIABILITY AND SAFETY

Gasifiers produce fuel (syngas), so it is important to distinguish between their operation and that of the generators that use the fuel. Ideally, the syngas is used immediately upon cleanup to power the generators, but this is not always practical. Instead, there should be separation of the gas infrastructure from electrical, which will minimize explosion risk and simplify maintenance. There will be one line out from the gasifiers to a mixing area where fuel appropriate to the turbines and fuel cells can be prepared and later piped in (Figure 8-3, left). The mixing area will have appropriate gas storage (i.e., storage tanks) to ease the control constraints on fuel supply/demand. Note that with zero available gas storage, supply has to equal demand. There should be enough gas for a two- to three-hour supply for the generators to allow workers to shut down the system in case of a gas shortage and a 24-hour supply to allow a simple repair or restart a gasifier as part of the system requirements.

8.5.3 PROCESS AND OPERATOR SAFETY REQUIREMENTS

Operations and safety are addressed through precautions and maintenance. Each piece of heavy equipment has fire, arcing, or leak hazards. The insulating fluid for the transformer is one example of a leak risk that can be addressed through routine maintenance. The turbine, fuel cells, and battery electric storage systems all provide arcing and fire hazards. Figure 8-7 illustrates a high-voltage tower that temporarily faults to ground and a human is in the loop. This is analogous to the situation with any device at high voltage when an object or a human is in the path between it and its ground connection.

As newer technologies, gasifier processes are proven from R&D experiences. Operator safety is based on standard training and awareness of high voltage and temperature areas of the plant. Procedures based on science and regulation can be developed in consultation with a chemical engineer. The negotiation with the utility for the interconnection will involve safety at each stage. An additional safety issue is workers on the utility grid that will turn off the switch to the power plant during grid maintenance, so the plant can cycle down (as discussed previously).

The gas supply must be insulated from any spark sources. This is common practice for liquid-fueled generators, such as certain turbines and fuel cells. Any fuel mixing and storage will occur outside a buffer zone. Fire extinguishers, coats, axes, and additional fire gear must be available at convenient access points in the plant. Key portions of the plant can be covered with partial roofing to reduce high-spot temperatures in the ModPlant. Solar panels on this roofing can support the energy storage systems in the balance of plant.

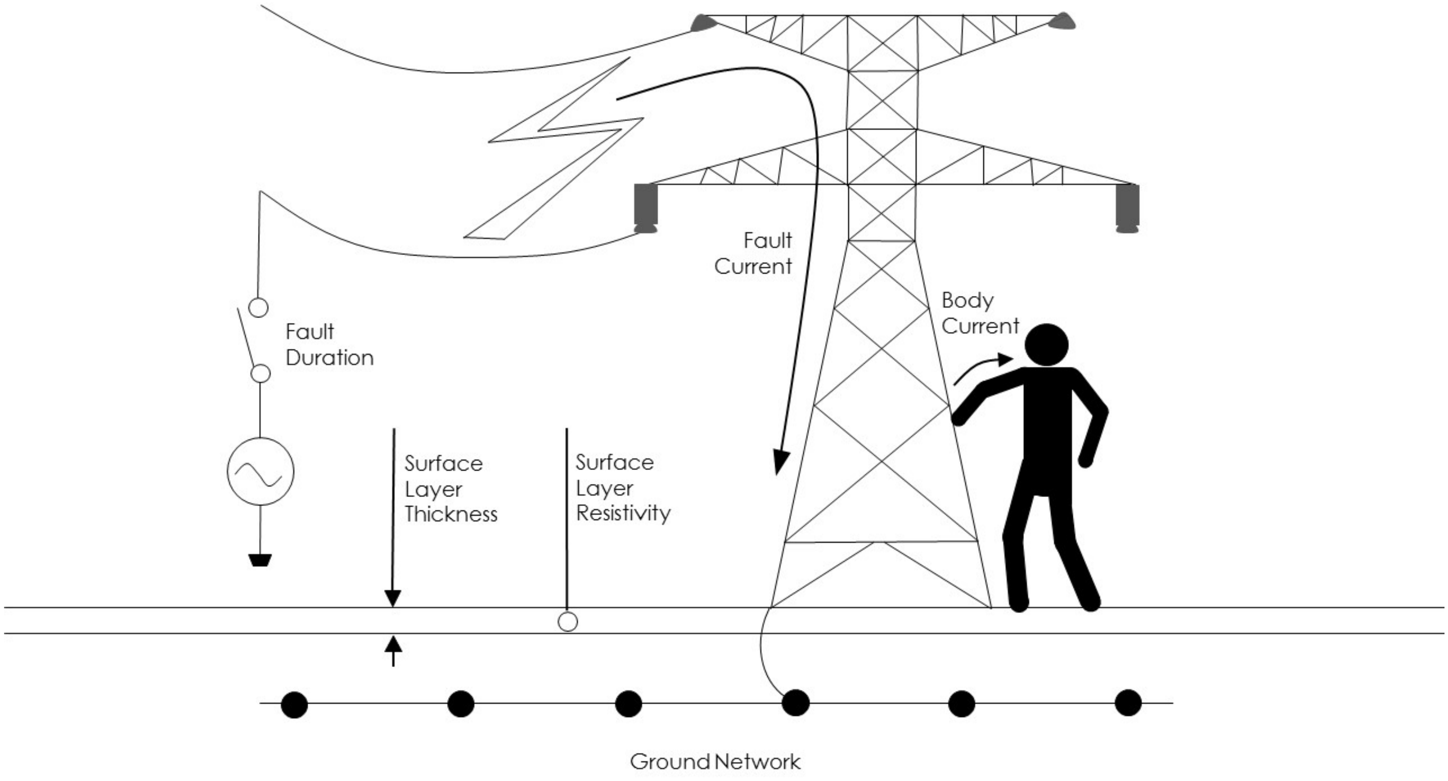


Figure 8-7. Touch potential; protection from hazardous differences in electric potential[38]

8.5.4 WORKING WITH VENDORS AND UTILITIES TO MAKE A MICROGRID

Utilities and vendors can assist in extending the ModPlant into a microgrid. The plant can serve local customers and export energy to the grid with the addition of a load connected to the ModPlant by a distribution connection. The tradeoff for this addition is increased control system complexity due to varying distribution customer loads and the potential for a distribution outage. To address this complexity, Duke Energy's OpenFMB implementation[39] showcases low-latency microgrid optimization and seamless islanding use cases using a variety of wired and wireless communications technologies that leverage open Internet of Things (IoT) publish-subscribe (pub/sub) protocols and industry-standard data modeling structures based on IEC's Common Information Model (CIM). The system vendor may also provide similar capabilities.

The ModPlant has a variety of generators, energy storage, and gasifier(s), resulting in similar management needs as DERs and microgrids (Table 8-2).

Table 8-2. ModPlant management needs

MANAGEMENT NEEDS		
DER CIRCUIT SEGMENT (POWER LINES) MANAGEMENT	PLANT MANAGEMENT	MANAGEMENT SERVICES
Coordination of PCC and POI Voltage, Frequency, and Power Factor (related to phase between voltage and current) Support Solar Smoothing (optional) Peak Power Management (e.g., shaving/shifting to optimize for market export) Volt-Var Management (this will impact the efficiency of exporting power to the electric grid)	Asset and Operations Optimization Unscheduled Islanding (covered under “Resilient Response to Outages”; see Section 8.5.1) Island Reconnection (black start, synchronization; see Section 8.5.1)	Cybersecurity Provisioning

8.6 OPERATIONS AND MAINTENANCE

When the turbines, fuel cells, and battery storage systems are integrated into the system and operating, they provide real time metering for the gross energy output (in MWh and megavolt-amp-reactive-hours [MVARh]) of the generator and the station service to the master controller. With the real power (MW) known, the various phases between voltage and current at various parts of the plant are measured. The temperature, humidity, and other environmental variables should be measured for O&M purposes. The utility grid operator receives information on the exported energy and power, keeping both the power plant and the utility grid stable and enabling the power generated to have market value.

Each piece of equipment has a maintenance schedule; however, condition-based maintenance allows for deviations to the schedule based on monitoring with sensing overseen by the operators and engineers looking for unusual behavior. This is a proven method for maintaining safety and reducing costs over time.

8.7 SUMMARY

The electric grid is designed to allow new sources of power to interconnect, including IPPs. With a new substation or upgrades consistent with a negotiation with the regional utility, the power plant can export power to the electric grid and make a profit for the operator and support local employment. The reuse of protection and control schemes based on documentation from the systems integration phase can make these systems modular and scalable for deployment across the United States and the world. The separation of key electrical circuits from one another and from gas pipelines maximizes operator and system safety.

Battery energy storage and gas storage make cycling the plant down (or off) and up (or on) during external and internal outages safer and more efficient. Careful planning and design, in consultation with consulting engineers, will lower investment costs and potentially increase future revenues, increasing ROI.

Endnotes

1 Generator Source (2020) "Paralleling generators for large power plants," https://www.generatorsource.com/Articles/Power-Sources/Paralleling-Generators-for-Large-Power-Plants.aspx, accessed May 15, 2020.

2 Energypedia (July 11, 2018) "Independent Power Producers (IPPs)," https://energypedia.info/wiki/Independent_Power_Producers_(IPPs), Accessed May 15, 2020.

3 California Public Utilities Commission (August 26, 2014) "Electric Transmission Fact Sheet," https://www.cpuc.ca.gov/Environment/info/aspen/cltp/archive/Files_8_26_14/_2ElectricTransmissionFactSheet.pdf, Accessed May 15, 2020.

4 Energy.gov (November 18, 2014) "The War of the Currents: AC vs. DC Power", https://www.energy.gov/articles/war-currents-ac-vs-dc-power, Accessed June 25, 2020.

5 PJM Learning Center (2020) "Market for Electricity", https://learn.pjm.com/electricity-basics/market-for-electricity.aspx.

6 PJM Learning Center (2020) "Capacity markets (RPM)", https://learn.pjm.com/three-priorities/buying-and-selling-energy/capacity-markets.aspx.

7 F. Gruffaz (September 20, 2016) "Grid Power Quality", https://blog.se.com/power-management-metering-monitoring-power-quality/2016/09/20/grid-power-quality/.

8 Tyco (n.d.) "Watts, Vars and VA," http://www.cromptonusa.com/Watt_Var_VA.pdf, Accessed July 13, 2020.

9 Electricity Advisory Committee (September 29, 2016) "The Value of a VAr – Perspectives on Electric Grid Voltage Support", https://www.energy.gov/sites/prod/files/2016/10/f33/EAC%20-%20Value%20of%20a%20VAr%20-%20Perspectives%20on%20Electric%20Grid%20Voltge%20Support%20-%20September%202016.pdf, Accessed June 3, 2020.

10 E Online (Jan-Feb 2010) "Volt-Var Optimization reduces peak losses, peak demands," https://electricenergyonline.com/energy/magazine/466/article/Volt-VAR-Optimization-Reduces-Losses-Peak- Demands.htm, Accessed May 19, 2020.

11 T&D World (March 6, 2018) "Power quality matters", https://www.tdworld.com/grid- innovations/article/20964027/power-quality-matters, Accessed June 24, 2020.

12 EIA (May 9, 2019) "Energy use in homes", https://www.eia.gov/energyexplained/use-of-energy/electricity-use-in- homes.php, accessed May 19, 2020.

13 Energy.gov (October 2012) "The U.S. Department of Energy's Microgrid Initiative", https://www.energy.gov/sites/prod/files/2016/06/f32/The%20US%20Department%20of%20Energy's%20Microgrid%20Initiative.pdf, Accessed June 25, 2020.

14 DOE (August 2006) "Benefits of Using Mobile Transformers and Mobile Substations for Rapidly Restoring Electrical Service, A Report to The United States Congress Pursuant to Section 1816 of the Energy Policy Act Of 2005", DOE, https://www.energy.gov/sites/prod/files/oeprod/DocumentsandMedia/MTS_Report_to_Congress_FINAL_73106.pdf.

15 P. Dvorak (June 18, 2014) "Report update: Large Power Transformers and the U.S. Electric Grid" (Figure of step-up, step- down transformers), https://www.windpowerengineering.com/report-update-large-power-transformers-u-s-electric- grid/.

16 OSHA (n.d.) "Illustrated Glossary: Substations", https://www.osha.gov/SLTC/etools/electric_power/illustrated_glossary/substation.html#Step-up, Accessed May 15, 2020.

17 SGIP (November 7, 2016) "OpenFMB DER Circuit Segment Management Use Case", https://openfmb.ucaiug.org/Documents/Resources/SGIPOpenFMBUseCases-DERCircuitSegmentManagement- 20161107a.pdf, Accessed May 17, 2020.

18 GeneralPac (March 17, 2015) Introduction to the Delta Wye Transformer Connection Part 1: Delta Connection, https://www.youtube.com/watch?v=j5tDh8QKftg, Accessed May 15, 2020.

19 Lawrence Berkeley National Laboratory (June 2016) "Recovery of Utility Fixed Costs: Utility, Consumer, Environmental and Economist Perspectives," https://emp.lbl.gov/sites/all/files/lbnl-1005742_1.pdf.

20 S. Blumsack (2020) "What went wrong with rate of return regulation?", https://www.e-education.psu.edu/eme801/node/536, Accessed July 14, 2020.

21 S. Blumsack (2020) "Electricity Industry Structure and Regulation", https://www.e-education.psu.edu/eme801/node/529, Accessed July 14, 2020.

22 S. Blumsack (2020) "Deregulation or restructuring?", https://www.e-education.psu.edu/eme801/node/534, Accessed July 14, 2020.

23 Kansas Municipal Utilities (n.d.) "Benefits of Municipal Utilities", https://www.kmunet.org/page/Benefits/Benefits-of-Municipal-Utilities.htm.

24 Investopedia (March 26, 2020) "Cost of Capital vs. Discount Rate: What's the Difference?", https://www.investopedia.com/ask/answers/052715/what-difference-between-cost-capital-and-discount-rate.asp.

25 Valparaiso University (n.d.) "Engineering Economics", https://www.valpo.edu/student/asme/FE%20Slides/EngEconSlides.pdf.

26 M. Gange (November 4, 2011) "Low-frequency and Tonal Characteristics of Transformer Noise", https://www.acoustics.asn.au/conference_proceedings/AAS$_2$011/papers/p23.pdf, Accessed June 24, 2020.

27 Electrical Engineering Portal (October 3, 2016) "Problems with audible substation noise and what you can do about it", https://electrical-engineering-portal.com/audible-substation-noise, Accessed June 25, 2020.

28 NERC (2020) "M-1 Reserve Margin", https://www.nerc.com/pa/RAPA/ri/Pages/PlanningReserveMargin.aspx, Accessed June 4, 2020.

29 Electrical Classroom (2020) "PLC, DCS, RTU, SCADA and PAC Systems," https://www.electricalclassroom.com/plc-dcs- rtu-scada-and-pac-systems/, Accessed May 19, 2020.

30 L. Fraas and L. Partain (December 19, 2019) "Solar Cells and their Applications, Second Edition", Wiley, 2010, ISBN 978-0- 470-44633-1, Figure 10.2, https://commons.wikimedia.org/w/index.php?curid=17181966, Accessed May 18, 2020.

31 Xcel Energy (December 5, 2019) "Xcel large generator interconnection request," https://www.transmission.xcelenergy.com/staticfiles/microsites/Transmission/Files/PDF/Interconnection/LGIP-Interconnnection-Request.pdf, Accessed May 18, 2020.

32 DSIRE (August 5, 2014) "Interconnection Standards", https://programs.dsireusa.org/system/program/detail/2882, Accessed May 19, 2020.

33 B. Briscoe, A. Odlyzko, and B. Tilly (July 1, 2006) "Metcalfe's Law is Wrong", IEEE Spectrum, *https://spectrum.ieee.org/computing/networks/metcalfes-law-is-wrong*, Accessed May 15, 2020.

34 Xcel Energy Interconnections (n.d.) "Xcel Interconnection instructions," Interconnection instructions, reference Large Generator Interconnection Study Request in the Definitive Interconnection Study Process ("DISIS"), Accessed May 19, 2020.

35 Stevens Engineering, (March 1995) "Medium Voltage Transformers", *https://stevenengineering.com/tech_support/PDFs/45TMEDVOLT.pdf*, Accessed May 15, 2020.

36 Ecofys US (October 2016) "Interconnection Guidebook", Energy Trust of Oregon, *https://www.energytrust.org/wp-content/uploads/2016/10/100908_Interconnection_Guidebook.pdf*, Accessed May 14, 2020.

37 T&D World (July 23, 2019) "Operating, Spinning Reserves Need to Be Optimized for Reliability", *https://www.tdworld.com/grid-innovations/generation-and-renewables/article/20972878/operating-spinning-* reserves-need-to-be-optimized-for-reliability, Accessed May 19, 2020.

38 OSHA (n.d.) "Protection from Hazardous Differences in Electric Potential", *https://www.osha.gov/pls/oshaweb/owadisp.show_document?p_id=9871&p_table=standards*, Accessed July 29, 2020.

39 Duke Energy (n.d.) "Designing ways to operate the grid more simply and cost-effectively", *https://www.duke- energy.com/our-company/about-us/smart-grid/coalition*.

9.0 GUIDELINES FOR THE DESIGN OF A SYNGAS-TO-LIQUID FUELS AND CHEMICALS CONVERSION MODULE

9.1 INTRODUCTION

Gasification produces raw syngas (see Chapter 3), a flexible chemical building block that can be converted to power, liquid fuels, and a variety of chemicals (see Chapter 2). Before conversion, the raw syngas undergoes cleaning to produce a clean syngas stream (see Chapter 6). This chapter discusses the design and operation of the catalytic reactors for conversion of the cleaned syngas to fuels and chemicals. The discussion focuses on liquid fuels usually obtained from petroleum (i.e., gasoline, diesel, and jet fuel), methanol, substitute natural gas (SNG), and hydrogen (H_2).

The production of liquid fuels and chemicals requires deep cleaning of the raw syngas to completely remove common catalyst poisons, such as sulfur, ammonia, chlorine, and heavy metals (see Chapter 6). The reduction of sulfur compounds to less than 30 parts per billion (ppb) is generally a requirement to prevent catalyst poisoning for all carbon monoxide-hydrogen (CO-H_2) synthesis reactions, except for those that make use of sulfided catalysts, such as ethanol and higher alcohol synthesis and sour water-gas shift (WGS) reactions.

In addition to the deep cleaning of syngas, adjustment of the H_2-to-CO ratio in syngas via WGS followed by carbon dioxide (CO_2) removal is also required (depending on the catalyst and product) before feeding the syngas to the catalytic reactor. WGS can be carried out using appropriate catalysts for both sour and sweet syngas.* For syngas from gasification, it is preferable to carry out a sour WGS (before acid gas removal) to adjust the H_2-to-CO ratio or produce H_2 as the primary product. Sulfur and CO_2 separation is then more conveniently carried out downstream of the WGS, resulting in a more streamlined and integrated overall gasification-based plant.

This chapter focuses on catalyst selection, reactor design, and reactor operation for the syngas conversion reactor module. Topics related to separation, purification, and/or upgrading of the primary syngas reactor products (e.g., separation of methanol followed by upgrading to gasoline and olefins) are discussed in Chapter 10.

Following a review of the chemistry, catalysts, and other pertinent technical aspects of CO-H_2 synthesis reactions, this chapter describes the factors related to catalyst selection and design/operation of the reactor module for the desired primary syngas conversion product (i.e., SNG, methanol, transportation fuels, or H_2). All these products can be commercially produced from syngas. Topics include desired syngas compositions, heat management, reactor types, and preferred catalysts. Finally, this chapter discusses the status of research and development (R&D) and future trends in direct syngas conversion to products other than SNG, methanol, H_2, and transportation fuels; small, efficient modular reactor systems; heat management using structured catalysts; and product yield and selectivity improvements using bi-functional catalysts.

9.2 REVIEW OF COMMERCIAL SYNTHESIS REACTIONS

9.2.1 SUBSTITUTE NATURAL GAS AND METHANOL SYNTHESIS

The development of catalysts and reaction conditions for the conversion of syngas (CO+H_2) with very high yield (greater than 95%) to a single high value C_2+ product is a highly desirable commercial goal. This goal has proved elusive (except for conversion to C1 compounds), despite significant effort by the research community. The only two single-step/single reaction CO hydrogenation routes that have achieved this goal are methane (CH_4)[1] and methanol (CH_3OH)[2, 3, 4, 5, 6, 7, 8] production.

$$CO + 3H_2 \rightarrow CH_4 + H_2O \qquad \text{(R9.1 CO Methanation, } \Delta Ho \sim -206 \text{ kJ/mol)}$$

$$CO + 2H_2 \rightarrow CH_3OH \qquad \text{(R9.2 Methanol synthesis, } \Delta Ho \sim -91 \text{ kJ/mol)}$$

Due to the extremely high selectivity achievable using catalysts available from major catalyst manufacturers, such as BASF, Clariant, Johnson Matthey and Haldor Topsøe, both routes have been commercialized, although improvements continue in catalyst performance and reactor design. A nickel (Ni)-based catalyst is used for SNG synthesis (methanation) at around 350 to 375°C (662 to 707°F), whereas a copper-zinc (Cu-Zn)-based catalyst is generally used for methanol synthesis at around 220 to 250°C (428 to 482°F). Both processes use elevated pressures to increase yield.

---

* "Sour" and "sweet" refer to the concentration of acid gases (namely hydrogen sulfide [H_2S] and carbonyl sulfide [COS], but also carbon dioxide [CO_2]) in the syngas. Acids are "sour," so "sour syngas" is syngas that has not yet undergone acid gas removal (AGR). It is for this reason that, when CO_2 capture is present, additional water-gas shift that is performed before AGR is called "sour-shift," whereas it is called "sweet-shift" when it occurs after AGR.

9.2.2 FISCHER-TROPSCH SYNTHESIS

The Fischer-Tropsch (F-T) process is another synthesis route that has achieved commercial status, but not widespread use, despite being a highly non-selective set of reactions resulting in a wide slate of C_1 to C_{60+} hydrocarbons (i.e., alkanes, olefins, and alcohols). The primary reason for commercialization is that F-T synthesis (FTS) with an appropriate catalyst can produce strategically important liquid transportation fuels, particularly diesel, although not with high selectivity.

9.2.2.1 BACKGROUND

FTS was discovered in the early 1900s and commercialized in Germany before World War II and in South Africa in the 1950s at a much larger scale.[9] FTS R&D efforts continue today, aimed to elucidate the reaction mechanism, improve catalyst performance, better manage reactor exothermic reactions by incorporating more efficient heat removal in the reactor design, and increase selectivity to specific transportation fuels or chemicals.[†]

The traditional catalysts for FTS to produce transportation fuels at commercial scale in fixed-bed, fluidized-bed, or slurry reactors are either iron (Fe)- or cobalt (Co)-based.[10, 11, 12, 13, 14, 15, 16, 17, 18, 19, 20, 21, 22, 23, 24, 25, 26, 27] Though more expensive than Fe-based catalysts, recent commercial emphasis (e.g., Sasol Oryx and Shell Pearl FTS plants in Qatar, Chevron-Sasol plant in Nigeria) has been on Co-based catalysts,[28] because of their better integration with syngas from natural gas, higher stability, higher activity, low CO_2-selectivity, and higher attrition resistance in slurry reactors.

Commercial reactor systems for FTS[16, 18, 28, 29] are discussed in a later section. In recent years, the R&D focus has shifted from catalyst/process development for large-scale commercial reactors toward niche, potentially profitable applications of small-scale heat-exchange reactors, structured reactor packing to facilitate heat management, and bifunctional catalysts to modify selectivity.[30, 31, 32, 33, 34, 35, 36, 37, 38, 39, 40, 41, 42, 43] The shift is driven by the investment risk required for large-scale plants amid volatile oil and natural gas prices—even though small-scale systems cannot bring about a shift in the overall energy mix.

9.2.2.2 CHEMISTRY

It is generally agreed that the F-T reaction process occurs through a chain growth mechanism, but the agreed-upon mechanism remains under debate despite much research.[17, 44, 45, 46, 47, 48, 49] The species that leads to chain growth is not fully agreed upon by the research community, because no mechanism has fully explained the formation of all the products, particularly oxygenates and branched hydrocarbon compounds. However, it is recognized that it may be a single carbon monomer fragment (e.g., [$-CH_2-$]) formed following CO dissociation on the catalyst surface. The monomer fragment ([$-CH_2-$]) then further polymerizes and reacts to form alkanes, alkenes, and alcohols. In addition, the WGS reaction and the Boudouard reaction also occur during FTS. The WGS can be undesirable depending on the syngas composition and catalyst. The Boudouard reaction can result in catalyst deactivation and modification due to carbon or carbide formation. The overall chemistry of FTS can be represented by the following set of reactions.

Table 9-1. FTS reactions (n is an integer, 1 or larger)

Alkanes (paraffins)	$(2n+1)H_2 + nCO \rightarrow CnH_{2n+2} + nH_2O$
Alkenes (olefins)	$2nH_2 + nCO \rightarrow CnH_{2n} + nH_2O$
Methanol and Higher Alcohols	$2nH_2 + nCO \rightarrow CnH_{2n+1}OH + (n-1)H_2O$
Water-Gas Shift	$CO + H_2O \leftrightarrow CO_2 + H_2$; $\Delta H^o = -41.2$ kJ/mol
Boudouard Reaction	$2CO \rightarrow C + CO_2$; $\Delta H^o = -172.5$ kJ

Like the methanation and methanol synthesis processes previously discussed, the FTS process is also highly exothermic with a standard enthalpy of reaction (ΔH^o) of around -165 kJ/mol of reactant, depending on the product distribution. The WGS reaction is mildly exothermic and the Boudouard reaction is highly exothermic (Table 9-1); thus, efficient heat management and removal is common and essential for all CO hydrogenation processes to maintain product yield/selectivity and prevent catalyst sintering/deactivation.

The relative FTS yield of alkanes, alkenes, and alcohols strongly depends on the catalyst and reactor conditions employed. When $n = 1$ (in Table 9-1), the alkane forming reaction is the same as methanation and the alcohol forming reaction is the same as methanol synthesis. These processes are considered special cases of FTS that are carried out with very high selectivity to produce CH_4 or methanol (respectively) using catalysts that are different from those used for FTS to produce transportation fuels.

9.2.2.3 CATALYSTS

FTS to produce transportation fuels is traditionally carried out using Co- or Fe-based catalysts. The preparation of these catalysts is discussed by Dry[16, 22] and van der Loosdrecht et al.[28] Ruthenium (Ru), which produces mostly long chain hydrocarbons at relatively lower operating temperatures,[50] has also been evaluated for FTS. However, Ru has emerged primarily as a Co catalyst promoter and not a serious contender for commercial FTS due to its high cost. Commercial Co catalysts contain 7 to 20 wt% Co, are typically promoted with less than 1 wt% Ru, and supported on high-surface area silica, alumina, or titania. On the other hand, commercial Fe catalysts are unsupported bulk and typically promoted with potassium (K), Cu, and silica (SiO_2). A well-known, widely used Fe catalyst is the German Ruhrchemie catalyst with a composition by weight of 100Fe/5Cu/4.2K/25SiO_2.[51]

Commercial Co and Fe FTS catalysts produce a distribution of alkanes (paraffins) and alkenes (olefins) and exhibit lower selectivity to alcohols and oxygenates, such as aldehydes and ketones. The product distribution depends on operating temperature. Co catalysts typically operate near 210 to 240°C

[†] The literature on FTS is extensive. A comprehensive list of FTS literature from 1920 to 1980 containing more than 7,500 citations and references is available at *www.fischer-tropsch.org*.

(410 to 464°F), whereas Fe catalysts operate near 250 to 270°C (482 to 518°F) in a low-temperature F-T (LTFT) process or near 300 to 340°C (572 to 644°F) in a high-temperature F-T (HTFT) process. Fe catalysts produce more olefins than Co catalysts. Lower temperature operation of either catalyst produces more diesel range and longer chain hydrocarbons, whereas the Fe-based HTFT is geared to produce more gasoline range and shorter chain hydrocarbons. Commercial Co and Fe catalysts with as low as possible CH_4 selectivity are preferred. Fe-based LTFT catalysts exhibit lower CH_4 and light hydrocarbon selectivity than Co catalysts, but produce a significant amount of CO_2 due to their high WGS activity.

In contrast to FTS, SNG synthesis is carried out using a nearly 100% CH_4-selective Ni-based catalyst, whereas methanol synthesis is carried out using a nearly 100% methanol-selective bimetallic Cu-Zn catalyst at appropriate reaction conditions to maximize their respective yields.

9.2.2.4 PRODUCT DISTRIBUTION

In addition to the catalyst type and temperature, the product distribution for an FTS process depends on other factors, such as total pressure, reactant partial pressures, and the type of reactor. The repeating selectivity pattern, with an increase in carbon number via the addition of the monomer fragment, has led to the development of many mathematical models to predict/estimate the product distribution. The simplest and most widely used model equation for approximating the FTS product distribution is the Anderson-Schultz-Flory (ASF) equation.[47] This is an idealized single parameter equation relating the chain growth probability factor, α (has a value from 0 to 1), the mole fraction of the nth carbon number compounds (X_n), and the number of carbon atoms (n):

$$X_n = (1-\alpha)\,\alpha^{(n-1)} \qquad \textit{Equation 9-1}$$

Equation 9-1 can be linearized by taking logs to yield:

$$\log (X_n) = n\log \alpha + \log [(1-\alpha)/\alpha] \qquad \textit{Equation 9-2}$$

A useful version of Equation 9-2 for material balance/design purpose relates α to the weight fraction (Wn) rather than the mole fraction:[25, 52]

$$\log [W_n/n] = n\log \alpha + \log [(1-\alpha)^2/\alpha] \qquad \textit{Equation 9-3}$$

Based on Equation 9-3, a plot of log (W_n/n) on the y-axis and n on the x-axis should yield a straight line with a downward slope of log α. This model predicts that catalysts with high α have a higher weight fraction of heavier hydrocarbons (Figure 9-1). Also, only C_1 compounds (n=1, CH_4 or methanol) can be formed with a selectivity of 100% when α is close to 0, depending on the monomer fragment. As seen from Equation 9-1, if α is very low (e.g., 0.01), meaning little to no chain growth, the mol or weight fraction of the single carbon product (n = 1), CH_4 or methanol is close to 1, meaning 100% selectivity. It follows that the commercial catalysts/conditions used for SNG and methanol synthesis that provide 99+% selectivity do so because of very low α values approaching zero. On the flip side, if the goal is transportation fuels, a catalyst with high α (as low as possible C_1 selectivity) is preferred.

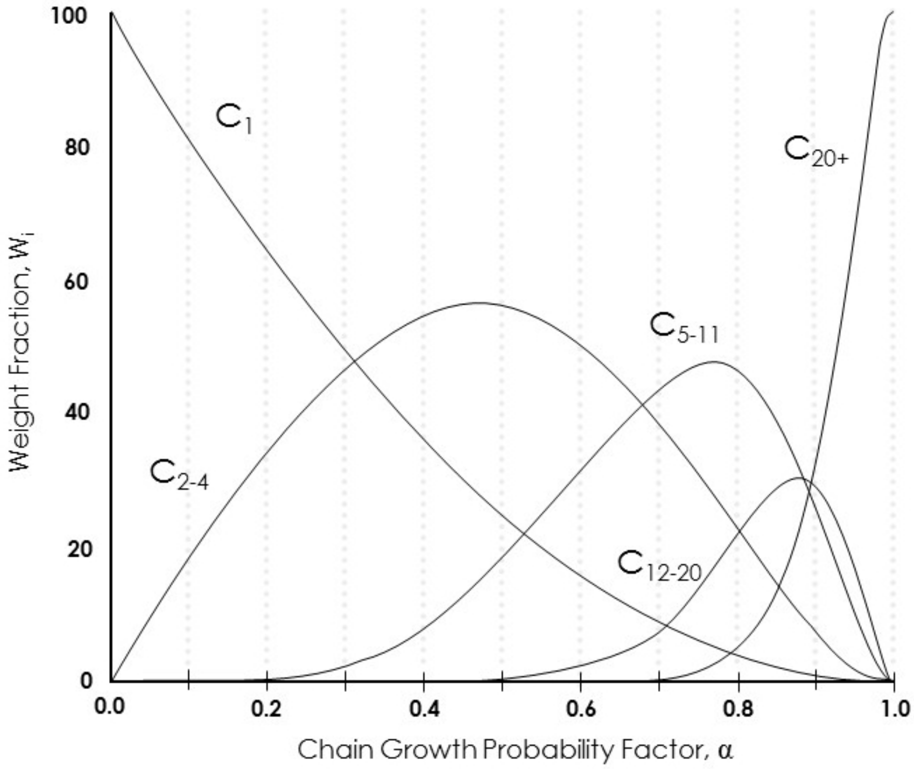


Figure 9-1. Hydrocarbon selectivity as a function of chain growth probability factor, α

Calculations based on Equation 9-3 to form Figure 9-1 shows that, if gasoline-level hydrocarbons (C_5-C_{11}) are to be maximized in FTS, catalysts and conditions resulting in an α of 0.75 to 0.8 should be used. On the other hand, to maximize diesel-level hydrocarbons (C_{10}-C_{20}), catalysts and conditions resulting in an α of 0.9 or higher should be used. Note that hydrocarbons and waxes heavier than C_{20} can be separated and subjected to mild hydrocracking to bring them back into the diesel pool, so high α catalysts/conditions approaching 0.91 to 0.95 are preferred if diesel is the desired product. Deviations from the ASF distribution are reported in literature as summarized by Lu.[29] These deviations include high CH_4, potentially due to a different catalytic site for methanation, heat/mass transfer limitations, or hot spots; low C_2 compounds, potentially due to re-adsorption and secondary reactions of ethylene; and the existence of two α (one that corresponds to compounds up to C_9 and another corresponding to C_{10}+ compounds), possibly due to the presence of catalytic sites that are different from chain growth sites, leading to termination reactions that govern C_9 compounds and below (e.g., hydrogenation, isomerization, and hydrogenolysis).

A case can also be made that, α increases gradually with carbon number rather than having two α values.[25] However, α calculated from the ASF plot is widely considered an important characteristic for the selection/purchase of a catalyst from a manufacturer depending on the desired range of liquid products. Knowing the α value (or the two α values), the project developer can use the ASF plot/equation to determine the expected product distribution, making it a useful 'a priori' reactor design and techno-economic evaluation tool for the plant under consideration.

9.2.3 HYDROGEN SYNTHESIS

Although there are 14 ways H_2 can be produced, the large-scale commercial production of H_2 for market distribution is mostly carried out by steam reforming of pre-cleaned natural gas to produce clean syngas followed by traditional WGS (high temperature and low temperature), CO_2 separation, and purification.[53, 54, 55] The advantage of steam reforming is that an air separation plant is not required, as the sensible and endothermic heat needed to drive the reaction is indirectly provided to the reactor by externally burning the required portion of natural gas. The reaction network is also simpler and easier to control compared to auto thermal reforming or partial oxidation. WGS is reversible and mildly exothermic, so lower temperatures thermodynamically favor higher CO conversion.

$$CO + H_2O\ (g) \leftrightarrow CO_2 + H_2 \qquad \text{(R9.3 WGS, } \Delta H^{0\sim}\text{ -41.2 kJ/mol)}$$

Hydrogen can also be made from syngas produced from coal, waste, biomass, or petroleum coke gasification. For better process integration, it is advantageous to subject syngas from gasification to sour WGS in the presence of hydrogen sulfide (H_2S) followed by sulfur cleanup, CO_2 removal, and subsequent purification steps. Another advantage of sour WGS is that carbonyl sulfide (COS) hydrolysis occurs during WGS, so a separate reactor for COS hydrolysis is not required. High-temperature, low-temperature, and sour-shift WGS catalysts differ in the active metals and temperature of operation (Table 9-2).[56, 57] These catalysts are commercially available from major catalyst suppliers such as BASF, Clariant, and Haldor Topsøe.

Table 9-2. WGS catalysts and corresponding operating temperature ranges

WGS REACTION TYPE	CATALYST	TEMPERATURE RANGE
High Temperature	Promoted iron-chromium	315 to 425°C (599 to 797°F)
Low Temperature	Copper-zinc	175 to 260°C (347 to 500°F)
Sour	Pre-sulfided cobalt-molybdenum	230 to 260°C (446 to 500°F)

9.3 SYNGAS COMPOSITION AND REACTOR MODULE DESIGN DISCUSSION

The synthesis of H_2, SNG, methanol, and F-T products from syngas are all commercially available processes offered by several process developers and/or catalyst vendors. This section is presented as a guide for project developers interested in installing and operating these processes. Using the information provided, project developers looking to invest in a gasification-based plant can establish a background in the reactors and catalysts for these processes. This background enables project developers to discuss their requirements with commercial process developers and catalyst vendors and down select to an appropriate commercial process/catalyst.

9.3.1 SYNGAS COMPOSITION

9.3.1.1 HYDROGEN PRODUCTION

For H_2 production from gasification using a sour WGS catalyst, the goal is to convert nearly all CO in syngas to H_2 and CO_2 using steam so that the steam-free relative concentrations of CO, CO_2, and H_2 in the syngas feed is unimportant. However, to maximize H_2 production it is important not to have high levels of carbon in the syngas tied to CH_4 and other light hydrocarbons to avoid a costly steam reformer. Syngas compositions from various gasifiers are reviewed in Chapter 3. If a moving-bed gasifier is used, a significant level of carbon in the fuel gets converted to hydrocarbons, making them undesirable for H_2 production. Gasifiers that produce low levels of CH_4 and other hydrocarbons in syngas, such as entrained flow gasifiers, are preferred for H_2 production. Excess steam can be added to the syngas to increase its concentration well above that needed stoichiometrically and thermodynamically to favor the forward WGS.

9.3.1.2 SUBSTITUTE NATURAL GAS AND METHANOL SYNTHESIS

For SNG, methanol, and FTS, the determination of the required feed syngas composition is not straightforward, because (depending on the process, operating conditions, and catalyst) CO_2 can also participate in the synthesis either directly or through reverse WGS. The CO_2 conversion reaction for SNG is also highly exothermic.

$$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2 \qquad \text{(R9.4 } CO_2 \text{ Methanation } \Delta H^o = -164 \text{ kJ/mol)}$$

Each mole of CO_2 requires 4 moles of H_2; it follows that the feed syngas H_2/CO ratio for SNG production needs to be in the range of 3 to 4 depending on the CO_2 concentration of the syngas. The preferred gasifiers for SNG are those that produce high H_2/CO ratios (like the Lurgi moving-bed used at a North Dakota SNG plant [USA]) and significant levels of CH_4 and C_2-C_4 hydrocarbons that are already suitable as components of SNG. However, one disadvantage is that extensive heavy hydrocarbon/tar/phenol removal is required for these gasifiers prior to sour WGS to adjust the H_2/CO ratio. The degree of WGS required (and/or the preferred H_2/CO ratio) before feeding the syngas to either the SNG synthesis reactor or the methanol synthesis reactor and the preferred CO_2 concentration in the syngas feed is usually specified by the commercial process and/or catalyst vendors. The shifted syngas may require CO_2 separation before synthesis.

Even small amounts of CO suppress the methanation of CO_2, because the Ni surface is preferably covered by CO rather than by CO_2.[1] However, CO_2 may participate in the reverse shift reaction, so the best option for SNG synthesis is to have the vendor specify the H_2/CO ratio and the CO_2 concentration in the feed syngas. As for the water levels in the syngas (it is a product of the reaction), it would be thermodynamically advantageous to remove it from the syngas. However, since the methanation reaction is highly exothermic, steam injection may help to moderate the reaction, as used in recent reactor designs.

The CO_2 conversion reaction for methanol synthesis ($CO_2 + 3H_2 \rightarrow CH_3OH + H_2O$, $\Delta H^o = -50$ kJ/mol) necessitates that the syngas feed H_2/CO ratio ranges from 2 to 3 (as opposed to from 3 to 4 for SNG synthesis) depending on the CO_2 concentration in the feed syngas. For methanol synthesis, a stoichiometric ratio, defined as $(H_2\text{-}CO_2)/(CO+CO_2)$ of around 2 is desirable.[25] Also, in contrast to a Lurgi gasifier for SNG synthesis, an entrained flow gasifier that produces a syngas with very low levels of both CH_4 and light hydrocarbons is preferred for methanol synthesis to avoid a costly steam-CH_4 reformer.

For methanol synthesis, a small amount of CO_2 (around 4 to 8 vol% dry basis) in the syngas is considered beneficial, because mechanistic studies suggest that CO_2 helps to initiate the methanol synthesis.[25] Carbon dioxide also appears to act as a gaseous co-catalyst for the synthesis by increasing the rate of reaction. The water should mostly be removed from the syngas feed, considering that water can be formed by the reverse shift reaction, as well as the CO_2 hydrogenation reaction to produce methanol.

9.3.1.3 FISCHER-TROPSCH SYNTHESIS

FTS is somewhat different and more complicated, because the catalyst choice (i.e., Co or Fe) also dictates the preferred syngas composition.[16, 19, 22, 49, 28] Cobalt possesses very low WGS activity, whereas Fe is an active WGS catalyst. Since the cost of syngas is high, it is important that the maximum amount of CO is converted to hydrocarbons. This requires the gasification block to deliver a syngas feed composition to the F-T reactor that matches the overall usage ratio (defined as the net rate of H_2 conversion divided by CO conversion [28]).

At one extreme, little WGS takes place for Co catalysts; the usage ratio is primarily determined by the F-T reactions and methanation and olefin saturation to some degree. The required feed H_2/CO ratio is higher than 2 (near 2.1) to allow for CH_4 formation that consumes 3 moles of H_2 per mole of CO and for olefin saturation to the corresponding paraffin, in addition to the F-T reaction.

At the other extreme, an Fe catalyst with very high WGS activity that shifts all the water formed by FTS to H_2 and CO_2 can require a feed H_2/CO ratio as low as 0.5.[28] In practice, the feed H_2/CO ratio depends not only on the Fe catalyst's shift activity, but also on the feed CO_2 concentration and operating temperature (LTFT versus HTFT). According to Dry,[8] for the LTFT process, the H_2/CO usage ratio is typically approximately 1.7. For HTFT, the WGS is rapid and presumably reaches equilibrium. This could allow CO_2 to be converted into F-T products via the reverse WGS reaction followed by the F-T reaction. If the syngas has a ratio of $(H_2\text{-}CO_2)/(CO+CO_2)$ equal to around 2, all the H_2, CO, and CO_2 can (in principle) be converted to F-T products.[8] This is the same as the requirement for methanol synthesis.[25]

Co catalysts are the preferred catalysts for natural gas-to-liquids (GTL) due to their very low WGS activity. The H_2/CO ratio of syngas derived from natural gas is sufficient (even excessive), requiring the separation of some H_2 using a membrane. Conversely, since Fe catalysts are active WGS catalysts, the water generated by F-T can react with CO to provide H_2, so a lower H_2/CO syngas is preferred as obtained from entrained flow gasifiers using fuel species with lower H/C ratios. An active Fe-based catalyst for F-T with high WGS activity may enable the elimination of an upstream WGS reactor, because WGS simultaneously occurs with F-T. This does not necessarily imply that Fe catalysts are the catalysts of choice for syngas from coal gasification, as other factors (e.g., stability, attrition resistance, activity, and commercial availability) must also be considered to make a choice for the most suitable F-T catalyst.

At present, the commercial space is moving toward the use of Co catalysts for GTL, so Fe catalysts are not readily offered in commercial quantities by vendors. However, it should be possible to obtain Fe catalysts in test quantities, particularly the well-established Ruhrchemie composition, from vendors. It may be advantageous to use an Fe-based catalyst over a Co-based catalyst for liquids production for small, modular gasification-based plants

9.3.2 REACTOR DESIGN

Due to the exothermic nature of the synthesis reactions, the reactor type and its operating conditions need to be designed so that the heat generated can be managed to maximize CO conversion and product yield without causing catalyst degradation from sintering while also avoiding unsafe temperature excursions or runaways. The reactor choices are typically fixed-bed, fluidized-bed, or slurry. All the commercial reactor installations for H_2 via WGS, methanol synthesis, and SNG synthesis are fixed-bed reactors (FBRs). Commercial FTS reactors include all three reactor types.

9.3.2.1 HYDROGEN, SUBSTITUTE NATURAL GAS, AND METHANOL SYNTHESIS

Catalysts for the WGS, SNG synthesis, and methanol synthesis generally have an operating life of more than four years, after which the spent catalyst is typically taken away by the catalyst vendor, who then helps to repack the reactors with fresh catalyst. The catalysts for these processes are commonly provided with guaranteed performance by the vendors for the life of the catalyst, as long as they are activated and operated within the window specified by the vendor. All these catalysts require a field activation procedure as specified by the vendor (typically H_2 reduction at specified conditions) that needs to be strictly followed to bring the catalyst to its fully activated state without deactivating the catalyst due to overheating during reduction. Sour WGS catalysts are normally supplied as pre-sulfided by the vendor so they do not require field activation by H_2 reduction; they do require inert blanketing prior to beginning heat up and synthesis.

9.3.2.1.1 HYDROGEN PRODUCTION

Commercial WGS installations for H_2 production from syngas employ a series of fixed-bed adiabatic reactors that use the same catalyst with inter-stage cooling. The catalysts and their temperature range are provided in Table 9-2. Depending on the plant's integration requirements, a high-temperature WGS reactor using a Fe-Cr catalyst followed by a low temperature WGS reactor using a Cu-Zn catalyst is sometimes used to maximize CO conversion. The process is pressure-neutral and conveniently carried out at the pressure of the available syngas. The space velocity (a measure of syngas volumetric throughput per unit volume of catalyst) to be used is normally provided by the catalyst manufacturer and typically has a value of 2,000 to 5,000 standard cubic feet (scf) gas per hour per cubic foot catalyst.

The approach of using adiabatic reactors in series with inter-stage cooling is the simplest and least costly approach compared to gas recycle and/or heat exchange reactor (shell and tube) approaches. The heat generated is manageable, because WGS is only mildly exothermic. The reaction is equilibrium limited, but inter-stage cooling and steam addition (if necessary) reduces the gas temperature to allow for additional shift. In three stages, the CO can usually be reduced to less than 0.2% when using a low-temperature WGS catalyst. Nearly complete removal of CO is carried out by converting the remaining CO to CH_4 via methanation.

Sour WGS is more suitable for integration with a gasification plant. Several catalyst vendors, including Clariant, BASF, and Haldor Topsøe, provide sour WGS catalysts along with their design operating conditions. In addition to H_2 production, WGS can be used for adjusting the H_2/CO ratio to make it suitable for subsequent synthesis. Haldor Topsøe suggests two approaches for a sour WGS operation depending on the H_2/CO ratio desired in the shifted gas. The first approach uses a sour high-temperature WGS in the first stage, followed by a sour low-temperature WGS converter in the second stage. The high-temperature conditions provide a higher rate of reaction, so higher space velocities (i.e., smaller reactor) can be used, while the low-temperature conditions allow additional conversion due to a more favorable equilibrium state. The second approach uses a single sour WGS reactor for applications that require a lower H_2/CO product ratio, such as an F-T reactor using an Fe catalyst.

9.3.2.1.2 SUBSTITUTE NATURAL GAS SYNTHESIS

SNG synthesis from syngas is a commercial process, first commercialized at the Dakota Gasification Co. Great Plains Synfuels Plant (North Dakota, USA)—based on the gasification of lignite—and more recently at several locations in China.[1] Although many reactor concepts have been studied at research scale and fluidized-bed reactors have been erected at demonstration scale, only FBRs have been commercialized in spite of the very high heat of the methanation reaction.[1] There is limited or no incentive to produce attrition-resistant, commercial Ni methanation catalysts that are needed for fluidized-bed reactors due to the well-established, fixed-bed catalyst formulation and reactor technology. A list of commercial suppliers of SNG fixed-bed catalytic synthesis processes, along with their heat management concept, is summarized in Table 9-3. All these processes use Ni-based catalysts, although the exact chemical composition is not published by these suppliers. The typical reaction conditions are 350°C (662°F) and elevated pressure that could be near the gasifier pressure or specified by the vendor.

Table 9-3. SNG synthesis processes on the market

SUPPLIER	PROCESS NAME	REACTOR STAGES AND HEAT MANAGEMENT METHOD
Air Liquide (Lurgi)	Lurgi Methanation	2 adiabatic FBRs with gas recycling and intermediate cooling
Haldor Topsøe	TREMP	3-4 adiabatic FBRs with gas recycling and intermediate cooling
Clariant and Foster Wheeler	Vesta	3 FBRs with steam addition; no gas recycle
Johnson Matthey (Davy Technologies)	HICOM	3 adiabatic FBRs with gas recycling and intermediate cooling
Linde	Linde Isothermal Reactor	1 FBR with an internal heat exchanger

Further details of these processes, including required feed composition, pressure, temperature, and space velocity, can be obtained by a project developer under a non-disclosure agreement (NDA) from these suppliers. Low conversion per pass with recycle is potentially used to control the high heat of the reaction in three of the offered processes (Air Liquide, Haldor Topsøe, and Johnson Matthey). In the Vesta process by Clariant and Foster Wheeler, steam addition is used to control the forward reaction and exothermicity without the use of recycle. All these processes use two to four adiabatic reactors in series. Linde offers a process design based on a single, potentially isothermal, FBR with an internal heat exchanger. It likely involves a shell and tube type reactor with catalyst in the tubes and steam generation/cooling on the shell side. This design has not been commercialized for SNG synthesis,1 but has been commercially applied for methanol synthesis.

9.3.2.1.3 METHANOL SYNTHESIS

Global methanol production was around 85 million metric tons in 2016, with worldwide full production capacity at nearly 130 million metric tons.[58] Methanol mostly undergoes further chemical conversion before being turned into a final product (see Chapter 10). Methanol is a fuel (or fuel additive) and a building block for a wide variety of chemicals, such as formaldehyde, acetic acid, methyl tert-butyl ether (MTBE), dimethyl ether (DME), dimethyl amine, and olefins.[25]

The reactor types for methanol synthesis at relevant scales (greater than 100 ton/d) include fixed-bed and slurry, although only FBRs are used for commercial applications since the first commercial plant was built in 1923.[2, 3, 5, 7, 8] A noteworthy example of a slurry reactor application is the liquid-phase methanol (MeOH) synthesis process known as the LPMEOH™ process from Air Products and Chemicals.[7, 59] The LPMEOH™ process development was supported by the U.S. Department of Energy (DOE) at pilot-scale and demonstration scale. A 260 ton/day demonstration plant was built and operated at Eastman Chemical Company's chemicals-from-coal complex in Kingsport, Tennessee (USA). The process uses an inert mineral oil/powdered catalyst slurry as a reaction medium and heat sink. The slurry bubble column reactor (SBCR) approach allows close to isothermal operation by heat removal using boiler tubes inserted into the slurry. It can process syngas of varying composition, containing greater than 50 vol% CO without the need for an upstream WGS unit. Also, the fine slurried catalyst provides a simple way to remove and add the catalyst using pumps. Despite these attributes, LPMEOH™ was not commercialized. The following discussion pertains only to FBRs.

The first methanol fixed-bed synthesis technology, commercialized by BASF in 1923, was based on a zinc chromite ($ZnO\text{-}Cr_2O_3$) catalyst that could hydrogenate CO to methanol at 240 to 300 bar and 350 to 400°C (662 to 752°F). This catalyst also produces CH_4 and other light hydrocarbons with 2 to 5 wt% selectivity. With advances in syngas cleaning technologies, a major improvement was achieved in the 1960s: a sulfur-free syngas that enabled Imperial Chemical Industries (ICI) to develop a methanol synthesis process based on the use of a more active Cu/ZnO catalyst that is easily poisoned by even trace amounts of sulfur. The ICI process can operate at much milder conditions (i.e., 50 to 80 bar and 220 to 275°C [428 to 527°F]), but requires a pristine clean syngas and has 99+% methanol selectivity. As a result, most of the emphasis shifted toward improving the ICI catalyst/process, leading to the closure of the last zinc chromite-based, high-temperature plant in the mid-1980s.

All current methanol synthesis plants use the Cu/ZnO catalysts and its variants. Plants as large as 5,000 tons per day (tpd) or greater are currently in operation, and plans exist to increase to 10,000 tpd to achieve further cost advantages of economy of scale. The most notable improvements to the Cu/Zn catalyst to increase activity and catalyst life include some addition of alumina and/or magnesium, potentially as surface area stabilizers and/or spacers, to prevent agglomeration of copper crystallites due to the exothermic reaction. In modern catalysts, typical relative atomic content of Cu:Zn:Al is around 6:2:1.[25]

Modern catalysts are expected to deliver a productivity of greater than 1 kg methanol/liter catalyst/hour and a useful life of more than four years. With proper activation, deep upstream removal of impurities and poisons, and operational care during synthesis so as not to overheat the catalyst, they may have a useful life of six to eight years.[25, 60] Despite the selectivity of modern catalysts exceeding 99.5%, there remains a need to remove dissolved gases, water, and various impurities from the condensed product, such as DME, ketones, aldehydes, ethanol, and fusel alcohols, to produce the required salable grade of methanol (see Chapter 10).

Major licensors of present-day, large-scale methanol plants and/or producers of methanol include Air Liquide (Lurgi); Johnson Matthey (Synetix); Mitsubishi Gas Chemical; Methanex; and Kellogg, Brown, and Root (KBR). Air Liquide also offers medium-scale plants (less than 1 million tons/year), whereas Maverick Synfuels offers small-scale plants. Major catalyst suppliers include Johnson Matthey, Haldor Topsøe, BASF, and Clariant. Project developers interested in gasification-based methanol synthesis should contact these process and catalyst suppliers to down select to a suitable vendor for their scale of process. Further details of these processes, including required feed composition, pressure, temperature, and space velocity and expected methanol productivity, can potentially be obtained under an NDA from these licensors and/or producers. Details include:

- The distinguishing feature of most methanol plants (or licensors) is the type of FBR system used to overcome equilibrium limitations while controlling and dissipating (or removing) the heat of reaction. Numerous adiabatic and isothermal reactor designs have been commercialized over the years, and many designs use a high enough recycle ratio to limit the CO conversion per pass (to limit temperature excursions) while achieving 99+% overall CO conversion.[5, 8, 25, 60] Some examples include series adiabatic reactors with recycle in the first reactor and with heat exchangers between reactors.
- Quench reactors that mix cold unreacted gas with hot gas from one reactor and distribute it across the next reactor.
- Shell and tube reactor with syngas recycle and with catalyst either in the tubes or on the shell side.

Modern reactors are mostly of the latter two types. They use either circulating gas or raise steam to optimize the temperature distribution in the catalyst bed. The goal is to maximize the reaction rate while maintaining a balance between slower kinetics at lower temperatures and lower conversion at high temperature due to thermodynamic equilibrium limitations.

9.3.2.2 FISCHER-TROPSCH SYNTHESIS

Unlike methanol synthesis, FTS for either GTL or coal-to-liquids (CTL) has not advanced commercially, even though both processes have been commercialized, likely due to the abundance of inexpensive crude oil in the Middle East. According to the U.S. Energy Information Agency (EIA), GTL production was 225,000 barrels per day (bpd), or approximately 0.23 percent, of the world oil consumption of 100 million bpd in 2019.[61, 62] If oil remains cheap, EIA predicts that large-scale GTL production will top out at 350,000 bpd. Even if oil prices increase, GTL production is predicted to rise to no more than 1.1 million bpd by 2040. EIA expects new plants using FTS to be small (less than 5,000 bpd) and modular for niche applications, such as stranded gas and landfill gas monetization, because large-scale GTL projects carry risk and require large investment. With the transition to carbon-free energy, GTL plants may play a more prominent role in the future provided carbon-neutral fuels like biomass are used in the creation of the products.

CTL, which was commercialized in South Africa in the 1950s, may progress in Asia according to a report by Acumen Research Consulting[63] due to high growth in China and some growth in India and Indonesia. This report predicts a 4.6% annual growth, with the CTL market size reaching $5.8 billion by 2026, which conflicts with U.S./global interests in reducing fossil fuel use to combat climate change. Chinese companies have government backing due to China's large coal reserves and the incentive to develop a coal-based transportation fuel industry to increase energy independence by reducing the volume of crude oil imports.[64]

The major companies involved in GTL and CTL projects include Shell, Sasol, Chevron, Qatar Petroleum, PetroSA, Shenhua Group, Synfuels China, Yitai Group, and Yanzhou Coal Mining.[65] There were 37 operating facilities in 2017, including 10 intermittently operating pilot-scale facilities and 27 large-scale facilities, including 15 with capacities greater than 5,000 bpd. Details of the reactor design/operation and catalyst formulations used in these facilities remain proprietary by the process developers and catalyst manufacturers, because FTS applications are not widespread.

The reactors used in the past at Sasol and/or currently in use for FTS in Qatar, Nigeria, and China are shown in Figure 9-2 along with information about the process, catalyst, and plant size.[28] All these reactors have been commercialized except for the microchannel reactor under development at pilot-scale by Velocys in the United States.

The most successful commercial FTS reactors can be categorized as fixed-bed, fluidized-bed, and slurry bubble column. These reactors operate at 20 to 30 atm pressure and elevated temperatures, as shown in Figure 9-2. A circulating fluidized-bed (CFB) reactor was operated by Sasol for HTFT from 1980 to 1995, but was replaced with the more advanced bubbling fluidized-bed (BFB) reactor. The main advantages of BFB over CFB reactors included simplicity, low construction cost, ease of operation, and lower overall catalyst consumption due to lower rates of attrition, deactivation, and removal.

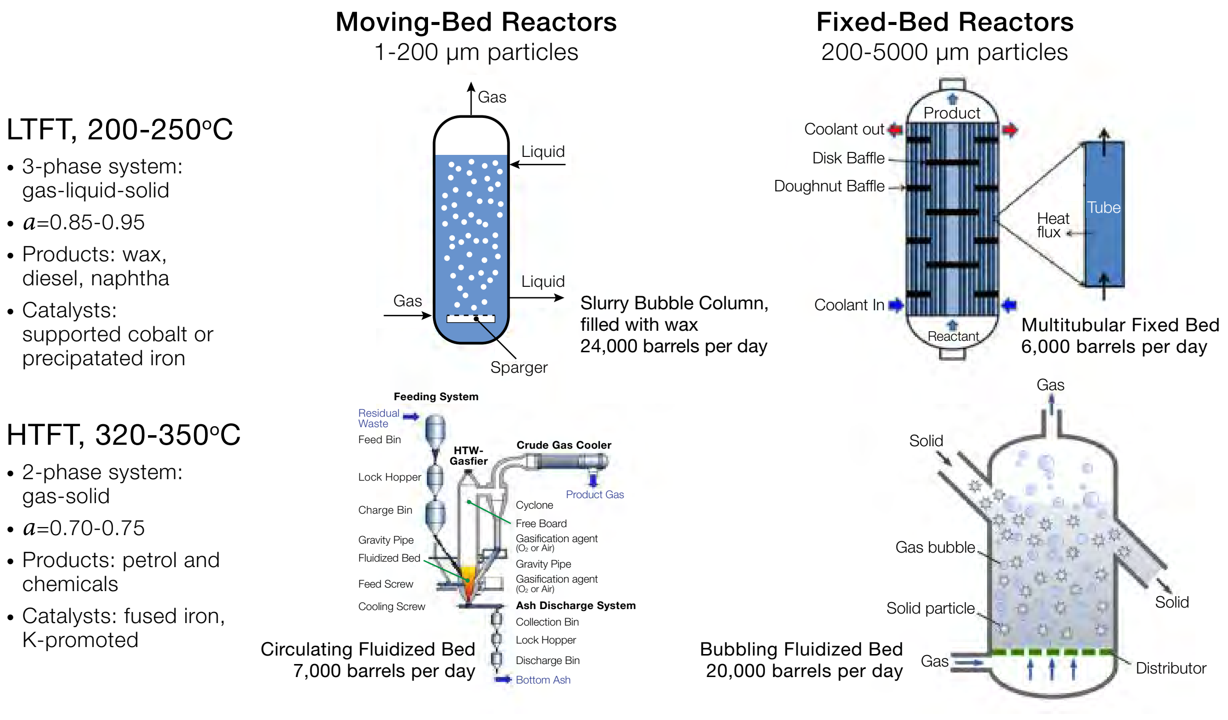


Figure 9-2. Fischer-Tropsch reactors[28]

9.3.2.2.1 FIXED-BED REACTOR

The Arge FBRs that later became the design basis of similar methanol synthesis reactors were designed by Lurgi and commercialized for LTFT in 1955 by Sasol for CTL. They are shell and tube type with a 10-foot diameter, 40-foot tall shell housing 2,050, two-inch diameter tubes containing high α-precipitated, Fe catalyst pellets.[29, 28] The syngas feed enters the top of the reactor tubes to take advantage of gravity by allowing the liquids produced to trickle down. Heat is efficiently removed by circulating high-pressure steam/water at close to water's boiling temperature on the shell side. The temperature is further controlled by maintaining low conversions per pass and recycling the unconverted gas after removing the condensable liquids and wax. Shell and tube reactors of higher capacity with a Co/TiO_2 catalyst are used by Shell at their Pearl Plant in Qatar, which is the world's largest GTL plant, producing around 140,000 bpd of F-T liquids and waxes.

Several FBRs are used in parallel, depending on the plant capacity, with at least 1 or 2 spare reactors to allow for catalyst replacement when the activity and/or selectivity of an operating reactor become unacceptable. The operating temperatures for LTFT are in the 200 to 250°C (392 to 482°F) range, with precipitated Fe catalyst (Arge reactor) run at the higher end of this range and Co/TiO_2 catalyst (Pearl) run at the lower end of this range. The desirable products from LTFT are jet fuel- and diesel-range hydrocarbons (C_8-C_{20}). The products are mostly alkanes and alkenes; they have high cetane ratings and are excellent for blending with diesel after saturating the alkenes via mild hydrogenation. The higher molecular weight waxes require further upgrading to produce lubricating oil and/or mild hydrocracking to bring them back into the diesel pool (see Chapter 10).

9.3.2.2.2 FLUIDIZED-BED REACTOR

The BFB reactor is used for HTFT at the Secunda CTL Plant in South Africa, also known as the "Synthol Plant." Due to the higher operating temperature and a lower-α fused iron catalyst, only a small amount of waxes (if at all) are produced. Temperature control is simpler in the fluidized bed—heat is efficiently removed using boiler tubes inserted into the reactor. As a result, fluidized-bed reactors also have nearly isothermal operation and higher throughput relative to the FBR. However, catalyst losses are higher than FBRs—losses occur not only due to chemical deactivation, but also due to mechanical attrition. However, replacement of the spent or elutriated catalyst with freshly activated catalyst can be carried out without shutting down the reactor.

The desired product for HTFT is gasoline-range hydrocarbons and aromatics (C_5-C_{11}), as opposed to jet fuel- and diesel-range hydrocarbons for LTFT. A poor gasoline with low octane number is produced from HTFT, which was acceptable in the past, because tetraethyl lead was blended in to boost the octane values, but this is no longer an option due to health and safety risks. Significant amounts of light gases, olefins, and oxygenates are also produced in HTFT. To produce an acceptable quality gasoline compatible with current requirements, costly product upgrading, consisting of oligomerization, isomerization, hydrotreating, and other steps, is required, rendering HTFT as complex as a multi-product crude oil refinery.[15, 16] This may explain the absence of new plants using HTFT beyond those in South Africa and it appears unlikely that HTFT plants will be built again. Consequently, HTFT will not be further discussed.

9.3.2.2.3 SLURRY BUBBLE COLUMN REACTOR

Like the FBR, the SBCR is also used for LTFT. Large operating commercial plants using SBCR include the Oryx GTL project (Qatar) and Escravos GTL project (Nigeria), each with a nameplate capacity of 34,000 bpd. These plants, jointly developed by Sasol and Chevron in partnership with local Government-owned entities, use the Sasol SBCR with a Co/Al_2O_3 catalyst. There are no commercial-scale, SBCR-based, fully operating CTL plants, but there is ongoing activity in China to build and commission a 100,000-bpd facility in Ningxia Hui.[66] Synfuel China's novel Fe-based SBCR that operates at 275°C/527°F (nearly 30°C/86°F higher than conventional fixed-bed LTFT) and its development in three demonstration-scale plants with 17 to 19 feet diameter SBCRs appear to be the driver for this commercial-scale plant.[67, 68, 69] The Fe catalyst is claimed to be attrition resistant. The commercial plant in Ningxia Hui has 24 dry-feed gasifiers and 10 Linde air separation units (ASUs). It uses Synfuels China F-T technology, consisting of two trains and eight SBCRs. The plant was scheduled to begin operation in 2017 to 2018, but no reports could be found to verify that the plant is fully operational (at the time of publication).

SBCRs are three-phase systems, involving the bubbling of syngas through a liquid suspension of finely divided catalyst (approximately 50 to 70 µm) that is typically prepared by spray drying a catalyst precursor slurry or salt solution. The liquid in the reactor is the F-T wax product that has low vapor pressure at the operating temperature and pressure. The slurry contains about 30% catalyst and acts as a heat sink. Excess heat is efficiently removed using boiler tubes within the slurry to maintain isothermal conditions. The liquid produced is withdrawn though a filter to maintain the liquid height. Several model-based designs of SBCRs have been reported in the literature, most notably by Krishna and coworkers at IFP, France.[23, 70, 71] A typical commercial design range reported by this group is 6 to 10 meters diameter; 30 to 40 meters in height; 20 to 40 bar; 230 to 250°C (446 to 482°F); 0.1 to 0.4 m/s superficial velocity; slurry-to-gas volume ratio in reactor 0.3 to 0.4; and 5,000 to 8,000, two-inch diameter vertical cooling tubes.

SBCRs have several advantages over shell and tube FBRs, including more efficient heat removal and nearly isothermal operation, simpler design and lower cost, higher conversion due to the ability to operate at higher temperatures, higher kinetics/productivity and lower intra-particle diffusional limitations from the use of small particles as opposed to pellets, and simple catalyst replacement ability to maintain activity without shutting down the reactor. However, one limitation of SBCRs versus FBRs is that they require mechanically and chemically robust catalyst particles to prevent their powdering due to attrition in the bubbling slurry. Powders can plug filters used for liquid withdrawal and exit the reactor with the liquid product, requiring replacement and complex catalyst-wax separation. A relatively soft particle can cause insurmountable problems in an SBCR.

Co catalysts for SBCRs are typically made by supporting 7 to 20 wt% Co on hard alumina micro-spheroidal particles that provide the necessary attrition resistance.[72] Fe catalysts are bulk iron oxides containing more than 70 wt% hematite (Fe_2O_3) that makes them relatively much softer compared to Co catalysts. Furthermore, Fe catalysts undergo various chemical changes during FTS that can weaken the particle, such as phase change due to reduction from hematite to magnetite, wüstite, or even elemental Fe. Slow Fe carbide formation from magnetite and/or carburization due to the Boudouard reaction typically occurs during and following an induction period to form the final active catalyst.[73, 74] The preparation of a suitable attrition-resistant Fe catalyst for SBCR is challenging and a bottleneck for the application of Fe catalysts in an SBCR. Synfuels China has claimed production of an attrition-resistant, Fe-based catalyst for SBCRs[66, 67, 68, 69] for their commercial plant at Ningxia Hui, but no published reports could be found that provide the run data from this plant that shows reduced catalyst losses due to attrition and their efforts related to catalyst wax separation.

9.3.2.2.4 CATALYST CHOICE

A comparison of the primary FTS catalyst candidates, Fe and Co, is summarized in Table 9-4 to assist project developers. The table applies to an LTFT plant using high-activity, high-α catalysts that maximize the production of jet fuel- and diesel-range hydrocarbons, because the development of a cost-effective HTFT plant is not expected in the future.

Table 9-4. Comparison of high-α Fe and Co catalysts for Fischer-Tropsch synthesis

	HIGH-A, FE-BASED CATALYST	HIGH-A, CO-BASED CATALYST
Typical Formulation	100Fe/5Cu/4.2K/25SiO_2	15-20 wt% Co on Al_2O_3
Promoter (s)	Cu, K	Ru~0.1 wt%
Preferred Syngas H_2/C0.71	2.05 to 2.10	
Acceptable Sulfur in Syngas	<1 ppm	<30 ppb
α	0.91 to 0.93	0.90 to 0.91
Typical Activation Procedure	Syngas, slow ramp to 280°C (536°F)	H_2, slow ramp to 350°C (662°F)
Temperature Range	240 to 275°C (464 to 527°F)	210 to 240°C (410 to 464°F)
(464 to 527°F)	~1,000 to 2,000 scf/(cf cat.h)	~1,000 to 2,000 scf/(sf cat.h)
(410 to 464°F)	~0.3 to 1 g/(g cat.h)	~0.3 to 0.7 g/(g cat.h)
Space Velocity (fixed-bed)	<5%	~8 to 11 %
C_3+ Productivity	High	Very low
Methane Selectivity	High	Very low
WGS Activity	Medium	Low
CO_2 Selectivity	High	High
Olefin/Oxygenates Selectivity	Medium to high	Very low
Wax Selectivity	High	Low
Chemical Deactivation Rate	Low	High
Attrition Rate in SBCR	High	Low
Commercial Production Cost	Low	High

The most important difference between Fe and Co catalysts is their preferred feed composition. Co catalysts require a high H_2/CO ratio so an upstream WGS is required, whereas it may not be required for an Fe catalyst where WGS occurs in the FTS reactor. The same amount of carbon is rejected as CO_2 for both catalysts, but the capital cost is lower for an Fe catalyst-based plant. Other advantages of an Fe catalyst include its higher tolerance to sulfur, lower CH_4 selectivity, and lower cost. Disadvantages include its higher selectivity to olefins and oxygenate, higher chemical deactivation rate, and higher mechanical attrition rate (if an SBCR is used).

SBCRs are more cost-effective for large-scale plants, whereas FBRs are more convenient for smaller scale plants. A reasonable choice for gasification-based syngas conversion via FTS would be parallel Fe catalyst-based FBRs for smaller plants and large Co catalyst-based SBCRs for larger plants. The reason for not choosing an Fe catalyst for SBCRs is that its poor mechanical attrition resistance can cause major operational problems (unless the Synfuels China Fe catalyst data shows high-attrition resistance). For a GTL plant of any size, a Co catalyst is a better choice, because of its higher stability and more favorable H_2/CO ratio achievable from natural gas.

9.4 REACTOR DESIGN, MASS BALANCE, AND PROCESS CONTROL

If the project developers are prepared to move forward with a gasification plant (based on a value proposition assessment [see Chapter 3]) and they have chosen H_2, SNG, methanol, or liquids as their primary product, the next task is proceeding with a reactor design and mass balance around the reactor. A sulfur guard bed upstream of the reactor should be an integral part of any synthesis plant to prevent accidental sulfur increase in the syngas feed from damaging the catalyst.

Vendors, catalyst details, space velocity data and productivity data are included in this chapter to enable preliminary reactor/vendor selection, reactor sizing, and mass balance. In most cases, an FBR system should be chosen.

The project developer should then carry out a detailed discussion with process suppliers and catalyst vendors to enable down selection. The reactor design can then move forward with the help of the chosen process and catalyst suppliers and an engineering company. This includes a mass/energy balance, piping and instrumentation diagram (P&ID), detailed reactor sizing and design, and process hazard analysis (PHA). Based on the PHA, any recommended safety modifications should be made to the P&ID.

The catalyst activation methods and required analytical and process control equipment should be specified by the process supplier. There should be enough analytical equipment and instrumental interlock control to carry out the catalyst activation according to the specified procedure; cool the reactor to reaction initiation temperature; bring in the syngas feed; gradually adjust the pressure and temperature to reach steady state while preventing the overheating of the catalyst; measure and control the reactor steady state performance, such as upstream and gas composition analyzers (both gas compositions and continuous analyzers); and detect and correct unsafe situations, such as a temperature excursion due to the exothermic reaction and leaks.

9.5 NEW TRENDS

Large-scale methanol synthesis plants have proven cost effective, and plants larger than those that exist today (e.g., 5,000 tpd) are being considered. Recent large-scale GTL FTS plants that have come online, such as Pearl GTL (Qatar) and Escravos GTL (Nigeria), are projected to be profitable due to site-specific benefits, but have required large capital investments ($130,000 to $265,000/bpd). This capital investment has rendered large GTL plants financially risky and future plans for some GTL plants, such as those by Shell and Sasol in Louisiana (USA), have been shelved. Some FTS companies, such as Syntroleum and Rentech, have exited the FTS business. Interest in commercializing FTS remains, as new trends emerge toward small-scale GTL by several companies, including Velocys, Compact GTL, and Infra.

Velocys's microchannel reactor-based FTS process is discussed in the following section. Compact GTL offers a proprietary, two-stage F-T system for GTL. The company was developing a 2,500-bpd GTL project in Kazakhstan to come online in 2018, but it does not appear to have started operations.[75] Infra offers a modular, small-scale GTL plant that uses a low-α, Co-based FTS catalyst that targets the production of gases and mostly naphtha range hydrocarbons.[76] Emphasis has shifted toward niche, potentially profitable, applications of efficient small-scale, heat-exchange FTS reactors and bifunctional catalysts to increase selectivity to a single high value product or class of products.[30, 31, 32, 33, 34, 35, 36, 37, 38, 39, 40, 41, 42, 43]

9.5.1 SMALL-SCALE HEAT EXCHANGE REACTORS

Two types of heat exchange reactors are prevalent at small scales: microchannel reactors and conventional large-diameter reactors. The pros and cons of each reactor are discussed in the following sections.

9.5.1.1 MICROCHANNEL REACTORS

The most developed (but not yet commercialized) small-scale heat exchange reactor designed for FTS is Velocys' microchannel cross-flow heat exchange reactor (Figure 9-2)[31, 37, 39, 77] for relatively small plants around 1,000 barrels per day (bpd) in size using their Co-based catalyst formulation. Microchannel reactors minimize heat and mass transport limitations to allow highly exothermic reactions, such as FTS, to occur isothermally at their maximum temperature, consistent with the desired selectivity to achieve high conversion (70% per pass) and productivity (greater than 1 g C_5+ per g catalyst per hour). The reactor module consists of stacked, parallel arrays of microchannels less than 0.2 inches deep that result in reduced land footprint and cost. When the catalyst activity and selectivity fall below an acceptable level, the catalyst can be periodically partially regenerated with H_2. The spent module can be replaced with a new one and sent back for catalyst replacement.

Velocys claims to have completed four pilot-scale demonstrations of its process and achieved a 5,000-hour run at its Oklahoma demonstration plant (USA) in 2018. It is currently developing two projects: Bayou Fuels in Mississippi (USA), which will convert wood waste to diesel for road transportation, and Altalto in Immingham, Lincolnshire (United Kingdom), which will convert household and commercial waste to sustainable aviation fuel. Both projects are slated to receive funding in 2021.[78]

It is unclear that the microchannel reactor can be cost-effectively commercialized for FTS, especially while running at high conversions per pass (70+%), unless the catalyst life cycle can be increased to more than four years. Deactivation can occur and propagate backwards due to oxidation and potential sintering from high-steam concentration (due to high conversions) at the downstream end of the reactor. Although the activity can be partially revived by intermittent H_2 treatment, the high steam environment can eventually lead to irreversible deactivation.

The limitations of microchannel reactors include the large volume of metal required per unit volume of the catalyst and their associated cost and potential deleterious catalytic activity due to the substantial metal surface area per unit volume of catalyst; preparing and uniformly packing attrition and crush resistant catalyst into small channels; high-pressure drop due to small particles and the associated cost and potential compression and over-packing of the catalyst, causing flow issues during the run; and effectively removing the catalyst from the spent module and repacking the module for reuse. There is no published information available from Velocys about preparing, packing the reactor with catalyst, and unpacking for reuse.

9.5.1.2 LARGE-DIAMETER HEAT EXCHANGE REACTORS

As microchannel reactors have limitations, another option could be conventional one- to two-inch diameter reactor tubes with a mechanism inserted to rapidly transfer heat form the center of the tube to the wall. Some concepts to develop the rapid radial heat transfer capability from two-inch tubes include coating of the catalyst on a monolith[34] inserted into the tube and installing conductive packing around the catalyst in the tube. The objective is to rapidly transfer the heat to the reactor wall, where it can be removed using boiling water circulation. Alternatively, the concept of loading the catalyst using a conductive packing is currently being explored.[38, 79, 80]

Intramicron offers a Cu-based microfiber entrapment technology[80] that was used in two- to three-inch diameter, bench-scale, fixed-bed FTS reactors with entrapped Co catalyst to convert coal/biomass derived syngas to jet fuel and diesel range hydrocarbons,[38] whereas a highly open structure aluminum foam with catalyst loaded was demonstrated in two-inch diameter FTS reactors by Fratalocchi et al.[79] Both Cu-based microfiber and aluminum foam were able to rapidly remove heat to the walls to maintain isothermal conditions. They potentially address the microchannel reactor limitations of high-pressure drop, catalyst packing, catalyst removal, and catalyst replacement.

9.5.2 SELECTIVE AND BIFUNCTIONAL CATALYSTS

Several studies have examined selective catalysts and bifunctional catalysts (i.e., metals combined with zeolites) to make products other than hydrocarbons or selectively produce a single product or class of products in a single step using FTS variations (Table 9-5).

Table 9-5. Single-step synthesis using selective and bi-functional catalysts

TARGET PRODUCT	CATALYST	COMMENTS
Ethanol and higher alcohols[81]	Cu-Co, MoS_2	Widely researched subject; insufficient selectivity and yield to justify commercialization
Dimethyl ether[82]	Cu-Zn-Al_2O_3/Clinoptilolite	Insufficient selectivity
Jet fuel and diesel[35, 38]	Co/Zeolite	75% selectivity to diesel or jet fuel range compounds; solid wax-free product
Gasoline[36]	Fe+ZSM-5	60 to 70% selectivity to gasoline range hydrocarbons, no wax production
Light olefins[83]	ZnCrOx/AlPO-18	86% selectivity (CO_2 free) to C_2-C_4 olefins
C_5+ Hydrocarbons[36]	30 wt% Co/0.1 wt% Ru/SiC	91% C_5+ selectivity at 77 % CO conversion

From 2005 to 2010, there was a push to develop a selective process for the direct conversion of syngas to ethanol, because of its anticipated demand as a gasoline additive. These efforts were unsuccessful. Conversely, the use of hybrid F-T zeolite catalysts has produced wax-free products with high gasoline- and diesel-range selectivity. Another development is the use of a methanol synthesis catalyst combined with an aluminophosphate catalyst to directly and selectively produce C_2-C_4 olefins with high selectivity and yield. A traditional Co catalyst on a silicon carbide support has yielded exceptional selectivity for C_5+ hydrocarbons (91% at 77% CO conversion).

9.6 SUMMARY

This chapter addresses the reactor module and catalysts for commercial syngas conversion processes that produce H_2, methanol, SNG, and F-T liquids. The processes to make H_2 and methanol from syngas are in widespread commercial use in the United States and overseas. On the other hand, SNG and FTS processes have primarily been commercialized overseas.

The processes for the conversion of syngas to H_2, SNG, and methanol are highly developed and offered commercially by several process suppliers and catalyst vendors. Although these processes are moderately to highly exothermic, they all use FBRs due to FBRs' relative simplicity compared to fluidized-bed or slurry reactors (the latter two reactors can provide better temperature control). The main reason is that commercial catalyst formulations having high activity and selectivity are well established for FBRs. Furthermore, FBR designs offered by the vendors incorporate the required temperature control by limiting conversion per pass using product recycling, shell and tube reactors, and/or multiple reactors in series with indirect intercooling or direct quench.

There are three different catalysts used for H_2 production from syngas through the WGS reaction. These include the high-temperature Fe-Cr catalyst and low-temperature Cu-Zn catalyst for shifting clean syngas and low-to-medium temperature Co-Mo catalyst for shifting sour syngas. Higher temperatures result in a higher reaction rate, but conversion is limited by equilibrium. Thus, some commercial designs use the first reactor with the Fe-Cr catalyst to take advantage of higher reaction rates, followed by a second reactor with a low-temperature Cu-Zn catalyst to take advantage of favorable equilibrium conditions. For coal-derived syngas, the sour shift Co-Mo catalyst that operates in its sulfided state is preferred, because it helps to streamline the overall process by shifting the sour syngas prior to sulfur removal.

Water-gas shift is not only used to make H_2: it is also a necessary initial step to adjust the H_2/CO ratio for SNG and methanol synthesis. Following sour-shift, the gas is cooled to remove water and cleaned using established solvent-based processes, such as Rectisol® or Selexol™, that remove sulfur and CO_2. SNG synthesis is selectively carried out using a Ni-based catalyst at 350°C (662°F) and requires an H_2/CO molar ratio in the range of 3 to 4 to convert both CO and residual CO_2 to CH_4. Methanol synthesis is selectively carried out using a Cu-Zn/Al_2O_3 catalyst at 220°C (428°F) and requires an H_2/CO molar ratio in the range of 2 to 3 to convert both CO and residual CO_2 to methanol.

Unlike the highly selective processes that produce single products (i.e., H_2, SNG, or methanol), FTS is a non-selective process that produces a wide slate of alkanes, alkenes, and oxygenates. It has been commercialized, but is not in widespread use, likely because its poor selectivity leads to high cost. There are many options to carry out FTS, depending on the source of syngas (and the selected catalyst (Fe or Co). It can be carried out in a fluidized-bed reactor using an Fe-based catalyst for HTFT or in FBRs or SBCRs using a Co- or Fe-based catalyst for LTFT. HTFT is complex and produces a variety of products involving several downstream unit operations. LTFT is suitable to produce naphtha, jet fuel, diesel-range hydrocarbons, and waxes that are primarily straight chain alkanes in the heavier range. The products from LTFT do not require as extensive downstream processing as those from HTFT.

Large GTL plants have been commissioned for LTFT using a Co catalyst in both FBRs and SBCRs, but are costly. There are no new plans to build large-scale GTL plants (at the time of publication). However, China is developing large CTL plants using an Fe catalyst in an SBCR. Fe catalysts are more compatible with syngas from coal gasification for FTS as opposed to Co catalysts, but they are prone to greater deactivation and mechanical attrition than Co catalysts in SBCRs, so they are better suited for FBRs.

New trends in FTS are being driven by the need to reduce costs and risks associated with large plants. These trends include the development of small, modular plants for niche applications; efficient fixed-bed heat exchange reactors; and bi-functional catalysts to selectively produce a single product or class of product.

Endnotes

1 S. Rönsch, J. Jens Schneider, S. Matthischke, M. Schlüter, M. Götz, J. Lefebvre, P. Prabhakaran, and S. Bajohr (2015) "Review on Methanation – From Fundamentals to Current Projects", Fuel 166:276-296.

2 PetroWiki (June 3, 2015) "Gas to methanol", *https://petrowiki.org/Gas_to_methanol*.

3 A. Galadima and O. Muraza (2015) "From synthesis gas production to methanol synthesis and potential upgrade to gasoline range hydrocarbons: A review", J. Natural Gas Sci. Eng. 25:303-316.

4 K.A. Ali, A.Z. Abdullah, and A.R. Mohamed (2015) "Recent development in catalytic technologies for methanol synthesis from renewable sources: A critical review", Renewable and Sustainable Energy Reviews 44:508-518.

5 J. Lange (2001) "Methanol synthesis: a short review of technology improvements", Catal. Today 1-2:3-8.

6 M. Marchionna, M.D. Girolamo, L. Tagliabue, M.J. Spangler, and T.H. Fleisch (1998) "A Review of Low Temperature Methanol Synthesis", Studies Surf. Sci. Catal., In Natural Gas Conversion V 119:539-544.

7 S. Khadzhiev, N. Kolesnichenko, and N. Ezhova (2016) "Slurry Technology in Methanol Synthesis (Review)", Petroleum Chemistry 56(2):77-95.

8 B. Balopi, P. Agachi, and Danha (2019) "Methanol Synthesis Chemistry and Process Engineering Aspects - A Review with Consequence to Botswana Chemical Industries", Procedia Manufacturing 35:367-376.

9 H. Schultz (1999) "Short History and Present Trends of Fischer-Tropsch Synthesis", Appl. Catal. A: General 186:3-12.

10 M.E. Dry and J.C. Hoogendoorn (1981) "Technology of the Fischer-Tropsch Process", Catal. Rev. Sci. Eng. 23(2):265-278.

11 C. Frohnig, H. Kolbel, and M. Ralek (1982) "Fischer-Tropsch Process" in Chemical Feed Stocks from Coal, J. Falbe, ed., John Wiley and Sons, New York, 309-432.

12 M.E. Dry (1982) "Catalytic Aspects of Industrial Fischer-Tropsch Synthesis", J. Molecular Catal. 17(2-3):133-144.

13 M.E. Dry (1988) "The Sasol Route to Chemicals and Fuels", Studies Surf. Sci. Catal. 36:447-456.

14 M.M.G. Sneden (1998) "The Shell Middle Distillate Synthesis Process: Commercial Plant Experience and Outlook into the Future", Petrole et Techniques 415:94-97.

15 A.P. Steynberg, R.L. Espinoza, B. Jager, and A.C. Vosloo (1999) "High Temperature Fischer-Tropsch Synthesis in Commercial Practice", Appl. Catal. A: General 186(1-2):41-54.

16 M.E. Dry (2002) "The Fischer-Tropsch Process 1950-2000", Catal. Today 71:227-241.

17 A.P. Steynberg (2004) "Introduction to Fischer-Tropsch Technology", in Fischer-Tropsch Technology, Studies Surf. Sci. Technol 152, A.P. Steynberg, and M.E. Dry, eds., Elsevier B.V., Chapter 1.

18 A.P. Steynberg, M.E. Dry, B.H. Davis, and B.B. Breman (2004) "Fischer-Tropsch Reactors", in Fischer-Tropsch Technology, Studies Surf. Sci. Technol 152, A.P. Steynberg, and M.E. Dry, eds., Elsevier B.V., Chapter 2.

19 M.E. Dry (2004) "Chemical Concepts Used for Engineering Purposes", in Fischer-Tropsch Technology, Studies Surf. Sci. Technol 152, A.P. Steynberg, and M.E. Dry, eds., Elsevier B.V., Chapter 3.

20 A.P. Steynberg and M.E. Dry. (2004) "Commercial FT Process Applications", in Fischer-Tropsch Technology, Studies Surf. Sci. Technol 152, A.P. Steynberg and M.E. Dry, eds., Elsevier B.V., Chapter 5.

21 L.P. Dancuert, R.de Haan, and A. de Klark (2004) "Processing of Primary Fischer-Tropsch Products", in Fischer-Tropsch Technology, Studies Surf. Sci. Technol 152, A.P. Steynberg and M.E. Dry, eds., Elsevier B.V., Chapter 6.

22 M.E. Dry (2004) "FT Catalysts", in Fischer-Tropsch Technology, Studies Surf. Sci. Technol 152, A.P. Steynberg and M.E. Dry, eds., Elsevier B.V., Chapter 7.

23 A. Forret, J. M. Schweitzer, T. Gauthier, R. Krishna, and D. Schweich (2006) "Scale-up of Slurry Bubble Column Reactors", Oil & Gas Sci. Technol-Rev. IFP 61(3):443-458.

24 I.B. Kamara and J. Coetzee (2009) "Overview of High Temperature Fischer-Tropsch Gasoline and Diesel Quality" Energy & Fuels 23:2242-2247.

25 D.C. Denton, B. Turk, and R. Gupta (2011) "Syngas Cleanup, Conditioning, and Utilization", in Thermochemical Processing of Biomass: Conversion into Fuels, Chemicals, and Power, R.C. Brown, ed., John Wiley & Sons, Ltd., Chapter 4:78-123.

26 NETL (2019) "Gasifipedia. 10.2. Fischer-Tropsch Synthesis", *https://www.netl.doe.gov/research/coal/energy-systems/gasification/gasifipedia/ftsynthesis*. Accessed December 11, 2019

27 NETL (2019) "Gasifipedia. 10.2.1. Commercial Use of Fischer-Tropsch Synthesis", *https://www.netl.doe.gov/research/Coal/energy-systems/gasification/gasifipedia/sasol*.

28 J. van der Loosdrecht, F.G. Botes, I.M. Ciobica, A. Ferreira, P. Gibson, D. Moodley, A.M. Saib, J.L. Visagie, C.J. Weststrate, and J.W. Niemantsverdriet (2013) "Fischer-Tropsch Synthesis: Catalysts and Chemistry", in Comprehensive Inorganic Chemistry II. From Elements to Applications 7:525-557.

29 X. Lu (2011) "Fischer-Tropsch Synthesis: Towards Understanding", PhD Thesis, Univ. Witwatersrand, Johannesburg, South Africa.

30 C. Cao, J. Hu, S. Li, W. Wilcox, and Y. Wang (2009) "Intensified Fischer-Tropsch Synthesis Process with Microchannel Catalytic Reactors", Catal. Today 140:149-156.

31 S.R. Deshmukh, A.L.Y. Tonkovich, K.T. Jarosch, L. Schrader, S.P. Fitzgerald, D.R. Kilanowski, J.J. Leron, and T.J. Mazanec (2010) "Scale Up of Microchannel Reactors for Fischer-Tropsch Synthesis", Ind. Eng. Chem. Res. 49:10883-10888.

32 V.G. Kiryachek, and S.A. Filipchenko (2011) "Fischer-Tropsch Slurry Process for Small and Medium Scale Plants Processing Associated Petroleum Gas", Russian J. General Chem. 81(12):2604-2610.

33 N. Hooshyar, D. Vervloet, F. Kapteijh, P.J. Hamersma, R.F. Mudde, and J.R. van Ommen (2012) "Intensifying the Fischer-Tropsch Synthesis by Reactor Structuring - A Model Study", Chem. Eng. J. 207-208:865-870.

34 W. Liu, Y. Wang, W. Wilcox, and S. Li (2012) "A Compact and High Throughput Reactor of Monolithic-Structured Catalyst Bed for Conversion of Syngas to Liquid Fuels", AIChE J. 58(9):2820-2829.

35 C. Kibby, K. Jothimurugesan, T. Das, H.S. Lacheen, T. Rea, and R.J. Saxton (2013) "Chevron's Gas Conversion Catalysis-Hybrid Catalysts for Wax-free Fischer–Tropsch Synthesis", Catal. Today 215:131-141.

36 Q. Zhang, K. Cheng, J. Kang, W. Deng, and Y. Wang (2014) "Fischer-Tropsch Catalysts for the Production of Hydrocarbon Fuels with High Selectivity", ChemSusChem 7:1251-1264.

37 S. Le Viness, S.R. Deshmukh, L. Richard, and H.J. Robota (2014) "Velocys Fischer-Tropsch Synthesis Technology—New Advances on State of the Art", Topics in Catal. 57:518-525.

38 S.K. Gangwal and K. McCabe, (2015), "Small-scale Coal-Biomass to Liquids Production Using Highly Selective Fischer-Tropsch Synthesis", Final Report, DOE Award Number: DE-FE0010231, September.

39 P. Stratmann (2017) "Rollout of Smaller Scale Gas to Liquids: Current Commercialization and Future Approach", presented at Advanced Bioeconomy Leadership Conference, Washington, DC, March 1-3.

40 C. Zhu and G.M. Boollas (2018) "Gasoline Selective Fischer-Tropsch Synthesis in Structured Bifunctional Catalysts", Applied Catal. B: Environmental 235:92-102.

41 W. Zhu, K. Cheng, J. Kang, C. Zhou, V. Subramanian, Q. Zhang, and Y. Wang (2019) "New Horizon in C1 Chemistry: Breaking the Selectivity Limitation in Transformation of Syngas and Hydrogenation of CO_2 into Hydrocarbon Chemicals and Fuels", Chem. Soc. Rev. 48:3193.

42 N. Li, F. Jiao, X. Pan, Y. Chen, J. Feng, G. Li, and X. Bao (2019) "High-quality Gasoline Directly from Syngas by Dual Metal Oxide-Zeolite (OX-ZEO) Catalysis", Angew. Chem. Int. 58:7400-7404.

43 J. Su, H. Zhou, S. Liu, C. Wang, W. Jiao, Y. Wang, C. Liu, and Y. Ye (2019) "Syngas to Light Olefins Conversion with High Olefin/Paraffin Ratio Using ZnCrOx/Alpo-18 Bifunctional Catalysts", Nature Communications 10:1297.

44 R.B. Anderson, R.A. Friedel, and H.H. Storch (1951) "Fischer-Tropsch Reaction Mechanism Involving Stepwise Growth of Carbon Chain", J. Chem. Physics 19(3):313-319.

45 R.C. Brady and R. Pettit (1981) "On the Mechanism of Fischer-Tropsch Reaction. The Chain Propagation Step", J. Am. Chem. Soc. 103:1287-1289.

46 P.M. Maitlis (1989) "A New View of the Fischer-Tropsch Polymerization Reaction", Pure and Appl. Chem. 61(10):1747-1754.

47 G.P. van der Laan and A.A.C.M. Beenickers (1999) "Kinetics and Selectivity of the Fischer-Tropsch Synthesis", Catal. Rev. Sci. Eng. 41(3-4):255-318.

48 B. Shi, and B.H. Davis (2003) "Fischer-Tropsch Synthesis: Evidence for Chain Initiation by Ethene and Ethanol for an Iron Catalyst", Topics in Catal. 26(1-4):157-161.

49 M. Claes, and E. van Steen (2004) "Basic Studies", in Fischer-Tropsch Technology, Studies Surf. Sci. Technol 152, A.P. Steynberg and M.E. Dry, eds., Elsevier B.V., Chapter 8.

50 H. Schultz (1999) "Short History and Present Trends of Fischer-Tropsch Synthesis", Appl. Catal. A: General 186:3-12.

51 K. Jothimurugesan, J.G. Goodwin, S.K. Gangwal, and J.J. Spivey (2000) "Development of Fe Fischer–Tropsch Catalysts for Slurry Bubble Column Reactors", Catal. Today 58:335-344.

52 H. Jehangiri, J. Bennett, P. Mahjaonbi, and K. Wilson (2014) "A Review of Advanced Catalyst Development for Fischer-Tropsch Synthesis of Hydrocarbons from Biomass-Derived Syngas", Catal. Sci. Technol. 4:2010.

53 P. Nikolaidis and A. Poullikkas (2017) "A Comparative Overview of Hydrogen Production Processes", Renewable and Sustainable Energy Reviews (67):597-611.

54 R. Chaubey, S. Sahu, O.O. James, and S. Maity (2013) "A review on development of industrial processes and emerging techniques for production of hydrogen from renewable and sustainable sources", Renewable and Sustainable Energy Reviews (23):443-462.

55 M. Voldsund, K. Jordal, J., and R. Anantharaman (2016) "Hydrogen production with CO_2 Capture", Int. J. of Hydrogen Energy (41):4969-4992.

56 Global Syngas Technologies Council (2020) "Syngas Conditioning & Purification, Water Gas Shift", *https://www.globalsyngas.org/syngas-conditioning-and-purification/water-gas-shift/*.

57 NETL (2019) "Gasifipedia. 6.2.6. Water Gas Shift & Hydrogen Production", *https://www.netl.doe.gov/research/coal/energy-systems/gasification/gasifipedia/water-gas-shift*. Accessed December 11, 2019.

58 D. Sheldon (2017) "Methanol Production – A Technical History", Johnson Matthey Technol. Rev. 61(3):172-182.

59 NETL (2019) "Gasifipedia. 10.3.2. LPMEOH™ Process." *https://netl.doe.gov/research/Coal/energy-systems/gasification/gasifipedia/lpmeoh-process*. Accessed December 11, 2019.

60 D. Sheldon (2017) "Methanol Production – A Technical History", Johnson Matthey Technol. Rev. 61(3):172-182.

61 EIA (December 8, 2020) "Short-Term Energy Outlook, Global liquid fuels," *https://www.eia.gov/outlooks/steo/report/global_oil.php*.

62 EIA (October 4, 2017) "Today in Energy, Global gas-to-liquids growth is dominated by two projects in South Africa and Uzbekistan", *https://www.eia.gov/todayinenergy/detail.php?id=33192*.

63 Acumen Research and Consulting (February 6, 2019) "Coal to Liquid (CTL) Market Size Worth USD 5.8 billion by 2026: Acumen Research and Consulting", *https://www.globenewswire.com/news-release/2019/02/07/1711835/0/en/Coal-to-Liquid-CTL-Market-Size-Worth-USD-5-8-billion-by-2026-Acumen-Research-and-Consulting.html*.

64 EIA (February 5, 2018) "Today in Energy, China surpassed the United States as the world's largest crude oil importer in 2017", *https://www.eia.gov/todayinenergy/detail.php?id=34812*.

65 Stratas Advisors (February 15, 2017) "37 XTL Plants in Operation; More Under Construction". *https://stratasadvisors.com/Insights/2017/021517-XTL-Growth*.

66 Syngaschem (December 12, 2016) "Synfuels China demonstrates First Fischer-Tropsch products from the new Shenhua Ningxia Coal-to-Liquids plant", *https://www.syngaschem.com/2016/12/12/synfuels-china-demonstrates-first-fischer-tropsch-products-from-the-new-shenhua-ningxia-coal-to-liquids-plant/*.

67 Y. Cui (n.d.) "Synfuels China CTL Technologies", *https://www.usea.org/sites/default/files/event-/Yongbin%20Cui%20EN.pdf*.

68 Y.-W. Li, Y. Yang, L. Zhou, J. Gao (2018) "CTL/GTL Technology based on Medium Temperature Fischer-Tropsch synthesis (MTFT): Fundamentals and Applications", presented at ISCRE25, Florence Italy, May 20-23, 2018 *https://www.aidic.it/iscre25/review/papers/516li.pdf*.

69 Institute of Energy Process Engineering and Chemical Engineering (2020) 10th Annual Freiberg Conference: Organizers *https://gasification-freiberg.com/organizer/*.

70 C. Maretto, and R. Krishna (1999) "Modeling of a Bubble Column Slurry Reactor for Fischer-Tropsch Synthesis", Catal. Today 52:279-289.

71 J. W. A. De Swart, R. Krishna, and S.T. Sie (1997) "Selection, Design and Scale Up of the Fischer-Tropsch reactors", Studies in Surf. Sci. and Catal. 107:213-218.

72 S.K. Gangwal (2000) Private communication with Clariant.

73 K. Jothimurugesan, J.G. Goodwin, S.K. Gangwal, and J.J. Spivey (2000) "Development of Fe Fischer–Tropsch Catalysts for Slurry Bubble Column Reactors", Catal. Today 58:335-344.

74 R. Zhao, J.G. Goodwin, K. Jothimurugesan, S.K. Gangwal, and J.J. Spivey (2001) "Spray-Dried Iron Fischer-Tropsch Catalysts. 2. Effect of Carburization on Catalyst Attrition Resistance", Ind. Eng. Chem. Res. 40:132-1328.

75 CompactGTL (2021) "Projects", *http://www.compactgtl.com/about/projects/*.

76 NFRA XTL Technology (2021) "M100", *https://en.infratechnology.com/products/m100*.

77 D. Atkinson and J. McDaniel (2010) "Microchannel Reactors in Fuel Production", *http://wiki.gekgasifier.com/f/ptq_2q_microchannel_reactors.pdf*.

78 Velocys (2020) "A commercially demonstrated technology solution", *https://www.velocys.com/technology/*.

79 L. Fratalocchi, C.G. Visconti, G. Groppi, L. Lietti, and E. Tronconi (2018) "Intensifying heat transfer in Fischer-Tropsch tubular reactors through the adoption of conductive packed foams", Chem. Eng. J. 349:829-837.

80 IntraMicron (2020) "Process Intensified Reactors", *https://www.intramicron.com/process-intensified-reactors*.

81 V. Subramani and S.K. Gangwal (2008) "A Review of Recent Literature to Search for an Efficient Catalytic Process for the Conversion of Syngas to Ethanol", Energy & Fuels 22:814–839.

82 R. Khoshbin and M. Haghighi (2013) "Preparation and Catalytic performance of Cu-ZnO-$Al2O_3$/Clinoptilolite Nanocatalyst for Single-step Synthesis of Dimethyl Ether", J. Nanosci. Nanotechnol. 11:4996-5007.

83 J. Su, H. Zhou, S. Liu, C. Wang, W. Jiao, Y. Wang, C. Liu, Y Ye., L. Zhang, Y. Zhao, H. Liu, D. Wang, W. Yang, Z. Xie, and M. He (2019) "Syngas to light olefins conversion with high olefin/paraffin ratio using ZnCrOx/AlPO-18 bifunctional catalysts", Nature Communications 10(1297):1-8.

10.0 GUIDELINES FOR THE DESIGN OF SYNGAS REACTION PRODUCT PURIFICATION AND UPGRADING MODULES

10.1 INTRODUCTION

The design and operation of catalytic reactors for the conversion of syngas from gasification to fuels and chemicals is discussed in Chapter 9. The primary products of syngas reactions include hydrogen, substitute natural gas (SNG), methanol, or Fischer-Tropsch (F-T) liquids. The products may contain unconverted reactants and side products that need to be removed; alternatively, it could be advantageous to upgrade via additional reactions to higher value secondary products before removal. For small modular plants, these products may need to be upgraded and/or purified onsite (and thus the project must include equipment/modules for such upgrading and/or purification), or, in some cases, it may be possible to transport/sell the crude products directly to a larger remote site, such as a chemical/petrochemical site or a refinery, where any necessary upgrading and/or product purification can be accomplished by blending into similar crude streams. Such remote upgrading and/or purification would normally be limited to only those crude products that exist in liquid or solid form due to transportation limitations of gaseous products. This chapter discusses options and design considerations when onsite upgrading and/or purification is needed.

Hydrogen and SNG need to undergo purification to achieve the required purity for in-plant use or commercial sale. Methanol can be purified using contaminant removal processes to produce a final methanol product that meets the required commercial specifications. It can also be used as an intermediate product in the same plant or in a separate, preferably nearby, plant, where it could be subject to further reactions followed by purification to produce final products, such as formaldehyde, dimethyl ether (DME), acetic acid, gasoline, or olefins.

Unlike methanol, which is a single product, F-T liquids include a wide slate of hydrocarbons that can range from C_5-C_{60}+ normal and branched paraffins, and can also contain significant amounts of C_5-C_{15} olefins. Thus, they need to be separated into individual product streams and undergo additional upgrading steps for conversion into drop-in liquid fuels, such as diesel (C_{10}-C_{20}) and jet fuels (C_8-C_{16}). A portion of the C_{21}+ F-T waxes can undergo hydro-isomerization to lubricating oil or the entire wax stream can be subjected to mild hydrocracking to produce additional C_8-C_{20} range hydrocarbons that can increase the jet fuel and diesel yields.

This chapter focuses on the purification and upgrading of the primary gasification-based syngas reaction products (i.e., hydrogen, SNG, methanol, or F-T liquids) to final products or secondary products of the required purity to be marketed. The processing of each primary product (i.e., hydrogen, SNG, methanol, and F-T liquids) is discussed in the following sections. The use of compact separation equipment and process intensification approaches can help to modularize the overall plant to reduce cost. Novel approaches, such as the use of compact equipment and combined reaction/separation, that could enable cost reduction for small, modularized plants are also discussed.

10.2 HYDROGEN

Hydrogen is primarily made by subjecting syngas derived from coal/biomass gasification or steam reforming of natural gas to the water-gas shift (WGS) reaction. The raw hydrogen produced is further purified to meet specifications. Hydrogen is either used near the plant where it is produced or stored for transportation as a liquid and/or in pressurized cylinders for end users, except along the U.S. Gulf Coast where there are some installed hydrogen transport pipelines from the large number of refineries and petrochemical plants in close proximity. Hydrogen is used in several large-scale industrial processes, notably for upgrading refinery products, ammonia production, metallurgical operations, and hydrogenation of fats and oils. An emerging smaller-scale end use of hydrogen is as a fuel in fuel cells to make electricity for light duty vehicles.

10.2.1 PRODUCTION OF RAW HYDROGEN

As of 2018, the majority of hydrogen (approximately 95%) is produced from fossil fuels by steam reforming of natural gas, partial oxidation of methane, and coal gasification.[1] Other methods of hydrogen production include biomass gasification and electrolysis of water. A recent article offers an even greater percentage (98%) of hydrogen production from fossil fuels, stating that of the 70 million metric tons of hydrogen produced worldwide each year, 76% comes from reforming natural gas, 22% from coal gasification, and 2% from electrolysis of water.[2]

Most of the nearly 10+ million metric tons of hydrogen produced annually in the United States is made by medium- to large-scale plants that use steam methane reforming (SMR), which is the most cost-effective method of hydrogen production.[3, 4] For F-T synthesis (FTS), reforming is also used to convert the methane content into carbon monoxide and hydrogen, as methane is undesirable for FTS. The potential markets for large-scale hydrogen production are in countries where viable solid fuels are abundantly available and used and there is a dearth of natural gas and/or its supporting infrastructure. [5]

Sour water-gas shift (SWGS) is carried out in a series of adiabatic fixed-bed reactors (two to three) to produce raw hydrogen from syngas (see Chapter 9). SWGS provides a more efficient integration of the overall plant and converts the difficult to remove carbonyl sulfide (COS) in syngas to hydrogen sulfide (H_2S) via hydrolysis ($COS + H_2O \rightarrow CO_2 + H_2S$). The raw hydrogen stream contains high parts per million (ppm) to percentage levels of H_2S, high percentage levels of carbon dioxide (CO_2) and steam (H_2O), and some residual (less than 0.5%) unconverted carbon monoxide (CO). Bulk H_2S and CO_2 removal is carried out using well-established processes, such as Selexol™ or Rectisol® (see Chapter 6). However, the hydrogen stream from these processes may still contain residual H_2S, CO_2, H_2O, and CO that needs to be removed for in process or commercial use.

10.2.2 HYDROGEN END USES AND INFRASTRUCTURE

Nearly all the hydrogen produced worldwide is used for ammonia production for the fertilizer industry, petroleum refining processes (e.g., hydrodesulfurization, hydrocracking and hydrotreating), metal ore reduction, specialty chemical production, and hydrogenation of fats and oils for the food processing industries. Smaller uses for hydrogen include rocket fuel, the production of hydrogen peroxide, and for fuel cells in hydrogen-powered vehicles and other electrical energy uses.

Hydrogen-powered fuel cell vehicles do not emit carbon and are less dangerous than liquid fuel-powered vehicles.[6] Hydrogen is much lighter than air, so it rapidly disperses in the event of a leak, thereby minimizing the possibility of accumulation and ignition. However, due to a relatively poor hydrogen transportation, storage, and fueling infrastructure and competition from improvements in battery technology for electric cars, the economic growth of hydrogen-powered fuel cell vehicles has been limited. There have been proposals in many countries, including the United States and the United Kingdom, to transport some hydrogen along with natural gas and recover it downstream at the delivery point (if necessary).[7,8] This appears attractive in the United States due to its massive natural gas infrastructure. However, this approach has not made any significant progress (at the time of publication), possibly due to the required amount and purity of the natural gas to be delivered by the distributor and the potential of leaks and pipe embrittlement due to hydrogen exposure.

The choice of hydrogen production scale affects the cost and method of delivery. Central production plants can produce hydrogen at a relatively low cost due to economies of scale, but the delivery costs are significant, whereas distributed production facilities have relatively low delivery costs, but the hydrogen production costs are likely to be more significant than centralized production due to economies of scale. Liquid hydrogen is currently the most cost-effective form of hydrogen to transport over relatively longer distances by truck or rail. Liquid hydrogen production facilities are limited, likely due to the low demand for hydrogen as a final product and the high cost associated with liquefying hydrogen. Substantial hydrogen pipeline networks exist along the U.S. Gulf Coast.

There are several companies that can supply nearly pure hydrogen or provide an onsite SMR-based hydrogen production system. Major vendors and contractors include Air Liquide (Lurgi), Air Products, Messer, Praxair/Linde, and Hydrogenics. Liquid hydrogen installations are typically used in high-volume, permanent commercial applications. On the other hand, compressed hydrogen tube trailers are typically used in low-volume commercial applications or temporary demonstration projects. Tube trailers provide convenient and portable fueling solutions, but are generally more expensive than permanent liquid hydrogen installations. The very high-purity hydrogen (i.e., 99.999 to 99.9999%) needed for electronic applications is generally supplied in individual small cylinders.

10.2.3 HYDROGEN PURITY REQUIREMENTS AND PURIFICATION

Nearly all end uses require high-purity hydrogen (see Section 10.2.2). The International Organization for Standardization (ISO) 14687:2019 document specifies the minimum quality characteristics of hydrogen and is currently in use in the United States pending a future revision by the Society of Automotive Engineers.[9] The values in Table 10-1, adapted from ISO 14687:2019, indicate the highly stringent hydrogen purity and quality requirements.

Table 10-1. Hydrogen purity and quality requirements

SPECIFICATION	VALUE
Hydrogen (minimum, %)	99.97
Total Other Gases (maximum, ppmv*)	300
H_2O (maximum, ppmv)	5
Total NMHC** (maximum, ppmv)	2
Methane (CH_4; maximum ppmv)	100
Oxygen (O_2; maximum, ppmv)	5
Helium (He; maximum, ppmv)	300
Nitrogen (N_2) and Argon (Ar; maximum, ppmv)	300
CO_2 (maximum, ppmv)	2
CO+HCHO+HCOOH (maximum, ppmv)	0.2
Total Sulfur (maximum, ppmv)	0.004
Ammonia (maximum, ppmv)	0.1

Table 10-1. Hydrogen purity and quality requirements

SPECIFICATION	VALUE
Halogenated Compounds (maximum, ppmv)	0.05
Particulate Size (maximum, µm)	10
Particulate Concentration# (maximum, µg/L)	1

** Parts per million by volume*
*** Non-methane hydrocarbons*
 # At normal temperature and pressure

Pressure swing adsorption (PSA) is the preferred commercial method to achieve and exceed the required hydrogen purity and quality of ISO 14687:2019.[5, 10, 11, 12] For hydrogen purification, PSA is based on the adsorption of impurities from a hydrogen-rich stream, typically by a molecular sieve of a certain pore size (typically five Ångstroms [5Å]). Gaseous impurities (e.g., CO, CO_2, H_2S, nitrogen, and oxygen [O_2]) are adsorbed by the molecular sieve, whereas hydrogen simply passes through, separating from the impurities and getting delivered at high pressure to the next downstream process.

PSA is a cyclic adsorption-desorption process. The adsorption is typically carried out at room temperature and at the pressure of the hydrogen-rich gas to be purified. Several fixed-bed adsorption vessels are employed and the feed gas is switched from one adsorption vessel to another. The complete cycle consists of five steps (Table 10-2), so six or preferably more beds are usually installed to run the cycles while achieving steady-state operation.[10] Total cycle time for a bed is on the order of minutes. High-cycle, tight shutoff valves are a critical component of reliable PSA operation. As adsorption is taking place in one vessel, the adsorbent in another vessel is being regenerated. The impurities are subsequently desorbed at low pressure into an off-gas stream, producing an extremely pure hydrogen product. The low-pressure off-gas stream, or tail gas, usually has a low British thermal unit (Btu) value, but can still be used as a fuel for fired heaters elsewhere in the process.

Table 10-2. Required steps for a PSA cycle

STEP		DESCRIPTION
1	Adsorption	Adsorption using a freshly regenerated bed from Step 5 in same direction (co-current) as Step 2 to produce high-purity hydrogen stream at high pressure. A small portion is sent to Step 5 to repressurize another bed.
2	Co-Current Depressurization	Recovery of hydrogen trapped in void spaces from Step 1 by partial depressurization. This hydrogen is first sent for some time to Step 4 to purge the bed and then to Step 5 to repressurize.
3	Counter-Current Depressurization	Final depressurization in the opposite direction to blowdown. Desorption of impurities into a low-pressure, off-gas stream.
4	Purge	Purge indicated in Step 2 using the same direction as Step 3.
5	Counter-Current Repressurization	Use of high-pressure hydrogen recovered from Steps 1 and 2 to fully repressurize the bed, making it ready for the Step 1 of the next cycle.

The sequence of Table 10-1 is generally controlled by a programmable logic controller (PLC), which manipulates the switching valves to maintain hydrogen purity. By using monitoring equipment, the control system can identify any problems with the process, locate the source, and alert the operator. Product purities exceeding 99.999% can be achieved.

10.2.4 RESEARCH TRENDS IN GASIFICATION-BASED HYDROGEN PRODUCTION

Hydrogen is a clean, carbon-free fuel, as well as a feedstock for many chemical processes. The main deterrent to a large-scale, hydrogen-based energy and supply infrastructure is that it is cumbersome and expensive to store and transport hydrogen in large quantities.[13, 14] In addition, since hydrogen is mainly produced from fossil fuels, a significant quantity of CO_2 is co-produced and needs to be removed from the product stream and stored, but it is costly to do so. The use of biomass instead of coal or coal-biomass mixtures could possibly avoid the need for CO_2 storage, as biomass is carbon-neutral (neglecting transportation and yield losses).

A gasification-based, large, centralized hydrogen supply infrastructure needs to include a gasification plant (with CO_2 storage) that generates pure hydrogen, a storage system that can hold a sufficient amount of hydrogen, and a system to transport the stored hydrogen to customers. Coal gasification plants currently producing hydrogen in China and other countries are essentially supplying hydrogen locally.[5]

Although a subject of interest and significant research and development (R&D) funding by the U.S. Department of Energy (DOE) and foreign countries, an economically attractive solution to the issue of large-scale hydrogen storage and transportation over long distances is yet to be developed.[5, 13, 14] Also, a number of technologies and process intensification approaches are under consideration in the United States for reducing the cost of gasification with CO_2 capture and for the development of small, location-specific modular gasification systems to reduce the risk associated with huge capital investments for large, centralized gasification plants (see Chapter 9). The following sections describe three selected technologies that relate to gasification-based hydrogen production.

10.2.4.1 HYDROGEN COPRODUCTION IN AN INTEGRATED GASIFICATION COMBINED CYCLE PLANT USING MEMBRANE TECHNOLOGY

Gas-separation membrane technology has been a subject of intense research and development (R&D) activities over the past 30 years, because of its operational and structural simplicity, but it is limited to relatively small-scale applications due to large-scale manufacturing challenges and the inability to achieve economy of scale.[15] A membrane that can separate hydrogen with nearly 100% selectivity would be needed to meet hydrogen purity requirements. Palladium-based membranes, when operated at elevated temperatures around 350 to 500°C (approximately 662 to 932°F), can provide nearly 100% selectivity, because hydrogen is the only gas among common gases present that dissolves and diffuses through the essentially nonporous palladium membrane.[15, 16] Also, by using an asymmetric large pore metal membrane as a support for a thin palladium layer, the hydrogen flux through the membrane can be sufficiently increased for hydrogen production from gasification.

The production of more than one product (e.g., hydrogen, methanol, F-T liquids, and/or electricity) can result in a variety of gasification plant configurations. One such scenario is the coproduction of pure hydrogen (using 10 to 20% of the hydrogen in syngas) and electricity with CO_2 capture using the following two options: (1) hydrogen is separated using a PSA system for a portion of the clean syngas (Figure 10-1) or (2) the entire clean syngas stream is passed through a hydrogen selective membrane to produce hydrogen, while the remaining hydrogen in the syngas is used to produce electricity after CO_2 capture (Figure 10-2).[17]

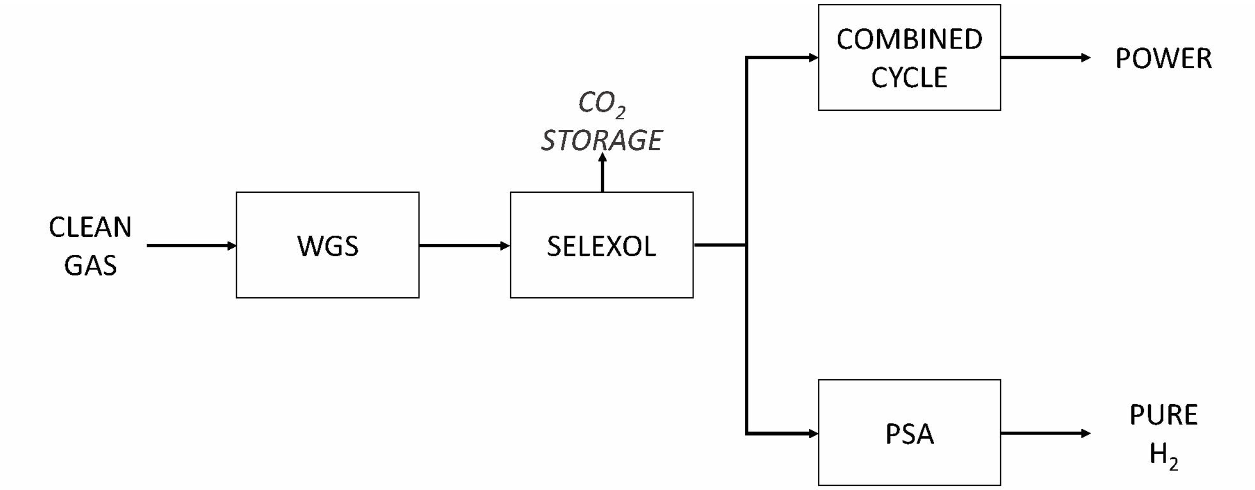


Figure 10-1. Coproduction of hydrogen (using PSA) and electricity with CO_2 capture

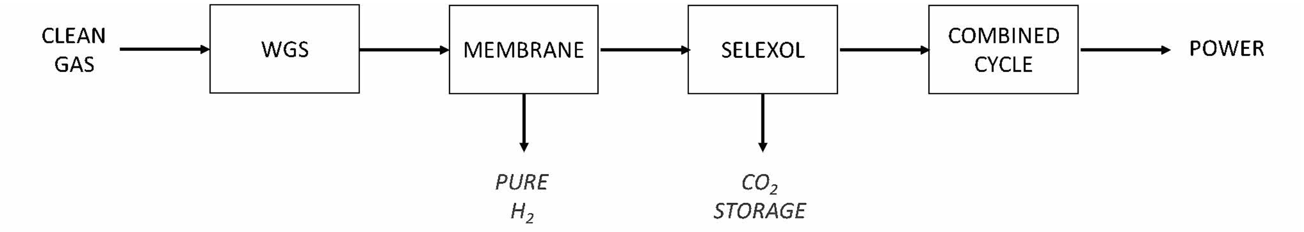


Figure 10-2. Coproduction of hydrogen (using membrane) and electricity with CO_2 capture

Based on a techno-economic analysis (TEA) study of the 335-megawatt-electric (MWe) Puertollano integrated gasification combined cycle (IGCC) power plant in Spain using a 50-50 coal and petroleum coke blend, Giuliano et al.[17] showed that both options had similar economics, lowered the additional cost of electricity, and reduced the cost of CO_2 capture.

Since palladium membranes operate at temperatures suitable for WGS, an interesting possibility of process intensification is to use the membrane as a WGS reactor by packing the membrane with WGS catalyst to accomplish simultaneous shift and hydrogen separation. Hydrogen separation shifts the equilibrium toward production of more hydrogen along with CO_2, because the shift reaction is equilibrium-limited, resulting in a reduction of the size of the reactor for the same amount of hydrogen production. This approach could be of interest in small modular plants where it is necessary to decrease the cost of equipment to reduce the economy-of-scale penalty due to scale down.

When employing palladium membranes, it is important to consider the deleterious effect of contaminants, particularly H_2S that can poison/foul the membrane by sulfidation. It is recommended to deep clean the syngas before the membrane using an amine or other solvent followed by a guard bed to reduce H_2S to low parts per billion (ppb) levels before the shift reactors. Also, rather than a sour shift reactor, either a high-temperature shift reactor or a low-temperature shift reactor (or both in series) should be employed if a membrane-incorporated reactor is under consideration.

ECN in Netherlands offers pre-commercial palladium membranes made using electroless plating.[16, 18] However, challenges associated with scale up and fouling by contaminants appear to have impacted full commercialization. PSA, on the other hand, is a fully commercial technology for hydrogen separation (see Section 10.2.3). PSA is much more tolerant of contaminants like H_2S, because it gets adsorbed by the PSA system and rejected in the off-gas stream.

10.2.4.2 SORBENT ENHANCED WATER-GAS SHIFT

In a manner like membrane-incorporated WGS reactors, the goal in sorbent enhanced water-gas shift (SEWGS) is also to shift the equilibrium toward the production of more hydrogen. This is accomplished by simultaneously adsorbing CO_2 and carrying out WGS using an adsorbent-catalyst mixture in a fixed-bed catalytic reactor-adsorber.[5, 19] When the CO_2 is about to break through the bed, the adsorber is switched and the used adsorber is regenerated in a manner similar to that of a PSA to release pure CO_2 ready for compression and storage. A conventional IGCC plant with CO_2 capture is compared to an IGCC plant with SEWGS in Figure 10-3 and Figure 10-4. The objective of SEWGS is to simultaneously accomplish both enhanced WGS and CO_2 capture to reduce the cost of equipment. SEWGS could be a useful technology for large IGCC systems and small, modular gasification systems to produce hydrogen and/or electricity.

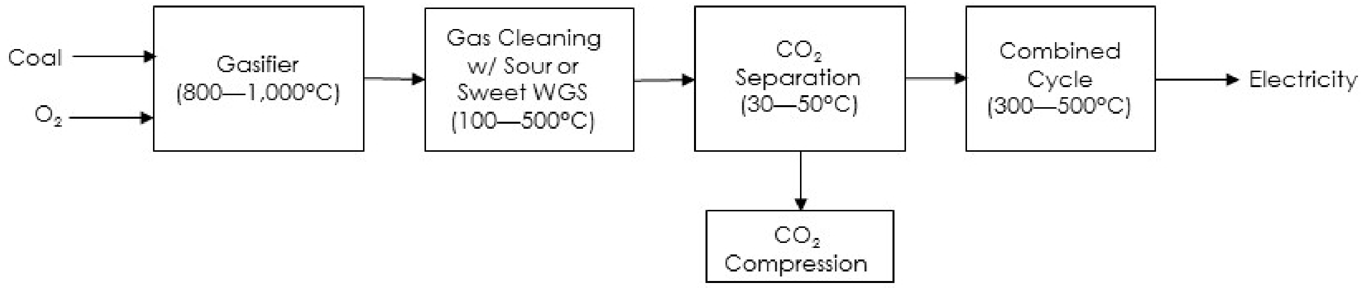


Figure 10-3. Conventional IGCC system with CO_2 capture

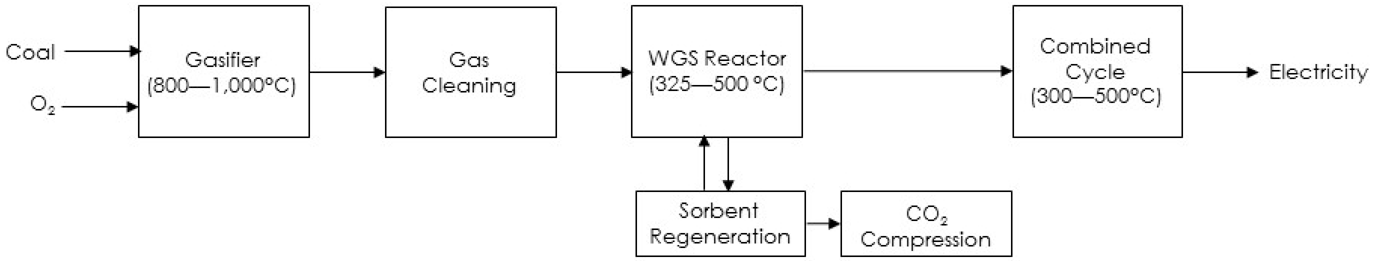


Figure 10-4. IGCC system with SEWGS

SEWGS has been a major research topic over the past 20 years.[20, 21] It could possibly be employed in steps, like a PSA system (Table 10-2). Several projects on sorbent development and scaleup have been funded by DOE's National Energy Technology Laboratory (NETL) and worldwide. Early studies evaluated naturally occurring calcium oxide-based sorbents (e.g., calcite, dolomite), but they suffered from sintering at their high operating temperatures. Synthesized, high-temperature sorbents (e.g., lithium silicates, sodium zirconate) did not degrade during cycling, but required very high regeneration temperatures to recover CO_2.

The research focus shifted to lower temperatures around 1995. Sodium oxide (Na_2O)-promoted alumina and potassium carbonate (K_2CO_3)-promoted hydrotalcite sorbents showed high selectivity for CO_2 chemisorption at 250 to 450°C (482 to 842°F), but CO_2 recovery was poor at 550°C (1,022°F). Most recently, sorbent development has shifted toward promoted magnesium oxide (MgO) sorbents.[21] These sorbents have shown very high CO_2 capacity, regenerability, and durability at 325 to 350°C (617 to 662°F) with a temperature swing of around 25°C (77°F), but slowly degraded over 500 to 1,000 cycles during regeneration in the presence of steam in a bench-scale reactor. Based on these past efforts, it is unlikely that SEWGS for hydrogen production and/or CO_2 capture in IGCC can be commercially applied unless the following challenges are met:

- Development of a highly durable sorbent that can release CO_2 without (or with) very low temperature increase. Like PSA for hydrogen separation, high capacity may not be as important a requirement.
- Heat management for the exothermic adsorption and endothermic regeneration cycle in a packed-bed reactor.

10.2.4.3 HYDROGEN STORAGE/TRANSPORT USING TOLUENE-METHYL CYCLOHEXANE CYCLE

Efficient hydrogen storage and transportation technologies are needed to encourage widespread use of hydrogen as a source of power and fuel for automobiles (see Section 10.2.4). Although coal gasification provides the lowest cost hydrogen production route in China, it is not widely commercialized due to a lack of low-cost, large-scale storage and transportation technologies, limited availability, and the high cost of hydrogen-fueled vehicles.[5] Currently, commercial hydrogen storage and transportation technologies (e.g., compressed cylinders, liquid hydrogen) are costly and potentially hazardous (especially at large scale). Technologies under development (e.g., metal hydride, metal-organic framework, and salts like sodium borohydride) suffer from very low capacity at room temperature and/or high cost.

Liquid, aromatic hydrocarbons have recently attracted attention as a hydrogen carrier due to their high hydrogen storage capacity at room temperature, relative safety, and ease of transportation. In particular, the toluene-hydrogen-methylcyclohexane (THM) reaction cycle (shown below) can reach storage and transport capacity up to 6.2 wt%, because of its durability and ease of reversibility and is being demonstrated at a scale of 200 metric tons/year by Chiyoda Corporation in Japan:[5, 22]

$$CH_3C_6H_5 + 3H_2 \rightarrow CH_3C_6H_{11} \qquad (\text{R10.1 - } \Delta H_{250^{\circ}C} = -241 \text{ kJ/mol})$$

$$CH_3C_6H_{11} \rightarrow CH_3C_6H_5 + 3H_2 \qquad (\text{R10.2 - } \Delta H_{450^{\circ}C} = 261 \text{ kJ/mol})$$

The THM cycle consists of hydrogenating toluene to methylcyclohexane (MCH) at a hydrogen production location, transporting liquid MCH to another location where hydrogen is needed, dehydrogenating MCH to toluene and hydrogen, and finally transporting the toluene back to the hydrogen production location to complete the cycle. The reactions are carried out in the vapor phase over hydrogenation and dehydrogenation catalysts under relatively mild reaction conditions of approximately 1 MPa and 250°C/482°F (hydrogenation) and 1 MPa and 400°C/752°F (dehydrogenation). The liquid products are condensed by cooling them to approximately 100°C/212°F, which separates hydrogen during dehydrogenation.

Chiyoda has carried out research on improving the efficiency of MCH dehydrogenation catalysts since 2002. During a 10,000-hour, lab-scale test, their dehydrogenation catalyst maintained an overall MCH conversion above 95% and a selectivity of toluene and hydrogen over 99%. In May 2020, hydrogen transportation by sea from Brunei Darussalam to a gas turbine at the Mizue Thermal Power Plant was successfully started by Chiyoda Corporation—the first demonstration of relatively large amounts of hydrogen transportation by sea.

Nonetheless, further improvements are potentially needed for the dehydrogenation step. Dehydrogenation of MCH is a highly endothermic reaction and requires high temperatures that can cause decomposition to low molecular weight hydrocarbon gases. The dehydrogenation catalyst can also gradually deactivate due to carbon deposition. Further improvements of the dehydrogenation catalyst and the use of non-noble metal catalysts could potentially advance the TMH technology closer to commercialization.

An alternative approach to hydrogen storage and transport is to react the hydrogen with nitrogen to produce ammonia, which can be readily stored and/or transported as a liquid, and then reconvert the ammonia to hydrogen at the intended use location and/or time. The downside of this approach is the need for additional equipment, and thus additional capital investment, to produce ammonia and subsequent regeneration of hydrogen.

10.3 SUBSTITUTE NATURAL GAS

SNG is produced by gasification of any carbonaceous fuel to produce a raw syngas followed by cleaning, WGS (to achieve the desired hydrogen to CO ratio), and catalytic methanation. The SNG must be further purified to meet natural gas pipeline specifications. Natural gas is cheaply and abundantly available in the United States, so SNG production is not commercially viable. The primary markets for SNG are China and other countries where natural gas is expensive or not readily available.

10.3.1 PRODUCTION AND INFRASTRUCTURE

Unlike hydrogen, natural gas in the United States enjoys a highly organized production, storage, and pipeline delivery infrastructure that could be available for SNG. In 2018, approximately 28 trillion cubic feet of natural gas was delivered to 75 million customers through 3 million miles of mainline and other pipelines that link natural gas production areas and storage facilities with consumers.[23] Also, unlike hydrogen, which is primarily used as a reactant for other processes, natural gas is primarily used as a fuel. The only non-combustion reactions of natural gas (methane) of any industrial significance are steam reforming to produce hydrogen itself (including hydrogen intended for conversion to ammonia/urea) and sequential halogenation of methane to produce methylene chloride (CH_2Cl_2), chloroform ($CHCl_3$), and/or carbon tetrachloride (CCl_4).

In the United States, the only currently operating SNG production facility is the Dakota Gasification Co. Great Plains Synfuels Plant (North Dakota), which was built in 1984. It produces more than 54 billion cubic feet of SNG and captures and sells approximately 100 million cubic feet of CO_2 for enhanced oil recovery. It also produces fertilizers, solvents, phenols, and other chemicals.

10.3.2 SUBSTITUTE NATURAL GAS PRODUCT REQUIREMENTS

The composition, pressure, and temperature of SNG produced from gasification must conform to the specification of the pipeline into which it is to be delivered and be interchangeable with the existing pipeline gas.[24, 25] As there is no general pipeline gas or SNG specification available, the operating conditions in the overall-SNG plant must be adjusted on a case-by-case basis to fit not only feedstock properties, but also the relevant product specification. Also, like a raw natural gas stream, SNG must also be treated to comply with emissions regulations. Typical pipeline gas specifications are shown in Table

10-3.[27] The gas must meet the pipeline's Wobbe Index (a function of the gas's higher heating value [HHV] and specific gravity) requirement,[26] which is typically about 1,350 Btu/standard cubic foot (scf). The index is an indication of interchangeability and the quality of operation of gas turbines and/or other combustion equipment to minimize emissions using the gas.

Table 10-3. Typical pipeline gas specification

CHARACTERISTIC	SPECIFICATION
Water	<4 to 7 lb/MMscf
H_2S	<0.25 to 1.0 grain/100 scf
Gross Heating Value	950 to 1,200 Btu/scf
Hydrocarbon Dewpoint	<14 to 40°F at specified pressure
Mercaptans	<0.25 to 1.0 grain/100 scf
Total Sulfur	<0.5 to 20 grain/100 scf
CO_2	2 to 4 mol%
N_2	4 to 5 mol%
$CO_2 + N_2$	4 to 5 mol %
O_2	<0.01 mol%
Dust, Gums, and Free Liquid	None
Typical Temperature	40 (min) to 120°F (max)
Typical Pressure	400 to 1,200 psig

10.3.3 FINAL SUBSTITUTE NATURAL GAS PURIFICATION

A major byproduct of methanation is water produced in equimolar or greater amounts compared to methane that ends up in raw SNG leaving the methanation process. Water can be removed to meet requirements by sufficiently cooling the gas exiting the methanation reactor (Table 10-3). The other requirements and the trace contaminants do not pose an issue in the final cleanup, because they all must be removed upstream of the methanation process to protect the nickel-based catalyst (Table 10-3). Nitrogen (N_2) (inert) limits can be met by using oxygen-blown gasification; CO_2, H_2S, and other trace contaminants can be removed using established processes like Rectisol® or Selexol™. The final gas is further cooled, cleaned, dried, compressed, and supplied to consumers via pipeline.[24]

SNG from gasification typically contains at least some hydrogen to nearly eliminate CO and reduce residual CO_2 via methanation. For example, the Topsøe Recycle Energy-efficient Methanation Process (TREMP) by Haldor Topsøe produces SNG containing around 3 vol% hydrogen, and the Lurgi process at the Great Plains Plant in North Dakota (USA) produces SNG with around 1 vol% hydrogen.[27] There is no specification for hydrogen in typical pipeline gas (Table 10-3). Thus, the hydrogen needs to be separated to produce liquefied natural gas (LNG) or pipeline quality gas (if required by the buyer).

The best (and cheapest) option is to sell the SNG to a pipeline operator without separating the hydrogen, rather than producing and selling LNG. If the 1 to 3 vol% hydrogen in SNG is required by the pipeline operator to be removed, PSA is not applicable, because the objective is to retain the pressure in methane rather than hydrogen. In contrast, PSA could be the preferred technology to separate CO_2 and nitrogen from the methane, as practiced by the natural gas industry.[28, 29] Currently, the best polymer-based hydrogen removal option appears to be hydrogen selective membranes made from polyaramide or polyimide that can achieve hydrogen/methane selectivities as high as 200.[15]

Hydrogen separation membranes and gas separation membranes (in general) are evolving and have mostly been applied in small-scale systems.[15] Asymmetric metallic membranes made using a perm-selective palladium layer are nearly 100% selective for hydrogen[16, 18] and could be considered. It may be possible to integrate them downstream of methanation prior to cooling, because they work at elevated temperatures around 350 to 500°C (approximately 662 to 932°F). Membrane-based separation technology (in general) could be attractive for smaller scale modular gasification-based SNG plants.

10.4 METHANOL

Methanol is produced by catalytic conversion of syngas derived from either steam reforming of natural gas or gasification of coal and other solid feedstock. The raw methanol produced must be further purified to meet the required commercial specifications. Methanol is used as a blend fuel with gasoline in some countries and serves as a building block to produce several industrially important products, including formaldehyde, DME, methyl tertiary-butyl ether (MTBE), acetic acid, gasoline, and olefins.

10.4.1 RAW METHANOL PRODUCTION

Global methanol production in 2016 was around 85 million metric tons at more than 90 plants worldwide with a full production capacity of nearly 130 million metric tons.[30, 31] Commercial methanol plants are large with capacities of some plants as high as 5,000 to 7,000 metric tons per day. Global methanol demand reached 75 million metric tons in 2015, driven in large part due to emerging energy applications for methanol (e.g., its use as a gasoline additive and for conversion to olefins, gasoline, DME, and MTBE), which now account for 40% of methanol consumption.

The synthesis of methanol is carried out by catalytic conversion of syngas derived from natural gas reforming or gasification using a highly selective copper-zinc (Cu-Zn)-based catalyst (see Chapter 9). Before the catalytic conversion step, the syngas is deep cleaned and its H_2-to-($CO+CO_2$) molar ratio adjusted by WGS to optimize methanol yield. When using SWGS, the deep cleaning, especially sulfur removal, is carried out after the shift. Very high methanol selectivities (reaching 99.5%) can be achieved using modern Cu-Zn catalysts.

The only commercial-scale gasification-based methanol plant in the United States is the highly successful Eastman Chemical Company chemical production facility located in Kingsport, Tennessee. The methanol product is used primarily to produce various acetyl products (acetic anhydride, acetic acid, and methyl acetate).

In the United States, only a few large, natural gas-based methanol plants are under construction (see Section 10.4.2). In contrast, China—the largest methanol producer in the world since 2006—is employing coal gasification for methanol production.[33, 32, 33] About two-thirds of China's methanol feedstock is produced from coal; the remainder is produced from coke oven gas and natural gas. Due to its abundant coal resources, China has significantly increased its capacity to manufacture methanol using coal as a feedstock over the past 10 to 15 years.

Although the carbon selectivity to methanol in methanol synthesis is more than 99%, the nearly 1% of the remaining carbon gets converted to side products, such as ethanol, acetone, or fusel oils. The so-called raw methanol produced from synthesis must be further purified to meet standards (see Section 10.4.3).

10.4.2 MAJOR U.S. PLANTS, CONVERSION PRODUCTS, AND END USES

As discussed in Chapters 2, 6, and 9, methanol is a building block for a variety of end products, including gasoline and liquified petroleum gas (LPG) additives, and it is also used as a blend fuel with gasoline in China.[34] However, the largest uses include (1) vapor-phase catalytic oxidation to produce formaldehyde (safely stored as formalin, which is 37% formaldehyde in water), (2) liquid-phase catalytic conversion to acetic acid via methanol carbonylation, and (3) liquid-phase catalytic reaction with isobutene to produce MTBE.[34] Newer methanol plants tend to be very large to allow economy-of-scale benefits.

Methanol production in the United States is mainly derived from natural gas and is used primarily as a chemical feedstock rather than directly as a fuel. Although COVID-19 appears to have impacted demand (at the time of publication),[35, 36] three new large plants were planned to come online in the United States from 2018 to 2020 with a combined nameplate capacity of around 3.3 million metric tons per year, which would increase total U.S. methanol capacity to 9.4 million metric tons per year. The 5,000 metric tons per day Natgasoline methanol plant in Beaumont, Texas (USA), began operation in June 2018. During 2019 and 2020, two new large methanol plants (Big Lake and St. James) on the Gulf Coast (USA) were expected to come online. Proman USA's (formerly G_2X) Big Lake facility (Louisiana, USA) started operation, making some 3,800 metric tons per day of methanol, which may be later converted to gasoline.[38, 37] Yuhuang's St. James (Louisiana, USA) 4,700 metric tons per day methanol plant was expected to begin operation in 2020, however no reports were found to indicate operations (at the time of publication).

Blending methanol with gasoline allows refiners to extend China's gasoline supply and increase the octane level of its gasoline. China has a national quality standard for 85% blends of methanol; a national standard for a 15% blend of methanol in gasoline is pending approval by the government (at the time of publication). MTBE is also blended into gasoline in China to increase octane levels. In the 1990s, MTBE was used as an oxygenated octane enhancer in reformulated gasoline sold in the United States (to reduce emissions). However, concerns over groundwater contamination led to its complete phase out in the United States by 2005, although its use continues in other countries.

Methanol is also being converted directly to gasoline, olefins, and DME in China.[33, 34, 35,38, 39] China built its first methanol-to-gasoline (MTG) plant in 2010 and since then another 10 MTG plants have come online. This conversion has occurred much less often than the blending of methanol or its derivatives into gasoline, because high MTG capital costs and low oil prices have created uncertainty for investment in new MTG plants.

Methanol can also be converted to light olefins using the so-called methanol-to-olefins (MTO) process.[40] China's MTO production capacities grew from 1.1 mt/y in 2010 to 15.5 mt/y in 2017 mainly due to favorable government policies. A mega-scale MTO process provided by Honeywell/UOP came online in China in 2018.[41] At a production capacity of 833,000 t/y, the unit is the largest single-train MTO unit in the world. It converts methanol from domestic coal into olefins for plastics production. Like MTG, MTO requires a high capital investment and cannot compete with cheap oil-based olefin production in the United States using steam cracking of alkanes. The technologies associated with MTG and MTO are discussed in Section 10.4.5.2.

DME, a colorless gas under ambient conditions, can be made via dehydration of methanol over an alumina catalyst. It is extensively used in the chemical industry and as an aerosol propellent. However, its use as a fuel has been limited. In China, it was blended into LPG until 2006. A DME ban in 2006 due to rubber seal corrosion by DME in LPG cylinders,[34, 40, 41] coupled with the poor price competitiveness of DME relative to LPG, have significantly reduced the level of DME blending. DME has also been touted as a clean burning substitute for diesel in compression ignition engines,[42, 43] but it has not been commercialized for the diesel engine market.

10.4.3 TRANSPORTATION AND STORAGE INFRASTRUCTURE

Methanol has an extensive distribution network and storage capacity in place throughout most of the world.[44] More than 500,000 tons of methanol is transported by rail, boat, and trucks each month to diverse and scattered users across the United States. China had a significant imbalance in 2013 between locations of large-scale production using coal and large-scale industrial use.[35] In northwest China, 38% of China's methanol was produced, but only 4% was consumed nearby. In eastern China, 44% of the methanol was consumed, but only 8% was produced. Consequently, a large amount of methanol was transported across China. Approximately 82.6% of methanol was transported by trucks at a high cost and with high security risk, 9% by sea (most was imported), and 8.4% by train. No reports could be found to indicate that these imbalance, cost, and security issues have been overcome.

Overland rail transport (with each car holding around 100 tons) is the preferred option for long-distance transportation of bulk quantities. For smaller volumes and distribution to local markets, tanker trucks with capacities of up to 30 tons are generally used. Where inland waterborne shipment through rivers and canals is possible, methanol can be transported by barges, which typically contain around 1,250 tons (i.e., 10,000 barrels) of methanol. Another means of transporting large quantities of liquids is via pipelines, which is also used extensively for oil, natural gas, and their products. Technically, transporting methanol through pipelines does not pose a problem, but methanol pipelines are only viable in regions where major methanol producers and users are concentrated in close proximity, such as on the Texas Gulf Coast between Houston and Beaumont (USA). In the future, if methanol is increasingly used as a fuel, as practiced in China, the widespread use of pipelines could make economic sense.

When methanol is produced in remote locations where cheap natural gas is available, it is shipped throughout the world by large, dedicated methanol ocean tankers ranging in size from 15,000 to 100,000 dead weight tons to reduce the transportation cost per ton. Once delivered, methanol can be stored in large quantities, much like petroleum and its products, in tanks with capacities exceeding 12,000 tons.

10.4.4 METHANOL PURITY REQUIREMENTS AND PURIFICATION PROCESSES

The following sections describe methanol purity specifications and purification.

10.4.4.1 PURITY SPECIFICATIONS

Methanol needs to meet several purity targets set by the International Methanol Producers and Consumers Association (IMPCA). Commercial methanol is generally classified according to American Society for Testing and Materials (ASTM) purity grades A and AA.[45] Methanol is also classified as Grades A, B, and C by the U.S. General Services Administration (GSA).[46] GSA Grade A is for solvent use and is equivalent to ASTM Grade A, whereas Grade B is for chemical use and is the same as ASTM Grade AA. Both these grades are made synthetically (i.e., from syngas), whereas Grade C is a denaturing grade wood alcohol. Both ASTM Grades A and AA have purity specifications that are similar (with minor differences) to the requirements specified by the IMPCA.[47] A condensed version of the IMPCA purity requirements is shown in Table 10-4. The full requirements specify several parameters, including minimum methanol content, maximum content for several impurities, color, distillation range, and specific gravity, along with their analytical or ASTM measurement methods. Note that the minimum methanol content required is 99.85%. Higher purity methanol (greater than 99.9%) is made in smaller quantities for analytical laboratory instruments, such as a liquid chromatograph.

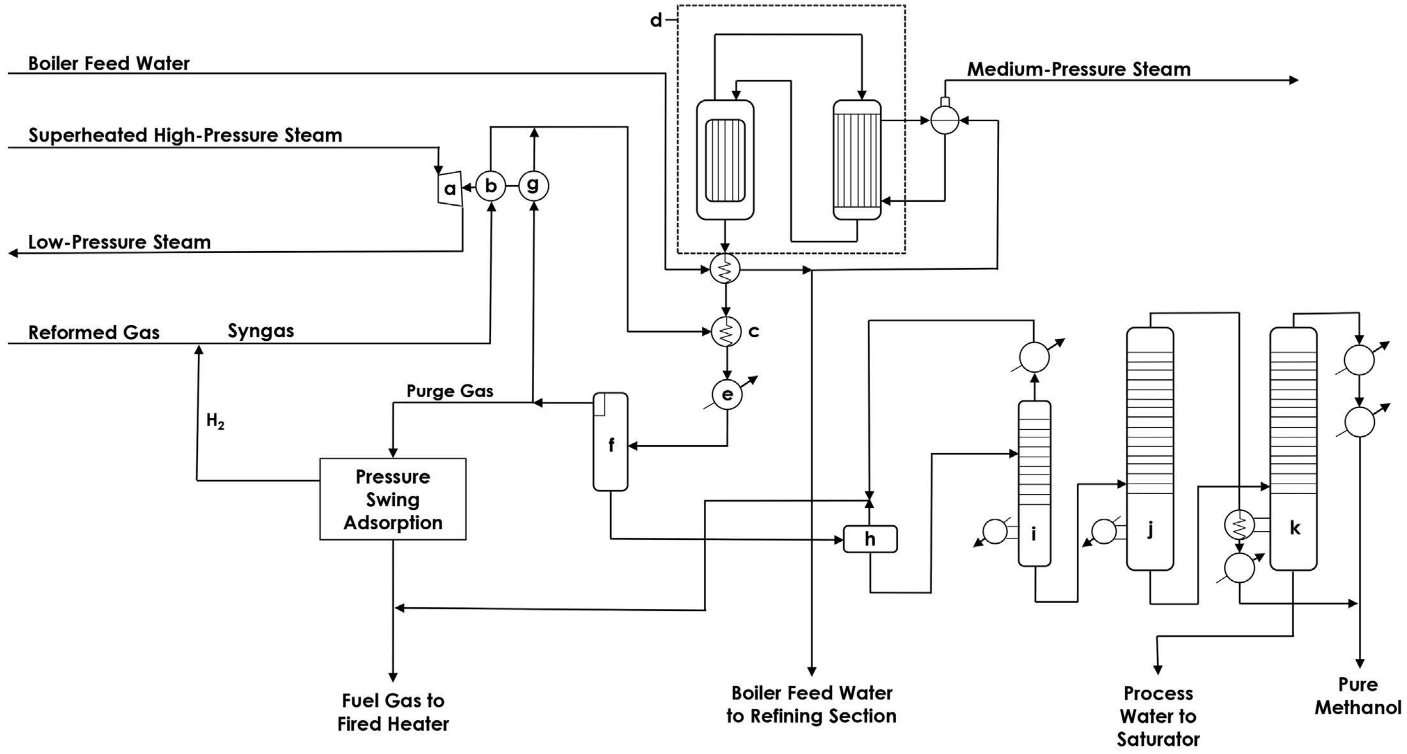
Table 10-4. IMPCA methanol purity requirements

Methanol (wt%)	99.85+ (dry basis)	Sulfur (mg/kg)	0.5 or less
Acetone (mg/kg)	30 or less	Acidity (as acetic acid)	30 mg/kg or less
Ethanol (mg/kg)	50 or less	Dissolved iron (mg/kg)	0.1 or less
Water (wt%)	0.1 or less	Non-volatiles	8 mg/1,000 ml or less
Chloride (mg/kg)	0.5 or less	Specific gravity (20°C)	0.791 to 0.793

10.4.4.2 PURIFICATION

Methanol can be synthesized from syngas with 99.0 to 99.5% carbon selectivity, but it needs further purification to meet the requirements (Table 10-4). The most prominent undesired products are higher alcohols, DME, formates, carboxylic acids, hydrocarbons, esters, and ketones.[48] The formation of these undesirable compounds can be promoted by catalyst impurities, such as alkali metals, the alumina support, and catalyst age. Methanol selectivity is generally controlled kinetically and methanol purity is mainly dominated by residence time and temperature. Crude methanol (after condensation of the synthesis reactor effluent) contains both low-boiling and high-boiling components (light and heavy ends) and significant amounts of water. The light ends include mainly dissolved gases (e.g., CO_2), DME, methyl formate, and acetone, and the heavy ends include higher alcohols, longer-chain hydrocarbons, higher ketones, and esters. In combination, all organic, undesirable products that have higher boiling points than methanol and lower boiling points than water are called fusel oils.[49]

Distillation columns, as shown in Lurgi's MegaMethanol process schematic (Figure 10-5), are used to produce high purity methanol to meet the specifications (Table 10-4). Distillation is the only separation technique that makes sense when separating bulk methanol from water and a much smaller and highly complex mixture of other organic compounds. Newer techniques, such as selective reverse osmosis membranes, cannot be applied for such a mixture.



(a) turbine for synthesis gas and recycle compressor; (b) synthesis gas compressor; (c) trim heater; (d) combined convertor system; (e) final cooler; (f) methanol separator; (g) recycle gas compressor; (h) expansion vessel; (i) light ends column; (j) pure methanol pressure column; (k) atmospheric methanol column

Figure 10-5. Lurgi MegaMethanol synthesis and purification process[50]

The crude methanol is generally first made slightly alkaline through the addition of small amounts of aqueous caustic soda to neutralize lower carboxylic acids and partially hydrolyze esters. Then the impurities are separated in at least two to three stages. The components boiling at a lower temperature than methanol are removed in a light-ends column. Pure methanol is then distilled overhead in distillation columns. If the columns operate at different pressures, the latent heat of the vapors of the currently operating high-pressure column can be used to heat the low-pressure column. If the process water contains slight impurities (e.g., the bottom product of the heavy ends column), either a side draw-off or an additional column for purification is required. Over the years, several variants related to the separation and/or use of fusel oils have appeared in the literature,[51, 50, 51] but the basic purification technology involving the addition of alkali and multiple distillation columns has remained the same.

10.4.5 NEWER TRENDS AND TECHNOLOGIES IN METHANOL PRODUCTION AND UTILIZATION

Methanol synthesis plants tend to be large scale on the order of 5,000 tons per day to achieve economies of scale. However, there has been recent interest in the development of small, modular distributed-scale methanol synthesis plants that reduce the financial risk associated with large plants. These plants may be a good fit for small, modular gasification plants. There may be opportunity for improvement of catalyst and reactor technology for the two relatively newer methanol conversion processes (i.e., MTG and MTO). Both MTG and MTO are currently commercially deployed in China at large scale using methanol made from coal gasification, but these plants have required financial backing by the government. If the costs associated with these processes can be reduced through technology improvements, there could be interest by private industry in deploying them in less risky, small, modular applications at niche sites where the products are locally needed, but unsuitable for long distance transportation.

10.4.5.1 SMALL-SCALE METHANOL SYNTHESIS

While there has been recent interest in the development of small methanol plants in the United States,[52, 52, 53, 54, 55, 56] there are no small plants in operation due to the current availability of cheap oil and gas and economies of scale from mega methanol plants all over the world. Also, the plants under consideration are slated to use municipal solid waste or waste gases, such as biogas and landfill gas, as feedstock. The key to the commercialization of small F-T or methanol plants[57] is reducing the cost of the reformer (gasifier in the case of coal or solid waste). Notable developments for small methanol plants appear to be from Johnson Matthey (JM) through Davy's compact reformer[54] and ThyssenKrupp through their renewable energy-based process for mini plants and Uhde's reformer for small-medium scale plants.[55]

The Davy compact reformer with BP's fixed-bed, F-T technology was first installed and proven over 18,000 hours in Nikiski, Alaska (USA), during the early 2000s to convert 90,000 Nm^3/day of natural gas to 300 bpd of synthetic crude. In early 2009, an improved compact reformer incorporating several design improvements was installed and successfully commissioned. The main differences between Davy's shell and tube design and conventional large systems are its ability to withstand thermal expansion during operation and use of a smaller size reforming catalyst. JM claims that their compact reformer is a highly heat integrated, small, lightweight, and modular factory-built unit that can be readily transported by road or rail for use in remote or offshore locations for small-scale methanol synthesis. ThyssenKrupp is another notable company offering methanol plants ranging from very small to very large (10 to 10,000 tons per day). Their mini-scale plants (ranging from 10 to 200 tons per day) use only renewable energy, whereas their small-medium scale plants (ranging from 200 to 3,000 tons per day) convert natural gas to syngas using Uhde's steam reforming technology.

10.4.5.2 METHANOL TO GASOLINE AND METHANOL TO OLEFINS

Both MTG and MTO are commercial processes currently deployed in China. Keil[58] provides an historical and technical review of these processes. MTG is considered the first major synfuels development after the discovery of F-T synthesis (FTS) in 1926. The MTG process, invented and announced by Mobil scientists in 1976, uses a zeolite catalyst (i.e., ZSM-5) that converts methanol to gasoline through consecutive reactions with DME and olefins as intermediates.

The first commercial installation of MTG started operation 10 years after initial discovery in Motunui, New Zealand, in 1986.[59, 60] It converted natural gas to methanol followed by MTG, producing a premium unleaded gasoline blend suitable for use directly in the transportation fuels market. The MTG portion of the plant was closed in 1997 due to the availability of cheap imported gasoline and is now only making methanol. During its operation, New Zealand's total gasoline requirement was around 42,000 bpd. The Motunui MTG plant supplied nearly a third of this total (approximately 14,000 bpd).

The MTG reaction sequence starts with dehydration of methanol to DME over a special alumina catalyst[60, 61] in a separate reactor to produce an equilibrium mixture of methanol, DME, and water. This mixture is then fed to the MTG reactor containing the zeolite catalyst. Further dehydration of DME to C_2-C_4 olefins occurs over the ZSM-5 catalyst and then these olefins react to selectively produce gasoline range olefins, n/iso-paraffins, aromatics, and naphthenes. The dehydration process produces a large amount of water. For example, the following is produced from 1,000 kg of methanol: 387 kg of gasoline, 46 kg of liquefied petroleum gas, 7 kg of fuel gas, and 560 kg of water. This MTG gasoline is compatible with petro-gasoline.

The first dehydration step to form the equilibrium mixture is carried out at 310 to 320°C (590 to 608°F) and 26 bar in a fixed-bed reactor. The reaction is exothermic, but controllable due to equilibrium. The mixture temperature rises to 350 to 366°C (approximately 662 to 691°F) and then it enters the fixed-bed MTG reactor at approximately 23 bar. Coke formation slowly occurs over the zeolite catalyst. This allows the use of fixed-beds reactors, such as those used in the Motunui, New Zealand, plant. Multiple fixed-bed reactors were used so that the process could continuously run using a sequence of reaction and periodic regeneration (i.e., coke burn off).[60, 62]

The MTO process was a logical offshoot of the MTG process, because highly desirable C_2-C_3 olefins are intermediate products of MTG. As opposed to ZSM-5, SAPO-34—a lower acidity zeolite—is used. SAPO-34 provides high olefin yield and selectivity and is considered an optimum catalyst for MTO. Also, the operating temperature is higher (470 to 490°C [878 to 914°F]) and operating pressure is lower to increase olefin yield.[60, 64] Through optimization of these conditions, olefin selectivities exceeding 90 to 95% can be achieved. MTO does not require the use of a separate DME reactor if maximizing ethylene yield is the goal. However, if propylene is the desired product, using a so-called methanol-to-propylene (MTP) process—an approach like MTG that uses a separate DME reactor—is needed.[63]

The use of fluidized-bed reactors for MTG has been demonstrated at bench to pilot scale, but has not been commercialized despite many advantages like good temperature control, constant product quality, and the ability to continuously remove and add catalyst. The drawbacks of using fluidized beds include the need for a highly attrition resistant catalyst that can withstand coking and coke burn off without mechanical weakening (to minimize loss), need for efficient dust control devices, and the delicate pressure balance and operational complexity compared to fixed-bed reactors. Despite these drawbacks, fluidized-bed reactors are preferred for MTO using the SAPO-34 catalyst due to the much shorter reaction-regeneration cycle caused by the higher temperature operation.

For the development of cost-effective, small-scale MTG- and MTO-type processes, a much higher level of process intensification is needed to reduce equipment cost. This explains Haldor Topsøe's TIGAS process development, which produces the required oxygenates (for gasoline production) directly from syngas using a different catalyst as opposed to selectively producing methanol, allowing the same pressure to be used throughout the process. Also, a combination of steam and auto-thermal reforming was used to produce the desired hydrogen to CO ratio without the need for hydrogen separation, and the gasoline synthesis reactor was directly integrated with the oxygenate synthesis reactor to reduce the number of recycle loops. Although a 1,000-tpd demonstration plant was built using this process and operated for three years, it has not been commercialized, presumably due to lower gasoline yield.

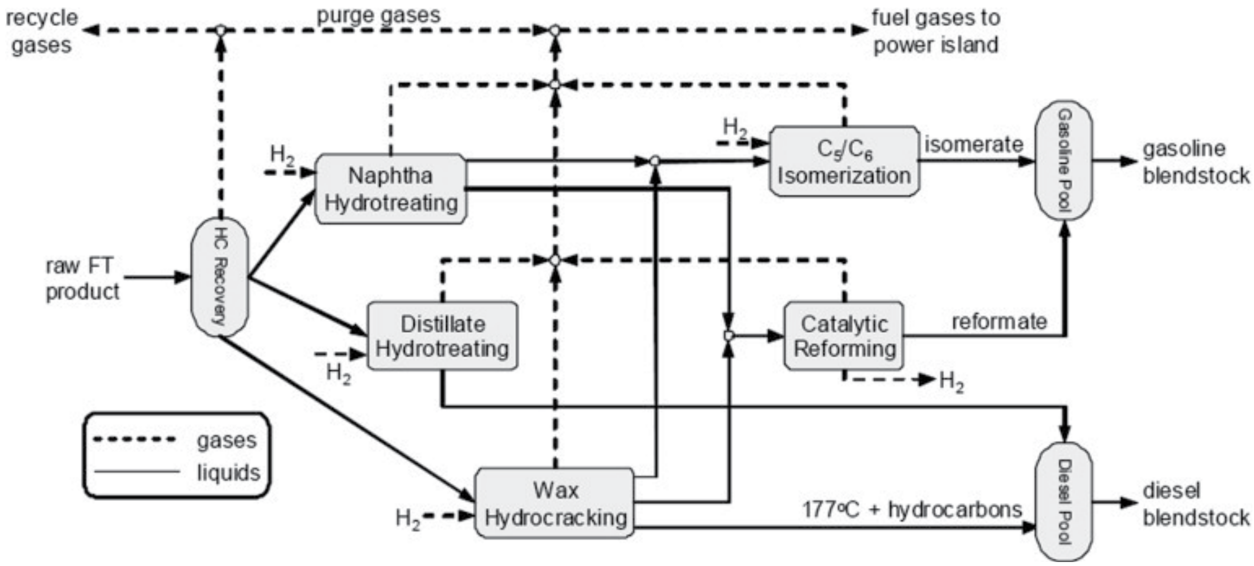
10.5 FISCHER-TROPSCH LIQUIDS

As discussed in Chapter 9, unlike methanol—a single selective compound made from syngas that can serve as a building block for a variety of end products—F-T liquids include a wide slate of C_5-C_{60+} paraffins, olefins, and oxygenates, so very limited separation and upgrading options are possible. F-T synthesis can be carried out at high temperature (i.e., high-temperature Fischer-Tropsch [HTFT]) and low temperature (i.e., low-temperature Fischer-Tropsch [LTFT]), but HTFT was eliminated from further consideration, because of its high complexity, like an oil refinery.

The simplest option for the large slate of LTFT liquids produced from conventional iron (Fe) or cobalt (Co) catalysts is as a premium zero sulfur blend with crude oil for processing in a refinery when the F-T plant is near the refinery and can be sold to the refinery. This may be the best option for mini to small modular gasification-based F-T plants. Small, conventional F-T plants cannot compete with low oil prices and large F-T plants due to economy-of-scale penalties. These plants make sense only at certain remote sites (e.g., Alaska or Hawaii, USA) or near military sites where oil transportation is costly. Even given this consideration, capital cost needs to be significantly reduced. Novel process intensification and modified catalyst approaches under development (see Chapter 9) need to be fully developed and commercialized before mini to small modular gasification-based F-T plants can be viable.

A better option for medium to large F-T plants is to separate the F-T products into the desired specific boiling point ranges of standard refinery cuts, such as naphtha (C_5-C_{12}), jet fuel (C_8-C_{16}), diesel (C_{10}-C_{20}), and wax (C_{21+}), by standard distillation methods followed by upgrading. This is currently practiced in the large Sasol/Chevron plants in Nigeria and Qatar and the Shell plant in Qatar.

Most compounds in LTFT liquids are n-paraffins, so using them directly as gasoline is not a practical option due to their low octane number. The best option for the naphtha range hydrocarbons (C_5-C_{12}) is to upgrade them using standard refinery type processes, such as steam cracking to make olefins or hydrotreating followed by isomerization and reforming to make gasoline. The paraffins in the C_{10}-C_{20} range make them an excellent candidate for blending with diesel, especially because of their high cetane number and zero sulfur content. Also, some of these same paraffins in the C_9-C_{16} range are desirable as jet fuel blends. A typical F-T liquid upgrading operation for making gasoline and diesel from F-T liquids is shown in Figure 10-6.[64]



Used with permission from Speight[66]

Figure 10-6. Raw F-T product upgrading schematic

The light gases are first separated and the higher boiling compounds are distilled to separate the naphtha, distillate, and wax fractions. All these fractions require at least some hydrogen for upgrading. The naphtha stream is hydrotreated to eliminate oxygenates and convert olefins to n-paraffins that are then split into two streams: a C_5-C_6 stream and a C_{7+} stream. The first steam is isomerized and the second stream is reformed to produce iso-paraffins and aromatics to boost their octane value before sending them to the gasoline pool. The distillate stream is also hydrotreated to remove oxygenates and saturate olefins, resulting in a finished diesel blend stock. The wax fraction is hydrocracked into a finished distillate stream and naphtha streams that augment the hydrotreated naphtha before the latter is sent for isomerization or reforming. Each of these blend stocks is of premium quality due to low aromaticity and zero sulfur content. If jet fuel is also the desired product, the schematic (Figure 10-6) can be modified as appropriate to produce the required amount of C_9-C_{15} kerosene-range hydrocarbons.

Waxes produced by the F-T process can be upgraded to either high-quality middle distillates by hydrocracking (Figure 10-6) or high-quality lubricant base oil by selective hydro-isomerization.[65] For both options, bifunctional catalysts containing a hydrogenation function and an acidic function are needed, as discussed in detail by Bouchy et al.[67] Since the wax feed is sulfur-free, the hydrogenation function is preferably provided by noble metals like palladium, rather than cobalt-molybdenum or nickel-tungsten-type catalysts that work in a sulfided state. The acidic function is provided by solid acids, such as silica-alumina or USY zeolite. Considering the effect of oxygenates present on the hydrogenation function, fine tuning of the relative strengths of the hydrogenation and acidic functions is required to produce an optimum catalyst for the specific wax composition to be processed.

Depending on where the LTFT-derived diesel is marketed, its specification must comply with local legislation, which may vary. The required LTFT diesel properties also depend on whether the LTFT diesel is used as-is or as a blend stock with another diesel.[66] In general, LTFT diesel can substitute for or be blended with petroleum-derived diesel. Additives may be required for sulfur-free F-T diesel to provide lubricity, which is a property of the sulfur impurities in conventional diesel. The low aromatics content of LTFT diesel results in a lower density than that of conventional diesel. As a result, its volumetric consumption in a compression ignition (CI) engine is higher than that of conventional diesel, but its brake-specific fuel consumption in the same system (expressed in mass of fuel per kWh generated) is better than that of the conventional product.

LTFT oils can be treated to meet the diesel and jet fuel ASTM and European specifications for petroleum-derived products.[67, 68] The required specifications for diesel include ASTM D975 in the United States and EN590 in Europe. For jet fuel, a maximum of 50% F-T-derived, kerosene-range paraffins can be blended with petroleum-derived jet fuel per ASTM D1655. This limit is necessary to have the required amount of aromatics to maintain engine seal performance.

10.6 SUMMARY

The primary products of syngas reactions include hydrogen, SNG, methanol, or F-T liquids. They may contain unconverted reactants and side products that need to be separated; it could be advantageous to upgrade them via additional reactions to higher value secondary products before separation.

Hydrogen and SNG must undergo separation/purification to achieve the purity required for in-plant use or commercial sale. PSA is the best option for hydrogen purification. SNG is clean, because of the deep cleaning requirements for syngas from which SNG is made, and is carbon-neutral if it is produced from biomass. It can contain hydrogen that can be separated using membranes.

Methanol can be purified by alkali addition and distillation to produce a final methanol product of the required commercial specifications. It can be used as an intermediate product either in the same plant or in a separate, preferably nearby plant where it could be subject to further reactions followed by purification to produce final products, such as formaldehyde, DME, acetic acid, gasoline, and olefins.

Unlike methanol, which is a single product, F-T liquids are a wide slate of hydrocarbons that can range from C_5-C_{60+} normal and branched paraffins and can also contain significant amounts of C_5-C_{15} olefins. They need to undergo further reactions, such as olefin hydrogenation, and additional upgrading steps for conversion into drop-in liquid fuels like diesel (C_{10}-C_{20}) and jet fuel (C_8-C_{16}). A portion of the C_{21}+ F-T waxes can undergo hydro-isomerization to lubricating oil or the entire wax stream can be subject to mild hydrocracking to produce additional C_8-C_{20}-range hydrocarbons that can increase the jet fuel and diesel yields.

Endnotes

1 International Renewable Energy Agency (IRENA) (2018) "Hydrogen Renewable Power: Technology Outlook for the Energy Transition", *https://www.irena.org/-/media/Files/IRENA/Agency/Publication/2018/Sep/IRENA_Hydrogen_from_renewable_power_2018.pdf*.

2 J. Rissman (October 7, 2019) "Hydrogen Could Become A $130 Billion U.S. Industry By 2050. Could It Also Cut Emissions?", Forbes *https://www.forbes.com/sites/energyinnovation/2019/10/07/how-hydrogen-could-become-a-130-billion-us-industry-and-cut-emissions-by-2050/#390f7d3b2849*.

3 EERE (2020) "Fact of the Month May 2018: 10 Million Metric Tons of Hydrogen Produced Annually in the United States", *https://www.energy.gov/eere/fuelcells/fact-month-may-2018-10-million-metric-tons-hydrogen-produced-annually-united-states*.

4 EIA (January 21, 2020) "Hydrogen explained: Production of hydrogen" *https://www.eia.gov/energyexplained/hydrogen/production-of-hydrogen.php*.

5 Z. Liu and Y. Li (2019) "Study on critical technologies and development routes of coal-based hydrogen energy", Clean Energy 3(3):202–210.

6 Ballard (October 2020) "Hydrogen Facts", *https://www.ballard.com/docs/default-source/spec-sheets/hydrogen-facts.pdf?sfvrsn=3930c280_0*.

7 Networks (March 17, 2016) "Could hydrogen piggyback on natural gas infrastructure?", *https://networks.online/gas/could-hydrogen-piggyback-on-natural-gas-infrastructure/*.

8 A. Witkowski, A. Rusin, M. Majkut, and K. Stolecka (2018) "Analysis of compression and transport of the methane/hydrogen mixture in existing natural gas pipelines", International Journal of Pressure Vessels and Piping 166:24-34.

9 International Organization for Standardization (November 2019) "ISO 14687:2019", *https://www.iso.org/standard/69539.html*.

10 Howe-Baker International (n.d.) "Recovery and Purification of Hydrogen Using PSA Technology", *https://howebaker.com/recovery-and-purification-of-hydrogen-using-psa-technology/*.

11 Air Liquide (2020) "Pressure Swing Adsorption (PSA) - Hydrogen Purification", *https://www.engineering-airliquide.com/pressure-swing-adsorption-psa-hydrogen-purification*.

12 C. Grande (2012) "Advances in Pressure Swing Adsorption for Gas Separation", ISRN Chemical Engineering 2012, Article ID 982934, doi:10.5402/2012/982934.

13 N. Sayfullin (2020) "Research Trends and Opportunities in Hydrogen Fuels", CAS, American Chemical Society, *https://www.cas.org/blog/hydrogen-fuels-research*.

14 D. Simbeck and E. Chang (2002) "Hydrogen Supply: Cost Estimate for Hydrogen Pathways- scoping Analysis" Mountain View, CA: SFA Pacific, Inc.

15 N. Sazali, M.A. Mohamed, and W.N.W. Salleh (2020) "Membranes for hydrogen separation: a significant review", Int. J. Adv. Manuf. Technol. 107:1859–1881.

16 T. Peters and A. Caravell (2019) "Pd-Based Membranes: Overview and Perspectives", Membranes 8(25):1-6. doi:10.3390/membranes9020025.

17 A. Giuliano, M. Poletto, D. Barletta (2018) "Pure hydrogen co-production by membrane technology in an IGCC power plant with carbon capture", Int. J. of Hydrogen Energy 43:19279-19292.

18 A. Doukelis, K. Panopoulos, A. Koumanakos, and E. Kakaras (2016) "Palladium Membrane Technology for Hydrogen Production, Carbon Capture and Other Applications", Johnson Matthey Technol. Rev. 60(3):190, doi:10.1595/205651316x691591.

19 W. Steen, C. Richardson, Y. Lu, H. Lu, and M. Rostam-Abadi (2014) "Sorbent Enhanced Water Gas Shift-Rethinking CO_2 Capture in IGCC", Energy Procedia 63:1686-1702.

20 W. Steen, C. Richardson, Y. Lu, H. Lu, and M. Rostam-Abadi (2014) "Sorbent Enhanced Water Gas Shift-Rethinking CO_2 Capture in IGCC", Energy Procedia 63:1686-1702.

21 S. Zhao, K. McCabe, and S.K. Gangwal (2019) "Combined Sorbent/WGS-based CO_2 Capture Process with Integrated Heat Management for IGCC Systems", Final Report, DOE Award Number DE-FE0026388.

22 Chiyoda Corporation (2017) "SPERA Hydrogen®: Chiyoda's Hydrogen Supply Chain Business", *https://www.chiyodacorp.com/en/service/spera-hydrogen/*.

23 EIA (December 3, 2020) "Natural gas explained: Natural gas pipelines", *https://www.eia.gov/energyexplained/natural-gas/natural-gas-pipelines.php*.

24 C. Wix, I. Dybkjær, and N.R. Udengaard (2007) "Coal to SNG – The Methanation Process", 24th Annual International Pittsburgh Coal Conference, Johannesburg, South Africa, September 10-14.

25 S. Mokhtab, W.A. Poe, and J.Y. Mak (2019) "Natural Gas Fundamentals", in Handbook of Natural gas Transmission and Processing, 4th Edition, ISBN 978-0-12-815817-3, Elsevier, Inc.

26 Mechanical engineering site (2017) "Wobbe Index and Modified Wobbe Index Formula", *http://www.mechanicalengineeringsite.com*.

27 W. Lin, J. Xu, L. Zhang, and A. Gu (2017) "Synthetic natural gas (SNG) liquefaction processes with hydrogen separation", International Journal of Hydrogen Energy, 42(29), 18417-18424.

28 J. Wills, M. Shemaria, and M.J. Mitariten (2004) "Production of Pipeline-Quality Natural Gas with the Molecular Gate CO_2 Removal Process", Society of Petroleum Engineers, SPE-87644-PA 19(1).

29 M. Mitariten (April 23, 2001) "New Technology Improves Nitrogen-Removal Economics", Oil and Gas J.

30 Methanol Institute (2020) "The Methanol Industry", *https://www.methanol.org/the-methanol-industry/*.

31 D. Sheldon (2017) "Methanol Production – A Technical History", Johnson Matthey Technol. Rev. 61(3):172-182.

32 EIA (February 23, 2017) "Today in Energy, China's use of methanol in liquid fuels has grown rapidly since 2000", *https://www.eia.gov/todayinenergy/detail.php?id=30072#*:~:text=Most%20of%20China's%20methanol%20supply,steel%20production)%20and%20natural%20gas.&text=Consumption%20of%20MTBE%20and%20other,230%2C000%20b%2Fd%20in%202016.

33 L.-W. Su, X.-R. Li, Z.-Y. Sun (2013) "The consumption, production and transportation of methanol in China: A review", Energy Policy 63:130-138.

34 B. Balopi, P. Agachi, and Danha (2019) "Methanol Synthesis Chemistry and Process Engineering Aspects - A Review with Consequence to Botswana Chemical Industries", Procedia Manufacturing 35:367-376.

35 EIA (February 21, 2019) “Today in Energy, New methanol plants expected to increase industrial natural gas use through 2020 *https://www.eia.gov/todayinenergy/detail.php?id=38412*.

36 S. McGinn (May 13, 2020) “North American methanol demand falls sharply”, *https://www.argusmedia.com/en/news/2105131-north-american-methanol-demand-falls-sharply*.

37 Proman (2019) “Companies”, *https://www.proman.org/companies/proman-usa/*.

38 C. Yang and R.B. Jackson (2012) “China’s growing methanol economy and its implications for energy and the environment”, Energy Policy 41:878-884.

39 *https://knowledge.energyinst.org/energy-matrix*.

40 I. Amghizar, L.A. Vandewalle, K.M. Van Geem, and G.B. Marin (2017) “New Trends in Olefin Production”, Engineering 3(2):171-178.

41 S. Jenkins (March 2, 2018) “World’s Largest Single-Train Methanol-To-Olefins Plant Now Operating”, *https://www.chemengonline.com/worlds-largest-single-train-methanol-to-olefins-plant-now-operating/*.

42 Advanced Motor Fuels (2020) “Dimethyl ether (DME)”, *https://www.iea-amf.org/content/fuel_information/dme*.

43 Oberon Fuels (2020) “DME Fuel Basics”, *http://oberonfuels.com/about-dme/dme-basics/*.

44 B. Williams (December 15, 2020) “Methanol Storage and Distribution”, *https://www.briangwilliams.us/methanol-economy/methanol-storage-and-distribution.html*.

45 Wikipedia (January 2, 2021) “Methanol”, *https://en.wikipedia.org/wiki/Methanol*.

46 *everyspec.com* (February 8, 2016) “O-M-232N, Federal Specification, Methanol”, *http://everyspec.com/FED_SPECS/O/O-M-232N_53980/*.

47 International Methanol Producers & Consumers Association (December 8, 2015) “IMPCA Methanol Reference Specifications”, *http://www.methanol.org/wp-content/uploads/2016/07/IMPCA-Ref-Spec-08-December-2015.pdf*.

48 E. Fiedler (2012) “Methanol”, in Ullman’s Encyclopedia of Industrial Chemistry, Wiley-VCH Verlag GmbH & Co. KGaA, Weinheim 10.1002/14356007.a16_465.pub3, 1-27.

49 S.R. Early (2011) “Process for the Purification of Methanol by Distillation”, US Patent Application, US 2011/0214981 A1.

50 BSE Engineering (2017) “Small scale Methanol Plants: a chance for re-industrialisation”, International Methanol Conference, May 8-10, *http://www.methanol.org/wp-content/uploads/2018/04/Small-Scale-Methanol-Plants-Christian-Schweitzer_bse-Engineering.pdf*.

51 A. Pinto (1984) “Methanol Production”, US Patent 4,455,394.

52 Johnson Matthey (n.d.) “Compact Reforming Technology”, *http://www.methanol.org/wp-content/uploads/2018/04/JM-Compact-Reformer-002.pdf*.

53 ThyssenKrupp (2017) “Methanol plants”, *http://www.methanol.org/wp-content/uploads/2018/05/thyssenkrupp-Green-Methanol-small-scale.pdf*.

54 Primus Green Energy (May 2016) “Intro to Primus Green Energy”, *http://www.methanol.org/wp-content/uploads/2018/04/Primus-Intro-May-2016-Methanol.pdf*.

55 ADI Analytics (2014) “Natural Gas Utilization via Small-Scale Methanol Technologies”, presented at Natural Gas Utilization Conference, October 14, *http://www.methanol.org/wp-content/uploads/2018/04/ADIAnalytics.pdf*.

56 J. Lane (March 2, 2015) “Maverick Announces Its Oasis BG Gas-to-Liquid Methanol Plant”, Advanced BioFuels USA, *https://advancedbiofuelsusa.info/maverick-announces-its-oasis-bg-gas-to-liquid-methanol-plant/*.

57 R. Saxton (2016) Private Communication.

58 F.J. Keil (1999) “Methanol to Hydrocarbons: Process technology”, Microporous and Mesoporous Materials 29:49-66.

59 Engineering New Zealand (2020) “Motunui Synthetic Fuels Plant”, *https://www.engineeringnz.org/programmes/heritage/heritage-records/motunui-synthetic-fuels-plant/*.

60 J. Haggin (March 25, 1985) “First methanol-to-gasoline plant nears startup in New Zealand”, Chem. Eng. News. *https://www.osti.gov/biblio/6335912*.

61 P. Basu (2018) “Production of Synthetic Fuels and Chemicals from Biomass”, in Biomass Gasification, Pyrolysis and Torrefaction, (Third Edition), Academic Press, Chapter 12, 415-443.

62 D.C. Denton, B. Turk, and R. Gupta (2011) “Syngas Cleanup, Conditioning, and Utilization”, in Thermochemical Processing of Biomass: Conversion into Fuels, Chemicals, and Power, R.C. Brown, ed., John Wiley & Sons, Ltd., Chapter 4:78-123.

63 K. Cui (2017) “How are Olefins made from CTO/MTO?”, *https://www.woodmac.com/news/editorial/how-are-olefins-made-from-cto-mto/*.

64 J.G. Speight (2011) “Hydrocarbons from Synthesis Gas”, in Handbook of Industrial Hydrocarbon Processes, Chapter 8, 281-323.

65 C. Bouchy, G. Hastoy, G.E. Guillon, and J.A. Martens (2009) “Fischer-Tropsch Waxes Upgrading via Hydrocracking and Selective Hydro-isomerization”, Oil and Gas Science and Technology-Rev. IFP 64(1):91-112.

66 F.G. Botes, L.P. Dancuart, H.G. Nel, A.P. Steynberg, A.P. Vogel, B.B. Breman, J.H.M. Font Freide (2011) “Middle Distillate Fuel Production via Fischer-Tropsch Process”, in Advances in Clean Hydrocarbon Fuel Processing, Woodhead Publishing, Chapter 11, 329-362.

67 Alternative Fuels Data Center (2020) “Renewable Hydrocarbon Biofuels”, *https://afdc.energy.gov/fuels/emerging_hydrocarbon.html*.

68 Q. Li (2012) “Making on Specification Jet Fuel Through Syncrude Refining”, Princeton University: China Energy Group, *http://www.princeton.edu/puceg/perspective/SyncrudeRefining.html*.

11.0 GUIDELINES FOR THE DESIGN OF ENVIRONMENTAL TREATMENT MODULE(S)

11.1 INTRODUCTION

Previous chapters have focused on the extraction of value-added products like electricity, fuels, and/or chemicals via gasification. To accomplish this objective, impurities were removed from the syngas to enable the most efficient and effective syngas utilization possible. Chapter 6 provides a description of different syngas cleanup technologies available for the different impurities and a design strategy for the selection of the appropriate syngas cleanup technologies. The objective of syngas cleanup was impurity removal from the syngas, as discussed in Chapter 6. In addition to a clean syngas product, all the syngas cleanup technologies described also produce a byproduct stream that contains the impurities removed from the syngas stream. The impurities in these byproduct streams typically require additional treatment to ensure environmental compliance of the overall gasification plant. Chapter 12 discusses the capture, storage, and utilization of carbon dioxide (CO_2) generated during gasification and/or combustion to reduce CO_2 emissions to the atmosphere. Chapter 11 addresses the remaining environmental treatment of byproduct streams generated by the fuel conversion processes.

The primary selection criterion for these environmental treatment processes is to maintain low impurity release to the environment per the applicable federal, state, and local regulatory requirements. Secondary selection criteria include minimizing costs and optimizing the opportunities to generate secondary value-added products with the impurities rather than waste streams requiring disposal. For Radically Engineered Modular Systems (REMS) plants, the size and footprint for these environmental treatment processes also needs to be considered. As these criteria are evaluated as part of the design process, modifications of the initial syngas cleanup technology should be considered to achieve the most efficient and cost-effective overall solution.

The large variety of gasification, syngas cleanup, and syngas conversion technologies available create an almost infinite number of different byproduct streams that need potential environmental treatment. This chapter provides an overview of the potential environmental treatment options available for the larger, more commonly used treatments Project developers will need to complete more detailed analyses of environmental treatment requirements during the detailed plant design. The most convenient means of organizing the environmental treatment options is by the physical nature of the byproduct stream at ambient conditions, including solids, liquids, and gases.

11.2 ENVIRONMENTAL TREATMENT FOR SOLID BYPRODUCT STREAMS

The first major pollutant category to be discussed is solid waste streams. For gasification systems, these solids consist of inert matter present in the fuel (also known as ash) and unreacted carbon (char) fines, which are sometimes called soot. In high-temperature gasifiers, the ash melts and becomes slag. The properties and behavior of these solids and solutions for properly capturing and disposing of them are discussed in the following subsections.

11.2.1 SLAG, FINE ASH, AND CHAR

The main solid byproduct generated during the fuel conversion process is from the ash impurities. A secondary solid byproduct from the conversion process is unconverted carbon in the form of char. Based on the available separation technologies, these two solid byproducts are simultaneously removed as a mixture. The chemical and physical properties of ash impurities and the char byproduct stream depend on the gasification technology selected and fuel species. For gasifiers that operate at higher temperatures, the carbon conversion process is more efficient, reducing the amount of char generated. The higher temperatures of these gasifiers also result in melting and fusion of the ash impurities, which generates a vitrified slag. As the gasifier operating temperature drops, carbon conversion drops and vitrification decreases due to melting/fusion of the ash. Gasification also results in a range of particulate sizes. Slag is generally recognized as the larger material generated. The char and fine ash are small and often light enough to be entrained out by the syngas. The amount of slag and ash produced, as well as the individual chemical species that comprise them, are determined by the composition of the fuel used.

In vitrified slag, the ash minerals are essentially coated with a glass-like material, which prohibits leaching of these minerals out of the slag. This significantly eliminates the potential environmental issues associated with the leaching of controlled elements from the slag, making it more suitable for use or disposal as a non-hazardous solid. As the gasifier operating temperature drops and less melting/fusion of the ash occurs, the potential for leaching the captured ash minerals increases. If the low-temperature ash contains leachable toxic elements (e.g., arsenic, barium, cadmium, chromium, lead, mercury, selenium) it cannot be disposed of as non-hazardous waste.[1]

One treatment option for the captured fine ash and char is to recycle these solids back to the gasifier to improve carbon conversion, allow capture of the fine ash during melting/fusion of the slag, and assist with composition control of the ash to promote better melting/fusion. Alternative commercial uses for the slag, fine ash, and char solids include cement and some road-based products and functions. The unconverted carbon in the char provides additional fuel in the cement production process. The high temperature of the cement production process also incorporates any leachable toxic elements into the final cement product, eliminating the leachability issue. The U.S. Department of Energy's National Energy Technology Laboratory (DOE/NETL) is also currently pursuing technologies that can extract rare earth elements (REEs) and critical or valuable elements from the slag and fine ash.[2]

The final issue for project developers to consider for the environmental treatment of slag, fine ash, and char is nature of the byproduct stream generated. In Chapter 6, the slag, fine ash, and particulates were removed by a dry, wet, or possibly a combination of dry/wet processes to achieve the desired concentrations of particulates in the syngas. Although wet processes are more commercially common, the water must be separated from the solids and requires additional treatment. The dry methods eliminate the need for separation and treatment of the water, but face more technical challenges in transporting and handling the extremely fine captured particulates.

11.2.2 SPENT SORBENT, CATALYSTS, AND FILTER MEDIA

Other solids generated by the fuel conversion process include sorbents, catalysts, and filter media. For catalysts, especially those comprised of expensive noble metals, recycling is a viable option over disposal. Generally, the catalyst vendor takes back the catalyst to recover the noble metals. Most of the solid byproducts generated during the conversion and utilization process end up as solid waste. Prior to disposal, these materials need to be tested to determine the most appropriate means of disposal. For example, the activated carbon used for mercury removal must be disposed of as a hazardous waste due to the adsorbed mercury and arsenic on the sorbent. Similarly, the filter media used in low-temperature, solvent-based acid gas treatments results in the accumulation of arsenic and other toxic elements, generating a hazardous waste material.

11.2.3 SOLID BYPRODUCT FROM FUEL HANDLING AND PRETREATMENT

Depending on the quality of the fuel used, there may be a need for handling and pretreatment, which could generate a solid byproduct of rejected material. The size of this stream can be reduced and/or eliminated by selecting a high-quality feedstock as a raw material that does not require any pretreatment. However, if a solid byproduct of rejected material from the pretreatment process is generated, this material needs to be disposed of in an environmentally acceptable manner.

11.2.4 ENTRAINED SOLID BYPRODUCT

Downstream of the gasification process, the syngas cleanup system includes cleanup technologies to remove entrained particulates in the syngas to minimize the abrasion/erosion of process equipment, fouling of heat exchangers, and plugging of catalyst beds (see Section 6.3.1). Therefore, the cleaned syngas should have very low concentrations of entrained solids, but any gas phase stream released to the environment should be evaluated for the entrained solids. If the entrained particulate concentration in any of these streams exceeds the regulatory limits, particulate control should be added to capture entrained particulates and achieve regulatory compliance for the released gas stream.

11.3 ENVIRONMENTAL TREATMENT FOR LIQUID BYPRODUCT STREAMS

Liquid byproduct streams include wastewater streams (including black water from slag quenching and fly ash scrubbing), tars, and waste fuel/chemical compounds. Each of these waste streams is produced in unique ways and generally require unique solutions for proper cleanup and disposal. The various methods of cleanup for liquid byproducts are given in the following sections.

11.3.1 WASTEWATER TREATMENT

Wastewater is the largest liquid byproduct stream generated in a power or chemical production plant. The technologies implemented to process industrial wastewater to achieve regulatory compliance for release to the environment are well established and commercially demonstrated. The following discussion identifies the specific and unique features of wastewater generated in a gasification-based plant.

The use of wet scrubbing technologies for the removal of particulates from the syngas during syngas cleanup generates a byproduct stream that contains water and solids. One function of the wastewater treatment process is to separate these solids from the water. The separated solid byproduct stream can be recycled within the process, sold as a raw material for cement or road production processes, or disposed of as a solid waste (see Section 11.2.1). The separated wastewater is referred to as grey water and is frequently reused within the feed water for slurry production and syngas scrubbing with the goal of minimizing the need for freshwater.

The wet scrubbing technologies used for syngas cleaning also result in the removal of chlorine, primarily as hydrogen chloride (HCl). Chlorine is present as a dissolved solid in the wastewater. The amount of chlorine present in the wastewater is dependent on the concentration of chlorine in the fuel. For high-chlorine feedstocks, the concentration of the chlorine in the wastewater could be sufficient to warrant special treatment to reduce the chlorine concentration to comply with regulatory requirements.

For high-pressure conversion processes, any wastewater generated has a large amount of dissolved gas species due to the higher solubility of the species at high pressure. These gas species include hydrogen (H_2), carbon monoxide (CO), CO_2, hydrogen sulfide (H_2S), carbonyl sulfide (COS), ammonia (NH_3), hydrogen cyanide (HCN), and nitrogen (N_2). When the pressure is reduced, the solubility of these gas species decreases and they are released to the environment. Although the amounts are small, there is sufficient gas released, especially in confined or poorly ventilated locations, to create explosion or toxicity risks/hazards. These risks/hazards need to be evaluated when performing maintenance on the wastewater disposal/treatment system for personnel protection. A stripper is usually included for the separation of these gas species to handle the release of these gas species during depressurization of the wastewater. This byproduct gas stream is usually treated with thermal decomposition in the Claus decomposition furnace (see Section 11.4.2.1).

The large amount of grey water recycled in a wet scrubbing process limits the removal of NH_3 present in the syngas. NH_3 removal from the syngas typically occurs during the removal of water vapor during the cooling for solvent-based acid gas treatment. The combination of low temperature and large amounts of condensed process water results in significant NH_3 removal. The dissolved NH_3 in the wastewater is typically removed by steam stripping. The NH_3-rich stream from this stripping process is typically thermally decomposed in the Claus furnace (see Section 11.4.2.1), but alternative NH_3 oxidation catalysts could be used.

Based on the contact between the wastewater and the slag, fine ash, and char, the wastewater may contain dissolved solids. Special treatment may be required for any of the toxic elements (e.g., arsenic, barium, cadmium, chromium, lead, mercury, selenium) to ensure that the concentration of these elements in the treated wastewater does not exceed regulatory limits. Additional treatment may also be required to control both general dissolved salts and/or specific salts that result from impurities in the fuel.

11.3.2 TARS

Lower temperature gasification processes are less efficient at carbon conversion and generate a significant amount of condensable hydrocarbons referred to as tars. As these tars can cause significant fouling of downstream heat exchangers and/or catalyst beds, they are typically removed from the syngas stream during syngas cleaning (see Chapter 6). The typical approach to dealing with these tars is to recycle them to the gasifier to improve the overall carbon conversion of the plant. Intermediate processing of these tars could also be used to extract value-added byproducts prior to the recycling of the low-value hydrocarbons back to the gasifier.

11.3.3 WASTE FUEL AND CHEMICAL BYPRODUCT STREAMS

Although the final fuel or chemical production processes are optimized to minimize the generation of waste fuel and chemical byproduct streams, these production processes generate some waste fuel or chemical byproduct streams. These byproduct streams have fuel value, which allows them to be used to generate heat or power. Alternatively, these byproducts streams can be recycled back to the gasifier for conversion into syngas.

11.4 ENVIRONMENTAL TREATMENT FOR GAS PHASE BYPRODUCT STREAMS

Gaseous pollutants include nitrogen-, sulfur-, and carbon-based compounds, as well as mercury and some trace elements, that are ordinarily solid or even liquid at room temperature, but exist in the vapor state due to the high temperatures present in the gasifier and early on in the gas cleanup system. Each of these pollutants and the various options for controlling their release to the environment is discussed in the following subsections.

11.4.1 NITROGEN OXIDE CONTROL FOR FLUE GAS FOR POWER GENERATION

When syngas is used as part of an integrated gasification combined cycle (IGCC) for electricity production, the syngas is combusted to generate electricity and a flue gas. This combustion process generates nitrogen oxide (NO_X) via three reaction pathways. One of these reaction pathways is through oxidation of any NH_3 and HCN innately present in the syngas (i.e., fuel NO_X), which are generated from the nitrogen present in the fuel. A second, related pathway is prompt NO_X, where nitrogen in the air used for combustion bonds with smaller hydrocarbons (e.g., methane) to form amine and cyanide groups, which then combust to form NO_X as in the fuel NO_X mechanism. The third reaction pathway is through thermal NO_X, where direct oxidation of the nitrogen in the air occurs due to high temperature. In these IGCC-based plants, the syngas cleanup system uses catalytic processes to convert the HCN into NH_3 and water wash processes to remove the NH_3 and HCN from the syngas (see Chapter 6). By effectively using syngas to control the nitrogen species in the clean syngas, the NO_X present in the flue gas from the combustion turbine is predominantly from thermal NO_X. This thermal NO_X production can be limited by using low NO_X burner technologies and using CO_2, steam, and/or nitrogen diluents in the syngas feed to the turbine to assist in flame temperature control. Through application of these different technologies, NO_X emissions below 15 parts per million by volume (ppmv) are possible.[3]

If additional NO_X reduction is required to meet regulatory limits, additional treatment of the flue gas is required. The standard commercial NO_X control technology used by most of the combustion processes is selective catalytic reduction (SCR). In SCR, controlled amounts of reductant, consisting of anhydrous NH_3, aqueous NH_3, or urea, are added to the flue gas before it passes over a catalyst bed. The catalyst consists of a ceramic support with vanadium, molybdenum, or tungsten oxide, zeolite, and various precious metals. Operating temperature is a critical operating parameter, with an optimal operating temperature range of 357°C/674°F to 447°C/836°F. Catalyst life can be extended by operating in the 227°C/441°F to 447°C/837°Frange, but catalytic activity may not be sufficient at the lower temperatures to meet NO_X destruction targets.

For NH_3 as a reductant, the key products are nitrogen and water. When urea is used as a reductant, it must first be converted to NH_3 and CO_2 by thermal decomposition before the NH_3 can reduce the NO_X, with the resulting products being nitrogen, water, and CO_2.

Other potential poisons include alkali metals, alkaline earth metals, arsenic, phosphorus, antimony, chromium, lead, mercury, and copper, which can deactivate the NO_X reduction catalyst and result in oxidation of the NH_3 reactant, increasing NO_X concentrations. Particulate matter entrained in the flue gas can also result in pore blocking in the catalyst, which prohibits adsorption of the NH_3 on the catalyst surface, resulting in loss of catalytic activity. This is not typically a problem for SCRs used for power generation from syngas, because of the low particulate tolerance of the gas turbines.

SCR processes are effectively tuned to ensure optimal addition and distribution of the NH_3 reductants to optimize the NO_X reduction and minimize the amount of NH_3 passing through the system unreacted (referred to as ammonia slip). SCR tuning also frequently develops the specific amount of NH_3 necessary for optimal NO_X reduction across the anticipated operating range for the process to support stable, reliable performance.

Although not common in practice, a lower cost alternative to SCR is selective non-catalytic reduction (SNCR), which operates in a manner similar to SCR, but without the need for any catalysts. SNCR requires much more reagent to be effective and must be utilized much closer to the combustion zone due to the lack of catalyst. If minimal NO_X reduction is required, SNCR can reliably achieve up to around 60% NO_X reduction at up to one-tenth of the capital costs of SCR. A unique application of SNCR is the use of a rich reagent injection (RRI) scheme, where the aqueous NH_3 or urea is directly injected into the combustion zone with the fuel. This allows for greater residence time of the reagent, allowing for higher NO_X reduction (though at the risk of higher reagent slip rates).[4] However, SNCR with or without RRI is unsuitable for gas turbine systems, as baseline NO_X concentrations are generally too low for selective reduction techniques to be effective without a catalyst. Rankine cycle boilers and reciprocating engines are appropriate for either technology (depending on the need). A comparison of SCR and SNCR is included in Table 11-1.

Table 11-1. Comparison of selective catalytic reduction and selective non-catalytic reduction[4]

CONTROL TECHNOLOGY	APPLICATION	NO_X REDUCTION (%)	CAPITAL COST	
			($/TON)	($/KW)
SCNR	Boilers/Process Heaters	40 to 60	500 to 1,000	10 to 20
SCR	Boilers/Process Heaters	80 to 90	1,000 to 10,000	20 to 200

11.4.2 SULFUR RECOVERY PROCESSES

During the syngas cleanup process, most of the sulfur present as H_2S, COS, and carbon disulfide (CS_2) is removed from the syngas to support the optimal operation and environmental performance of downstream syngas conversion processes. The impurity byproduct stream generated by most sulfur removal processes consists of an H_2S-rich gas stream. Due to its concentration and toxic nature, this byproduct stream must be treated to recover more than 99.8% of this sulfur to meet environmental regulatory requirements. The standard commercial processes for treating this H_2S-rich byproduct stream focus on conversion of the sulfur into either elemental sulfur or sulfuric acid. The elemental sulfur product is preferred, as it is easily stored and can be conveniently and inexpensively transported over large distances. The sulfuric acid product is preferred in areas where local demand for sulfuric acid is high due to local phosphate mining or significant NH_3-based fertilizer production. The two standard commercial processes to convert this H_2S-rich byproduct stream into elemental sulfur include a Claus process with tail gas treatment (TGT) and wet oxidation or redox processes. The wet oxidation or redox processes are described in Chapter 6 and produce a low-quality, low-value sulfur without additional processing.

11.4.2.1 CLAUS PROCESS

The Claus process is based on the Claus reaction ($2H_2S + SO_2 \rightarrow 3/x\ S_X + 2H_2O$), which is equilibrium limited. Under typical operating conditions, the sulfur is present as a gas, with the primary molecular species being S_2, S_6, and S_8. The gas temperature dictates the relative abundance of each species with mostly S_2 at higher temperatures and mostly S_8 at lower temperatures. The reaction is catalyzed by aluminum oxide (alumina).

The general block flow diagram for a standard commercial Claus system is shown in Figure 11-1. The first step of the process is to oxidize approximately one-third of the H_2S into sulfur dioxide (SO_2) with air to meet the required stoichiometry of the Claus reaction. Heat from these highly exothermic oxidation reactions is recovered from the product gas to generate high-pressure steam in a waste heat boiler. Any sulfur produced in the furnace is recovered in a sulfur condenser, which is used to generate low-pressure steam. The resulting gas is then passed through two or three catalytic conversion stages. Each of these catalytic stages consists of a reheater, catalytic reactor, and sulfur condenser. The lower operating temperature in the catalytic reactor and sulfur condenser are used to effectively maximize sulfur production for the equilibrium-limited Claus reaction. The liquid elemental sulfur product is collected from the sulfur condensers into a sulfur pit prior to transport for use or disposal.

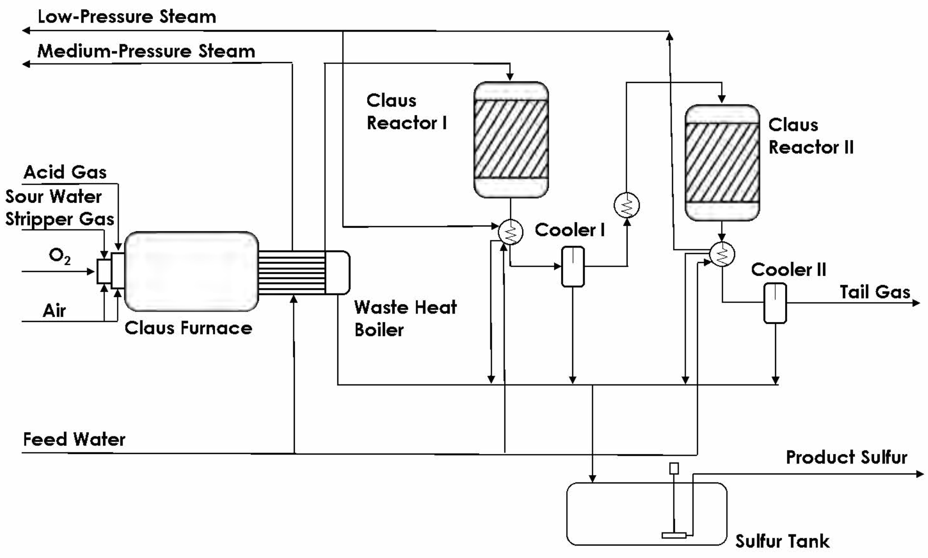


Figure 11-1. Block flow diagram for a typical Claus system

The first catalytic stage is also operated at a slightly higher temperature than the remaining catalytic stages to ensure adequate hydrolysis of COS and CS_2 into H_2S and reduce the concentration of COS and CS_2 in the final tail gas.

The oxidation of H_2S in the Claus furnace at high temperatures provides the opportunity for thermal destruction of byproduct gas containing NH_3, HCN, H_2S, and other trace syngas components generated from wastewater treatment during depressurization and steam stripping. Destroying NH_3—the most difficult species to thermally destroy—requires stoichiometric oxygen concentrations, flame temperature of more than 2,280°F (1,250°C) and a residence time greater than 0.8 seconds. For extremely high concentrations of nitrogen-based species or in the event of incomplete thermal destruction of these species, ammonia salts can be produced, which can result in plugging of the downstream catalytic reactors.

One of the key performance criteria for the Claus system is the recovery of elemental sulfur. One of the key operating issues affecting this sulfur recovery is maintaining a stable and suitable temperature in the H_2S oxidation furnace. The operating temperature of the H_2S oxidation furnace is predominantly set by the concentration of the H_2S in the sulfur byproduct stream. The impacts of H_2S concentration in the sulfur byproduct stream are shown in Table 11-2.

Table 11-2. Effects on sulfur recovery based on H_2S concentration in the sulfur byproduct stream[5]

H_2S CONCENTRATION (MOLE %)	SULFUR RECOVERY (%)	COMMENT
>70	98	N/A
50 to 70	96	N/A
20 to 50	94 to 96	To achieve these sulfur recoveries, a portion of the feed gas is bypassed around the H_2S oxidation furnace to the first catalytic stage. The bypass helps stabilize the temperature of the furnace.
<15	N/A	Additional process modifications (e.g., air preheating and/or oxygen enrichment) are required to maintain stable operation of the H_2S oxidation furnace. Lower sulfur recoveries and higher cost are associated with these conditions.

Many modifications of the standard commercial Claus system have helped improve sulfur recovery and dilute H_2S concentrations in the sulfur byproduct streams. Information about these systems and their modifications is provided in Table 11-3. The primary modification has been switching from air as an oxidant to oxygen. The primary challenge for this transition is limiting and controlling the temperature in the H_2S oxidation furnace.

Table 11-3. Different modifications of the standard Claus system[5]

NAME	OXYGEN BLOWN	MODIFICATION	COMMENT
Claus Oxygen-Based Process Expansion (COPE)	Yes	Control of the temperature in the H_2S oxidation furnace is provided by the recycle of cooled effluent gas from the first sulfur condenser. This modified process also uses specially designed burners with separate feeding of air, oxygen, recycle gas, and fuel gas.	With these modifications, sulfur recoveries can be increased by 0.5 to 1.0% over conventional air-blow Claus systems, depending on the H_2S concentration in the sulfur byproduct gas.
Oxy-Claus	Yes	The furnace temperature is controlled by a proprietary burner that maintains temperature below furnace refractory limits even with pure oxygen.	This system boosts the capacity of existing Claus and tail gas units by adding air and oxygen for more efficient combustion. This increases operational flexibility—temporarily high or low amounts of H_2S are efficiently managed by an automatic change over from air to oxygen operation and vice versa, without the need for additional equipment. Capable of processing feed gases with H_2S concentrations as low as 20%. [6]
Sulfur Recovery (SURE)	Yes	This modification uses staged combustion. All the sulfur byproduct gas is fed to the first stage with a portion of the oxygen and air. A proprietary burner is used in the first stage. After the first stage, the effluent gas is cooled and fed into the second stage with the remaining oxygen. Since the temperature of the gas feed to the second stage is maintained above the auto-ignition temperature, no burner is needed for the second stage.	A special waste heat boiler design has been developed by Siirtec Nigi and Parsons, which combines the waste heat boiler and second stage into a single piece of equipment.
SuperClaus	No	For this process, the last catalytic stage is replaced with a selective oxidation based on a silica/alumina catalyst. The selective oxidation reaction ($H_2S + \frac{1}{2} O_2 \rightarrow S + H_2O$) is not equilibrium limited like the Claus reaction.	The selective oxidation reaction is not equilibrium limited, especially with high steam concentrations and does not have the tendency to form undesirable species like SO_2, COS, CS_2, and CO.
Extended Cold-Bed Claus	No	The final catalytic stage is operated below the sulfur dewpoint to permit higher conversion. In this process, the product sulfur is adsorbed on the catalyst. A regeneration step is needed to recover the sulfur.	The process requires multiple sub-dewpoint reactors to allow for regeneration of the sulfur-laden catalyst and permit continuous operation.

Although the modified Claus processes can achieve sulfur recoveries high enough to meet regulatory limits in some cases, the effluent sulfur concentrations for most of the advanced Claus systems described in Table 11-3 are too high to permit environmental release. For gasification plants, one option for treating this tail gas includes hydrogenation of the effluent stream to reduce all SO_2 to H_2S and recycle the stream back into the acid gas removal (AGR) system. Alternatively, the effluent from the Claus system needs to be treated in a TGT system to meet regulatory limits for environmental release. Optimizing the sulfur recovery in the Claus system helps to lower the overall cost of the sulfur recovery system by reducing the capacity and size of the required TGT system.

11.4.2.2 TAIL GAS TREATMENT

The sulfur species present in small and varying concentrations in the effluent tail gas from the Claus system are H_2S, COS, CS_2, SO_2, and elemental sulfur vapor. The Claus system tail gas may also contain H_2, CO, H_2O, nitrogen, and CO_2. To meet regulatory sulfur limits of more than 99.9% reduction, the remaining sulfur in the tail gas must be captured. The first step in a TGT system is to convert all the sulfur species present into H_2S via catalytic hydrogenation/hydrolysis. The elemental sulfur and SO_2 are hydrogenated. The hydrogenation reactions result in almost complete conversion of the SO_2 and elemental sulfur when there is excess hydrogen and the concentration of the sulfur species is less than 10 ppmv. The hydrogen for these hydrogenation reactions is provided by hydrogen already present in the tail gas, by partial oxidation of the fuel gas used to reheat the tail gas, or from an external source. The COS and CS_2 are converted into H_2S via hydrolysis reactions. After hydrolysis, the CS_2 concentration is reduced to less than 1 ppmv and the COS concentration to less than 10 ppmv. The water vapor needed for the hydrolysis rection is usually present in the Claus system tail gas.

After the hydrogenation/hydrolysis reactions convert the various sulfur species into H_2S, the treated gas is cooled and the H_2S is removed in a conventional amine solvent unit. The sulfur loading in the solvent is low, because of the low concentration of sulfur in the Claus system tail gas, which allows a simplified regeneration of the amine solvent with a stripper. The treated tail gas is finally incinerated before release to the environment with sulfur concentrations meeting the regulatory limits. The H_2S-rich product gas from the stripper is recycled back to the H_2S oxidation furnace in the Claus system.

Various vendors offer Claus TGT systems based on their proprietary solvents and catalytic reduction reactors. The Shell Claus Off-Gas Treating (SCOT) process is the most widely used commercial TGT system (Figure 11-2).

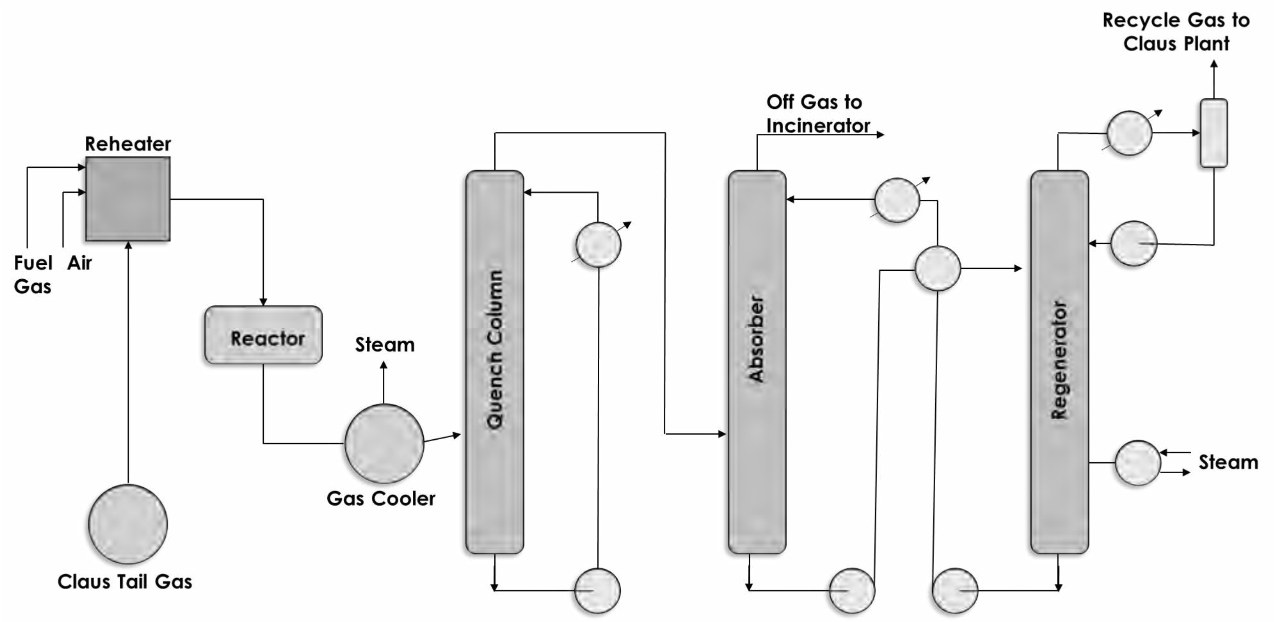


Figure 11-2. Shell Claus off-gas treating (SCOT) process

The capture of the H_2S from the Claus tail gas after hydrogenation/hydrolysis offers significant potential for process intensification with the existing desulfurization processes in the syngas cleanup system. With the extremely lean sulfur loading added by removing sulfur from the treated Claus system tail gas, this absorption unit can be combined with the amine desulfurization system used for the syngas, eliminating the need for a separate stripper in the TGT system. Alternatively, the treated tail gas stream can be compressed and recycled back into the syngas stream upstream of the desulfurization process. This option works best when the volumetric flow of the treated tail gas is low enough to not significantly dilute the syngas stream. The project developer should evaluate these opportunities for process intensification as part of the design and cost estimation processes.

11.4.2.3 SULFURIC ACID PRODUCTION

When local economic conditions provide a strong, stable market for sulfuric acid, commercial process technology is available to convert the sulfur byproduct from the syngas cleanup system into sulfuric acid. The sulfur in the sulfur byproduct stream from the syngas cleanup process is first oxidized to generate a stream of SO_2. This high-temperature process can be used for decomposition of the NH_3 and HCN byproduct stream from the wastewater treatment like the H_2S oxidation in the Claus system. Heat is recovered from the hot stream of the thermal oxidation unit in a waste heat boiler to generate steam. If the water is removed from the cooled SO_2 stream, the conventional commercial sulfuric acid production process can be used. If the water is not removed, Haldor Topsøe offers their commercial wet sulfuric acid (WSA) process.[7] In both processes, the SO_2 is oxidized to sulfur trioxide (SO_3) using a multi-bed catalytic process with interstage cooling to maximize the equilibrium conversion of SO_2 into SO_3. The SO_3 is then absorbed into water in a controlled process to generate sulfuric acid. The tail gas from the sulfuric acid process with sulfur concentrations meeting regulatory requirements is released to the environment.

Previous sections focused on sulfur byproducts from the syngas cleanup when the predominate sulfur species is H_2S. Some desulfurization technologies produce a sulfur byproduct stream with SO_2 rather than H_2S. For these sulfur byproduct streams, the Claus and sulfuric acid production technologies can be readily adapted to effectively capture the sulfur and generate a treated gas stream and meet regulatory requirements. For sulfuric acid production, the initial oxidation process to generate the SO_2 is not required prior to the oxidation of SO_2 into SO_3. The first step in TGT is the catalytic hydrogenation/hydrolysis of all sulfur species, including SO_2 into H_2S. This same catalytic process can be used to convert around two-thirds of the SO_2 into H_2S, which supports the stoichiometry of the Claus reaction: $2H_2S + SO_2 \rightarrow 3/x\ S_X + 2H_2O$. For either of these approaches, the high-temperature oxidation process is eliminated, prohibiting the use of this process for the decomposition of the NH_3 and HCN byproducts generated from wastewater treatment. The byproduct streams with these nitrogen species can be effectively treated by incineration, but the overall cost, equipment, and footprint effects of the desulfurization technology and sulfur byproduct treatment need to be considered by the project developer during the design phase.

11.4.3 CARBON DIOXIDE FOR ENHANCED OIL RECOVERY OR STORAGE

There are currently no regulatory requirements for the release of CO_2 to the environment in the United States. However, there is a general agreement that CO_2 emissions regulations may be enacted to deal with the contribution of greenhouse gases. These emissions regulations could have a serious impact on the design, operation, and cost of a REMS plant, so project developers need to include evaluation and analysis of the impact of potential CO_2 emissions regulations (see Chapter 12). The two key factors associated with CO_2 capture in a REMS plant are the amount of CO_2 that is released to the environment by the conversion plant and ensuring the specifications of CO_2 byproduct stream meet the ultimate intended use of utilization and/or storage. Without the regulation(s), there is no CO_2 value or incentive for either of these factors. The general approach used for CO_2 emissions has been to set a target for CO_2 captured by the plant. NETL has based their research platforms on 90% CO_2 capture or more based on the carbon introduced into the process through the fuel and air. NETL's general approach to the CO_2 byproduct stream has been based on a general specification for CO_2 for saline reservoir storage provided in the Quality Guidelines for Energy System Studies (QGESS) CO_2 Impurities Design Parameters document; the findings of which are summarized in Table 11-4.[8]

Table 11-4. QGESS CO_2 impurities specification for saline reservoir storage[8]

COMPOSITIONS		QGESS VALUE	LITERATURE VALUES
CO_2	vol % (min)	95	90 to 98
N_2	vol % (max)	4*	0.01 to 7
Ar	vol % (max)		0.01 to 4
O_2	vol % (max)		0.01 to 4
H_2	vol % (max)		0.01 to 4
CH_4	vol % (max)		0.01 to 4
H_2O	ppmv (max)	300	20 to 650
H_2S	vol % (max)	0.01	0.002 to 1.3
NH_3	ppmv (max)	50	0 to 50
SO_2	ppmv (max)	100	10 to 5,000
NO_X	ppmv (max)	100	20 to 2,500
CO	ppmv (max)	35	10 to 5,000
Delivery Pressure	psia	2215	–

**This 4 vol% represents the total maximum combined concentration of N_2, Ar, O_2, H_2, and CH_4.*

Details on the technologies and design configurations needed to achieve the target specifications for CO_2 capture and the CO_2 byproduct specification are discussed in Chapter 12, which also describes a general approach for evaluation of the importance/need to include CO_2 capture in the design of a REMS plant.

11.4.4 MERCURY AND OTHER TRACE CONTAMINANTS

Mercury and the trace impurities (e.g., arsenic, cadmium, lead, nickel, chromium) present in fuels represent potential process and environmental hazards. As process hazards, these species can poison catalysts; foul heat exchangers; and result in material erosion, corrosion, and embrittlement in process equipment. A summary of the environmental regulatory standards and accepted concentrations to avoid process hazards for many of these species, including trace contaminants, present in the fuel and/or generated during gasification is available in Table 6-1 (see Chapter 6). Most of these contaminants are treated as part of the syngas cleanup process due to the following:

- The separation of contaminants from the syngas treats the smallest volume of process gas with the highest concentration of contaminant, reducing the size and increasing the efficiency of the separation process.
- Multiple major contaminant removal processes result in the simultaneous removal of trace contaminants.
- Accepted concentrations that avoid process hazards require some, if not more, contaminant removal than that required to meet environmental limits.

The separation technologies for many trace contaminants and the strategy for their integration into the syngas cleanup process is discussed in Chapter 6.

The challenge is addressing the impurity byproducts generated by the separation processes during syngas cleanup. The large variety of trace contaminants present means that impurity byproducts with trace contaminants can be solids, liquids, or gases. This chapter has mentioned issues with solid impurity byproducts like slag, fine ash, char, catalysts, and filter media that may need special treatment and disposal due to the leachability of toxic elements. Similarly, special wastewater treatment may be needed to deal with dissolved trace contaminant species. Finally, vapor phase trace contaminant species like mercury and arsenic need to be evaluated when considering the release of byproduct gases to the environment.

The general goal of acceptable disposal, processing, or release of the trace contaminant byproduct is to minimize the overall treatment required. This begins with the selection of the syngas cleanup system to ensure not only effective separation of the trace contaminant from the syngas, but also the generation of an impurity byproduct stream that can be effectively disposed of or released to the environment with little or no additional treatment. This selection process involves an iterative approach to ensure this objective can be achieved with all trace contaminant species.

11.5 SUMMARY

There are two specific objectives for the contaminant separation technologies used in the syngas cleanup system. The first is to ensure optimal performance and efficiency of the syngas utilization process for producing electricity, fuels, and/or chemicals with minimal adverse impact on process equipment. The second objective is to generate impurity byproduct streams that require little or no additional treatment prior to acceptable disposal or release to the environment to meet regulatory requirements. Chapter 11 illustrated the key areas related to acceptable disposal or release of the impurity byproduct streams generated by a REMS plant.

The two largest impurity byproducts that require additional treatment are wastewater and sulfur byproduct. In general, conventional commercial technologies are available to treat all the special impurities introduced into wastewater, resulting in a tradeoff between simplicity of syngas cleanup versus a more complex wastewater treatment. For the sulfur byproduct, the toxic nature of the byproduct sulfur streams (H_2S or SO_2) prompts the conversion of these species into either elemental sulfur or sulfuric acid (as marketable products). The high-temperature conversion used in the treatment of the sulfur byproduct is also used to acceptably dispose of NH_3 and HCN generated from wastewater treatment.

Although electric power production from syngas in an IGCC plant tends to have very low NO_X concentrations in the flue gas, many of the commercially planned IGCC projects have included an SCR system to meet legal NO_X emissions requirements. The need for SCR or SNCR treatment of the flue gas for a REMS gasification-based system needs to be adequately evaluated by the project developer, while considering the cost of this additional technology versus the cost of failing to meet regulatory requirements. In the absence of CO_2 emission regulations, there is no current need for CO_2 capture and disposal or utilization of the CO_2 byproduct stream (see Chapter 12).

Finally, Chapter 11 shows the need to include an evaluation of the treatment of the impurity byproduct streams as a criterion during the design of a selection process for contaminant cleanup technologies. This ensures that the overall cost associated with both the removal of the impurity from the syngas and the ultimate disposal or release of the impurity to the environment result in the simplest and most cost-effective overall process possible. For REMS plants, the size and footprint of these environmental treatment processes also needs to be considered. When these additional criteria are added to the iterative design and selection criteria, project developers can select an optimal syngas cleanup system and the most minimal impurity byproduct treatment to acceptably meet regulatory requirements for the release or disposal of all impurity byproduct streams.

Endnotes

1 NETL (2020) “Gasifipedia. 4.3. Chemical and Leachate Characterization of Gasifier Slag”, DOE/NETL, *https://netl.doe.gov/research/Coal/energy-systems/gasification/gasifipedia/chemical-leachate*. Accessed July 15, 2020.

2 NETL (n.d.) “Feasibility of Recovering Rare Earth Elements”, DOE/NETL, *https://netl.doe.gov/coal/rare-earth-elements*. Accessed September 16, 2020.

3 R. Jones and N.Z. Shilling (2002) “IGCC Gas Turbines for Refinery Applications”, presented at Gasification Technologies Conference, San Francisco.

4 R.D. Bell and F.P. Buckingham (n.d.) “An Overview of Technologies for Reduction of Oxides of Nitrogen from Combustion Furnaces”, MPR Associates, Inc., Alexandria, VA, USA.

5 N. Koren, D. Simbek, and D. Wilhelm, “Process Screening Analysis of Alternative Gas Treating and Sulfur Removal for Gasification”, Revised Final Report, SFA Pacific Inc., 2002.

6 Air Liquide (2014) “Sulfur Management Solutions”, *http://www.chemwinfo.com/private_folder/Uploadfile2014December/AIR_Sulfur_Management.pdf*.

7 C. Higman and M. van der Burgt (2008) “Gasification”, Second edition, Elsevier, pp. 364-365.

8 NETL (January 2012) “CO_2 Impurity Design Parameters”, DOE/NETL, 341/011212.

12.0 GUIDELINES FOR THE DESIGN OF A CARBON DIOXIDE CAPTURE, UTILIZATION, AND STORAGE MODULE

12.1 INTRODUCTION

Carbon capture, utilization, and storage (CCUS) strategies are crucial to reducing carbon dioxide (CO_2) emissions across the globe. The four primary categories of CO_2 capture methods are absorption (solvents), adsorption (sorbents), membranes, and cryogenic processes. The technical and economic challenges are (1) minimizing the energy intensity of CO_2 capture due to the thermodynamic penalties associated with the entropy of de-mixing CO_2 from the process gas and (2) the compression work required to bring the captured CO_2 to the required end-use pressure. The technical and economic success of a technology for overcoming the challenges for a specific application heavily depends on specific operating conditions, which include feed gas CO_2 partial pressure, other impurities in the process gas, flow rate, target capture rate, and target CO_2 purity.

Once the CO_2 is captured, a suitable end use must be identified that prevents the release of this CO_2 to the environment. One of the challenges with implementing CO_2 capture is finding suitable end uses that can handle the enormous amounts of CO_2 produced. Enhanced oil recovery (EOR)—where CO_2 is pumped into an oil well to raise the well pressure—increases the amount of oil that can be harvested and is the only current large-scale user of CO_2, but even this use is small in scale when considering the quantities of CO_2 that must be captured Once the oil has been extracted, these wells can be capped to store the CO_2, Other potential small-scale end-uses of CO_2 include carbonated beverages, as a refrigerant, as a flame retardant (liquid CO_2 is commonly found in most fire extinguishers), as a pneumatic gas in pressure-based tools and mechanisms, and many other niche applications. Most CO_2 producers source their product from other chemical processes or even natural CO_2 sources that produce nearly pure CO_2 and require little treatment. There is considerable research and development (R&D) in progress to expand the potential use of CO_2 as a component in the production of commodity chemicals and building materials. When a suitable end use cannot be identified, the only remaining option is to store (or sequester) the CO_2 underground inside a suitable geologic formation. Each end use of the captured CO_2 has distinct specifications for the purity, temperature, and pressure of the captured CO_2. These specifications increase the challenges associated with CO_2 capture from process gas streams and impact the selection of CO_2 capture technology.

Consequently, any planned industrial plant needs to evaluate the impacts of including or adding CO_2 capture systems. These impacts include capital and operating costs of the CO_2 capture process, technical and physical (i.e., space and plant layout) integration issues of the CO_2 capture process into the overall plant, potential end uses of the captured CO_2 and/or government incentives that could generate a revenue stream to offset the costs associated with CO_2 capture, and cost and requirements for storing the captured CO_2. This chapter helps prospective project developers understand the technical and economic impacts of including a CO_2 capture system in their planned facilities.

12.2 CARBON DIOXIDE CAPTURE TECHNOLOGY SELECTION

For power generation, the project developer should consider the three primary process approaches for CO_2 capture: pre-combustion, oxy-fuel combustion, and post-combustion capture. A summary of the technical features of each approach, along with advantages and disadvantages is provided in Table 12-1. An analysis of the advantages and disadvantages of these approaches shows that for pressurized gasification (using high-purity oxygen as an oxidant) pre-combustion CO_2 capture can potentially be part of a cost-competitive power process.

Table 12-1. Summary of three primary approaches for CO_2 capture for power generation

APPROACH	TECHNICAL DESCRIPTION	ADVANTAGES	DISADVANTAGES	SUITABILITY FOR GASIFICATION
Pre-Combustion	CO_2 or carbon is removed prior to utilization.	• High-pressure environment leads to reduced equipment sizes. • Oxygen-blown systems yield higher CO_2 concentration in syngas, making capture more energy efficient and enable chemical and fuel co-production.	• CO-shift process reduces syngas volumetric heating value and impacts turbine design. • Post-capture CO_2 product usually requires pressurization.	Good to excellent.
Oxy-Fuel Combustion	Fuel is combusted with pure oxygen.	• With minimal nitrogen (N_2) in flue gas, purifying the CO_2 consists of removing moisture, sulfur oxide (SO_X), NO_X, and ash. • When CO_2 is used as the working fluid, an Allam cycle can be used, which has demonstrated 60% efficiency (lower heating value [LHV]) with natural gas as a fuel.	• A source of oxygen is needed. • Oxygen purification costs must be balanced with resulting downstream CO_2 purification costs. • Traditional air separation units (ASUs) have high auxiliary loads.	Extremely poor due to competing chemistries of gasification and combustion.
Post-Combustion	CO_2 is removed after combustion.	• Any existing plant that requires capture can be retrofitted with CO_2 capture technology.	• Lower gas pressure and CO_2 concentrations require more complicated, expensive, and energy-intensive separation technologies compared to pre-combustion.	Acceptable, but pre-combustion is recommended due to economics.

The greatest potential for cost-competitive CO_2 capture with gasification for both power generation and chemical production is pre-combustion CO_2 capture using high-pressure, oxygen-blown gasification processes. The remainder of this chapter focuses on pre-combustion CO_2 capture from high-pressure, oxygen-blown gasification processes.

Block flow diagrams of generic power and chemical production plants based on high-pressure, oxygen-blown gasification with CO_2 capture are shown in Figure 12-1. The process flow diagrams shown in Figure 12-1 contain the key processes necessary to enable efficient CO_2 capture for power or chemical production. In the current regulatory environment (i.e., CO_2 emission allowed), only the process blocks within the blue-dashed boxes are required. This demonstrates that CO_2 capture is essentially a requirement for chemical production using gasification with or without CO_2 emission regulations. However, for power production, multiple additional processes, like water-gas shift (WGS) and CO_2 capture, must be included to achieve the prescribed CO_2 emission levels.

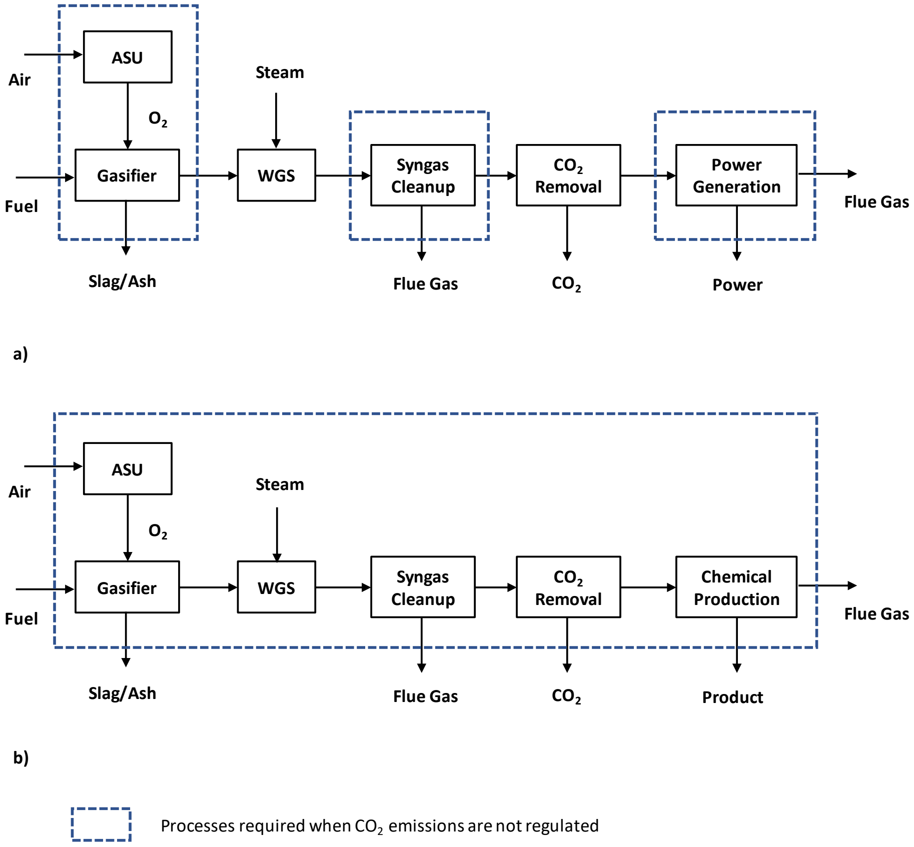


Figure 12-1. Block flow diagrams with CO_2 capture (a) power generation (b) chemical production

One selection criterion is the amount of CO_2 that needs to be captured. For chemical processes, the amount of CO_2 to be captured is determined by the stoichiometry of the chemical transformation reactions. For methanol production, the main reaction is $CO + 2H_2 \rightarrow CH_3OH$. This requires that the syngas mixture have a molar hydrogen (H_2) to carbon monoxide (CO) ratio of at least 2.0. To make the raw syngas more suitable for chemical production, the WGS reaction is used to convert CO and water (H_2O) into H_2 and CO_2 to produce a syngas with a molar H_2-to-CO ratio necessary for the chemistry of the chemical product (i.e., 2 for methanol). Although some CO_2 is beneficial to the methanol reaction, most of the CO_2 in the syngas used for chemical production is removed from the syngas prior to the chemical conversion process. This shifting of the syngas composition enables maximum production of the product compounds and ensures maximum efficiency of the syngas conversion. This fixes the amount of carbon in the original fuel that is converted to CO_2 and not the desired product compounds and consequently the amount of CO_2 to be captured.

For power production, progressively higher levels require additional and/or more sophisticated equipment and process complexity, which, in turn, increases costs and reduces overall plant efficiency. The WGS process converts CO in the raw syngas into CO_2 (Figure 12-1), which can be captured. As the CO is converted into CO_2, the H_2 concentration in the syngas increases, which impacts combustion properties and specifications for the gas turbine. Commercial turbines are currently used by defining the maximum hydrogen concentration in the turbine feed. This limit reflects the practical high level for CO_2 capture. National Energy Technology Laboratory (NETL) system analysis has shown that 90% CO_2 capture is technically achievable for currently available commercial gas turbines.[1]

Beyond simple technical challenges, the project developer should also consider the economic impact of CO_2 capture on power production and the technologies developed to enable low carbon emissions discussed in this chapter. The conversion of CO in the raw syngas using the WGS reaction reduces the energy that can be extracted from the syngas as power. Part of this reduction in power is related to the lower energy of the fuel molecules (the heating value of CO is -283.0 kilojoules per mole [kJ/mol], whereas the heating value of H_2 is -241.8 kJ/mol). Another part of the power loss is associated with the energy required to generate high-pressure steam for the WGS reaction. The WGS reaction ($CO + H_2O \leftrightarrow CO_2 + H_2$)—with the amount of forward and reverse reaction limited by thermodynamic equilibrium—requires using high steam concentrations and multiple reaction steps to maximize the conversion of CO. In addition, WGS catalysts experience less deactivation due to coking at steam-to-CO ratios of more than 2. Even with extensive integration and optimization of heat and steam generation, the generation of this required steam significantly reduces the available energy that would otherwise be used to generate power in the non-capture plant. The net result is lower overall plant energy efficiencies for plants with CO_2 capture. Finally, CO_2 provides nitrogen oxide (NO_X) control during combustion. Removing CO_2 from the syngas, coupled with hydrogen's high flame temperature requires additional steam or nitrogen (N_2) to control combustion temperatures and keep NO_X emissions low. If the need for low CO_2 emissions is greater than the economic or efficiency downsides to CO_2 capture outlined above, or a carbon tax is implemented that raises costs above those that CO_2 capture would otherwise impose, potential designers/developers should carefully weigh these costs against each other before choosing a CO_2 capture scheme. In the case that CO_2 capture is used, the cleanup system with capture should be heavily optimized to mitigate the above-mentioned issues as much as possible.

In the following sections, design selection and constraints for the WGS and CO_2 capture processes are discussed.

12.2.1 WATER-GAS SHIFT PROCESS

The WGS reaction ($CO + H_2O \leftrightarrow CO_2 + H_2$) is used to convert CO and H_2O in the raw syngas into H_2 and CO_2. For chemical production, this conversion is to achieve the necessary H_2-to-CO ratio to support the chemical transformation of the H_2 and CO in the syngas into the targeted chemical products (discussed in Chapter 6).

For chemical production, the CO_2 capture requirements are based on the H_2-to-CO ratio necessary for chemical production. As the carbon intensity of the chemical product decreases (i.e., chemical products with higher H_2-to-CO ratios), the capture requirements increase. Hydrogen production requires essentially 100% CO_2 capture, meaning a complete shifting of the CO into H_2 to achieve this high level of CO_2 capture.

For power production, the amount of CO conversion is determined by the desired level of CO_2 capture and the efficiency of the CO_2 removal process. In NETL system analysis of gasification with CO_2 capture, the key CO_2 capture technologies, Selexol™ and Rectisol®, can achieve more than 95% efficiency in CO_2 capture. To achieve 90% CO_2 capture for the plant, the CO conversion in the NETL system analysis studies was more than 97%.[1]

For hydrogen production and power production with more than or equal to 90% CO_2 capture, essentially all the carbon in the fuel must be converted to CO_2 and captured. Therefore, CO_2 capture processes for plants producing hydrogen and/or power with more than or equal to 90% CO_2 capture will be similar.

To understand the engineering design constraints associated with the WGS process, it is critical to understand the nature of the WGS reaction. The WGS reaction is an equilibrium-limited exothermic reaction with a forward reaction enthalpy of -41 kJ/mol. This means that the reaction releases heat, which increases reaction temperature, but lower reaction temperatures are necessary to achieve higher conversion. The engineering constraints are that high conversion requires either efficient heat removal from the WGS reactor or multiple reactor stages with inter-stage cooling. In general, standard commercial applications use multiple reactor stages with inter-stage cooling, because of the operational complexity and challenges of efficient heat removal.

The WGS reaction is equilibrium-limited, meaning that the distribution of reactants and products at equilibrium is determined by temperature and the initial concentrations of CO, H_2O, CO_2, and H_2 in the feed stream. The significant steam addition is beneficial, because it helps maximize the amount of CO shifted into H_2 by the equilibrium-limited WGS reaction. Another benefit of significant steam addition is that the steam provides additional thermal mass to absorb the heat generated by the WGS reaction, which lowers the adiabatic temperature rise during reaction. Lower reaction temperatures favor CO conversion for the WGS reaction, so the lower adiabatic temperature rise enabled by the additional steam also favors higher CO conversion.

The impact of these constraints on the design of the WGS process is that (for the lower CO conversion required for chemical production, excluding hydrogen) it consists of a single reactor with a syngas bypass loop to provide control of the CO conversion to optimize production of the target chemical product. For higher CO conversions, for power and hydrogen production applications with 90% CO_2 capture or more, the typical WGS process employs multiple reactors with inter-stage cooling and possible steam addition to maximize CO conversion.

The available commercial catalysts for WGS also add constraints for the WGS process. The three commercial WGS catalysts are high-temperature shift (HTS), low-temperature shift (LTS), and sour-gas shift (SGS) catalysts. As discussed in Chapter 6, the wider operating temperature range for the HTS and SGS catalysts reflects that these catalysts are typically used for bulk conversion in process streams with large CO concentrations and where high levels of CO conversion are needed. LTS catalysts are typically used for final polishing to achieve effluent CO concentrations of less than 1 mol% due to its high reactivity. High concentrations of steam in the syngas are required to minimize the deactivation of the catalyst by coking, because HTS and SGS catalysts will be exposed to high temperatures. A general rule of thumb for the steam-to-CO ratio is 2, but some manufacturers recommend a steam-to-dry gas ratio of 2.

Although other potential poisons for these catalysts (e.g., arsenic [As], chlorides) can be found in syngas, the primary poisons in syngas for these catalysts are hydrogen sulfide (H_2S), carbonyl sulfide (COS), and carbon disulfide (CS_2). Optimal operating conditions for HTS and LTS catalysts require the highest level of sulfur removal, so WGS processes based on these catalysts would need to be operated after syngas cleanup. In contrast, SGS catalysts preferentially operate in high-sulfur conditions since their active component is sulfided cobalt/molybdenum (Co/Mo).

The simultaneous operating constraints of temperature, high steam concentration, and suitable sulfur concentration limit the potential integration options for the WGS process in a gasification-based application. The HTS and LTS processes would need to be downstream of the desulfurization process in the syngas cleanup process. However, as the syngas effluent from the standard commercial desulfurization processes are at lower temperatures (less than or equal to 100°F) and have low steam concentrations, significant preconditioning of the syngas would be required as an enabling system for the WGS process. Alternatively, an SGS process could use a quenched raw syngas stream. By appropriately quenching the raw syngas, almost all the steam requirement is provided by vaporization of quench water during the quench process at a suitable operating temperature for the SGS catalyst. The high sulfur concentration in this quenched syngas, which would be present for most available fuels is also fully compatible with SGS catalyst operation. The SGS process also catalyzes COS hydrolysis, which is necessary to ensure adequate sulfur removal for most acid gas removal systems, including Selexol™. Using the SGS reactor to simultaneously complete CO-shift and COS hydrolysis eliminates the capital and operating costs for a dedicated COS hydrolysis process. Consequently, an SGS-based WGS process is the standard process used for CO conversion for coal gasification applications among all available commercial technologies. Emerging technologies are looking at higher temperature desulfurization processes and potentially decoupling sulfur and CO_2 removal, which would enable potential integration options that are not practical with current commercial technologies. One example is Research Triangle Institute's (RTI) warm gas desulfurization process, which enables the use of an HTS system and can be used with any of the available commercial CO_2 capture processes, achieving sulfur removal essentially identical to that of a Rectisol® process at approximately 68% of the cost of a dual-stage Selexol™ process [5]. Prospective project developers should consult with vendors to determine the most appropriate catalyst to suit their needs for a given application and known raw syngas composition.

12.2.2 CARBON DIOXIDE CAPTURE TECHNOLOGIES

The four primary CO_2 capture processes are absorption (solvents), adsorption (sorbents), membranes, and cryogenic separation. Selection of a CO_2 capture process is dependent on the operating conditions of the process stream; CO_2 partial pressure; temperature, pressure, and composition of the process stream; target amount of CO_2 capture; and required specifications for both the effluent process gas and CO_2 product stream. The key technical challenge for CO_2 separation is overcoming the thermodynamic penalty associated with extracting the CO_2 from a mixture. This thermodynamic penalty can be reduced for applications where favorable CO_2 operating conditions exist, namely high CO_2 partial pressure. These conditions are typically available for pre-combustion CO_2 capture from high-pressure, oxygen-blown gasification and are the basis for the selection criteria discussed in this chapter. A significant amount of R&D is being conducted for post-combustion CO_2 capture applications with the goal of reducing the energy penalty associated with CO_2 capture. Some highlights of this research are included, as any disruptive technology developed could be readily adapted and exploited to lower the energy penalty for pre-combustion CO_2 capture. Project developers should select a CO_2 capture scheme/technology that optimizes the CO_2 capture rate and resulting CO_2 purity while staying within plant budget parameters.

The following sections offer descriptions of these different CO_2 capture process technologies.

12.2.2.1 ABSORPTION (SOLVENT)

Carbon dioxide capture by absorption refers to CO_2 removal from process gas into a liquid medium through physical or chemical mechanisms and regeneration through temperature/pressure swing. A simple thermal swing system for monoethanolamine (MEA) is shown in Figure 12-2. The ideal solvent has the following characteristics: readily available in large-scale, low-cost, high affinity toward CO_2, low affinity for all non-CO_2 compounds, low-volatility (minimized losses), and easily regenerable (low-energy). It is also desirable to have highly stable solvents that are resistant to degradation in the presence of impurities in the process gas and at typical operating conditions.

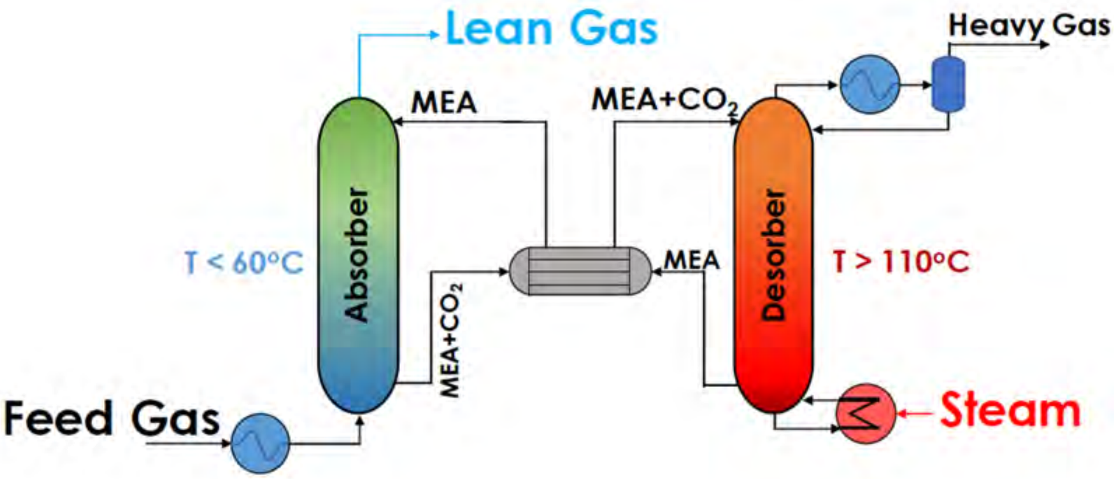


Figure 12-2. CO_2 capture by MEA absorption and thermal regeneration of MEA

The absorption of CO_2 from natural gas is a mature technology that has been used since the 1930s. The selection of absorption technology depends on the CO_2 partial pressure in the feed gas. Carbon dioxide partial pressure is the product of feed gas pressure (process gas stream that contains CO_2) and the CO_2 mole fraction. Typically, natural gas streams are available at elevated pressures, though their CO_2 content can vary widely, resulting in a range of CO_2 partial pressures and technology options. The solvent technology options for CO_2 capture from natural gas based on CO_2 partial pressure in the process gas (Y-axis) and the desired recovery pressure in the CO_2 product (X-axis) are shown in Figure 12-3.

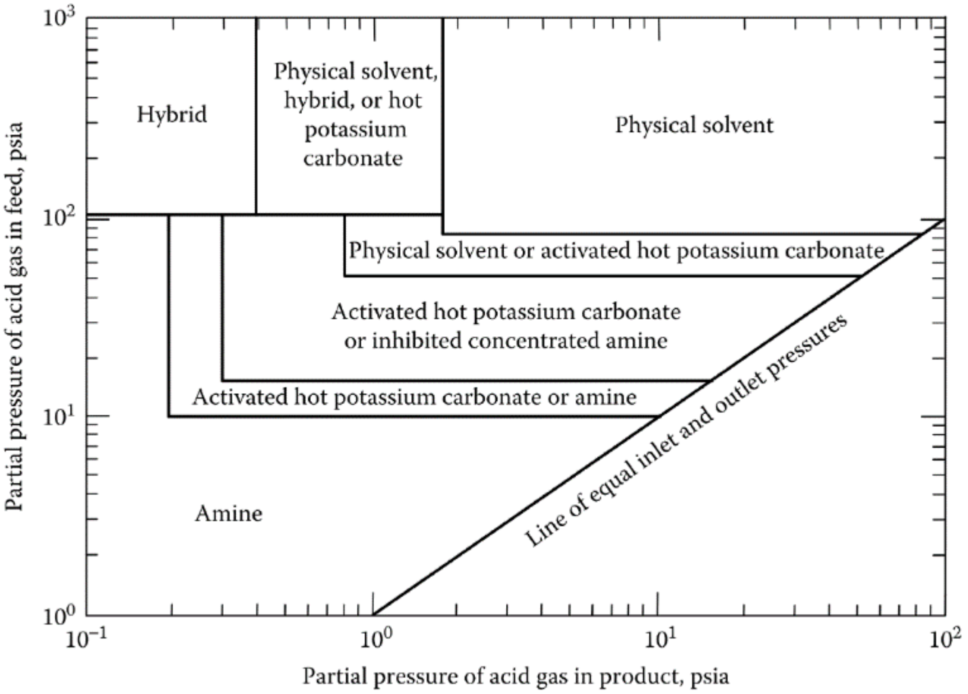


Figure 12-3. Options for commercially available CO_2 removal technologies[2]

Solvents (amines) based on chemical mechanisms for CO_2 separation are the preferred separation option when CO_2 partial pressures in the process gas are low (Figure 12-3). These are the typical conditions that exist in post-combustion CO_2 capture where the flue gas is at atmospheric pressure with low CO_2 concentration (5 to 15 mol%). The CO_2 byproduct is recovered at an even lower CO_2 partial pressure, which requires compression of the CO_2 byproduct to support CO_2 utilization and/or storage. By contrast, physical solvents are commercially used in applications with high CO_2 partial pressure in the process gas. These physical solvents also generate a CO_2 byproduct with higher pressure, lowering compression requirements for CO_2 utilization and/or storage. Therefore, physical solvents offer commercially viable processes for pre-combustion CO_2 capture from high-pressure, oxygen-blown gasification processes.

The primary reason that physical solvents are more suitable for CO_2 capture from process gases with high CO_2 partial pressure is lower energy requirements. Physical solvents are typically regenerated by pressure swing, in which the pressure of the solvent is reduced, allowing desorption of the physically absorbed CO_2. For chemical solvents, regeneration typically requires the addition of thermal energy (temperature swing) to reverse the chemical absorption mechanism. The CO_2 capacity of chemical solvents is also fixed by CO_2 solubility of the solvent, whereas the CO_2 capacity of physical solvents increases with CO_2 partial pressure in the process gas. These properties cause process economics to become more favorable for physical solvent as the CO_2 partial pressure in the process gas increases. At intermediate CO_2 partial pressures, the economics of chemical and physical solvent processes are similar for CO_2 capture.

Another factor influencing the selection of the absorption technology is the other contaminants in the process gas. For pre-combustion syngas generated using high-pressure, oxygen-blown gasification, sulfur compounds, primarily H_2S, will be present in significant concentrations (more than 1,500 parts per million by volume [ppmv]). As H_2S is an acid and interacts with the solvents in a similar manner to the acidic CO_2, H_2S and CO_2 are generally removed simultaneously. Although this offers an opportunity for process intensification, it results in challenges when recovering separate high-purity sulfur and CO_2 byproducts. The two most predominant commercial processes for acid gas removal from high-pressure syngas are Selexol™ and Rectisol®. Both use physical solvents and capture CO_2 and H_2S. Adjusting the gas pressure during regeneration allows for more control (Henry's Law) for preferentially desorbing H_2S or CO_2.

12.2.2.2 CHEMICAL SOLVENTS

Chemical solvents are used for the removal of CO_2 from syngas derived from natural gas steam reforming, where there is essentially no sulfur. This is true for the use of syngas to produce ammonia (NH_3). The NH_3 production catalyst is extremely sensitive to oxides, which includes CO_2. Therefore, chemical solvents, which can achieve significantly lower CO_2 partial pressures in the treated gas, are advantageous. BASF's OASE® White is a standard commercial process used for CO_2 capture in NH_3 production from natural gas.

Chemical solvent systems involve absorption and regeneration where lean (i.e., CO_2-free) solvent absorbs CO_2 in the process gas at low temperature by a complex chemical mechanism. The CO_2 loaded (rich) solvent is regenerated (desorbed) by applying steam heating in the desorber column (also called stripper) at elevated temperatures. The absorber operates at low temperature, whereas the desorber (or stripper) is at high temperature (temperature swing shown in Figure 12-2). Large-scale applications of CO_2 removal with solvents are utilized in the natural gas processing industry. Amines, a class of chemical solvents, are excellent absorbents for CO_2, H_2S, and COS. Typically, a single amine or a blend of two or more amines in aqueous solution chemically reacts with CO_2 in an acid-base reaction, forming a salt. The salts can be converted back into acid and the base (amine) at high temperatures and reduced acid gas partial pressures, regenerating the solvent and releasing CO_2.[2] The same acid-base reactions occur with H_2S and, to a more limited extent, COS, with H_2S being released during regeneration.

MEA has become the benchmark commercial technology for solvent-based post-combustion CO_2 capture due to its higher rates of reaction with CO_2 compared to other amines. The reboiler duty required to strip CO_2 out of 30 wt% aqueous MEA solvent in a state-of-the-art amine plant (intense heat integration) is 3.6 gigajoules (GJ)/metric ton CO_2 (1,548 British thermal units [Btu]/lb CO_2 or 3,600 kJ/kg CO_2). R&D efforts (Table 12-2) have focused on advanced, high-efficiency solvents that could result in significant reductions in reboiler duty. These R&D efforts on post-combustion flue gas CO_2 capture are increasing the potential for the development of a disruptive CO_2 capture technology that could also be applied for pre-combustion syngas generated by high-pressure, oxygen-blown gasification. A noteworthy feature of this R&D is the state-of-the-art MEA plant used as the benchmark reference.

Table 12-2. R&D for post-combustion CO_2 capture solvent-based processes[3]

DEVELOPER	SOLVENT	BENEFIT(S)
BASF	OASE® Blue	• More stability than MEA, leading to lower operational costs. • Lower steam rate due to lower circulation rates than MEA (2.4 to 2.6 GJ/ton CO_2 versus 3.6 GJ/metric ton CO_2 for MEA), resulting in higher plant efficiency and lower fuel costs.
Kansai Electric Power Company (KEPCO) and Mitsubishi Heavy Industries, Ltd. (MHI)	KS-1™	• Lower capital costs with the use of carbon steel equipment. • KS-1™ achieves 20% lower regeneration energy compared to MEA. • NH_3 emissions in the treated flue gas were also lower than with MEA.
Shell	Shell Cansolv DC-201®	• Fast kinetics, low degradation rates, and high resistance to oxidation. • Separates both CO_2 and sulfur dioxide (SO_2) from the flue gas. • Demonstrated construction and material cost reduction. • 50% lower solvent circulation rate compared to MEA and a thermal regeneration energy of 2.4 GJ/metric ton CO_2 (35% lower than that of MEA). • Optimum solvent concentration for minimizing thermal regeneration energy is 45 to 55 w/w%.
Babcock Hitachi	H_3-1	• In pilot studies, thermal regeneration energy was reduced by 33% compared to MEA (2.4 GJ/metric ton CO_2 versus 3.6 GJ/metric ton CO_2) while capturing a higher fraction of the incoming CO_2 (96% capture versus 92% capture with MEA).
Korea Electric Power Company (KEPCO)	KoSol Family of Solvents	• The thermal energy requirement for regeneration is estimated to be 30% lower than MEA (2.5 to 3 GJ/metric ton CO_2). • KoSol solvents offer better resistance to oxidative and thermal degradation, lower vapor pressure, higher loading capacity (with less corrosion), and lower energy consumption (30 to 40% lower) compared to 30 w% MEA
Pacific Northwest National Laboratory (PNNL)	CO_2BOL	• Lower volatility and lower stripper temperature via polarity-swing-assisted regeneration, which reduces the steam demand for solvent regeneration.
General Electric	Aminosilicones	• Lower vapor pressure, higher thermal stability, lower thermal energy requirement, resulting in up to 27% lower thermal energy consumption than MEA.
ION Engineering	Amine-Organic Solvent	• Low water-content solvent, which has relatively low heat of regeneration (1.6 to 2.5 GJ/metric ton CO_2), is resistant to thermal degradation, and has low buildup of heat-stable salts on exposure to SO_X and NO_X.
University of Texas	Piperazine (PZ)	• Fast kinetics; low degradation rate. Carbon dioxide is regenerated at elevated temperature to produce higher pressure CO_3 during regeneration.

12.2.2.3 PHYSICAL SOLVENTS

Physical solvents utilize a similar absorption-regeneration process used for separating the CO_2 from the process gas. The separation of the CO_2 from the process gas is based on physical absorption of the CO_2, rather than a chemical reaction between the solvent and CO_2. In a physical absorption process, the absorber requires a high CO_2 partial pressure to efficiently capture the CO_2. The CO_2-rich solvent leaving the absorber is depressurized, making desorption of the CO_2 more thermodynamically favorable and regenerating the solvent. It is also possible that some thermal regeneration is included to remove additional CO_2, effectively increasing the CO_2 capacity of the solvent. The high CO_2 partial pressure requirements in the absorber also make physical absorption processes ideally suited for pre-combustion CO_2 capture from high-pressure, oxygen-blown gasification applications.

Solvent-based separation processes were developed for removing acid gases, including CO_2 and H_2S, from natural gas. Although this simultaneous separation of CO_2 and H_2S from the process gas allows for process intensification of the absorption process, the process requirement to generate a treated process gas, CO_2 byproduct, and a sulfur-rich byproduct increases the complexity of the regeneration process. Due to the potential for pressure swing regeneration, physical solvents offer the potential to selectively control desorption of these compounds and generate distinct CO_2 and sulfur byproduct streams.

For commercial gasification, the two most prominent physical absorption processes are Selexol™ and Rectisol®. The Rectisol® process is exclusively used for chemical production applications, especially gasification, because the absence of impurities in the treated syngas maximizes the life of downstream catalysts. The prime catalyst poisons present in syngas are the sulfur compounds consisting of H_2S and COS. The Rectisol® process reduces the sulfur concentration in the treated process gas to less than 100 parts per billion by volume (ppbv). The Rectisol® process removes hydrocarbons, NH_3, and hydrogen cyanide (HCN); achieves effluent CO_2 concentrations in the ppmv range; and remove both H_2S and COS, which many other solvents cannot. The simultaneous removal of CO_2 with other contaminants is a significant benefit for chemical processes where the CO_2 is either an inert diluent or (at worst) an inhibitor to the downstream chemical processes. In most chemical plants, the captured CO_2 is released to the atmosphere, although it does represent a valuable opportunity for near-term implementation of CO_2 utilization and/or storage applications.

The Rectisol® process uses chilled methanol as a solvent and was independently developed by Linde and Lurgi. Due to the high vapor pressure of methanol, the process is operated at a temperature in the range of -30 to -100°F. The low operating temperature enables superior separation performance for multiple contaminants, resulting in thermal penalties due to the higher operating temperatures required in downstream processes and the need for refrigeration to achieve the low operating temperatures.

These thermal penalties, which decrease the thermal efficiency in gasification-based power plants, is the reason for NETL using dual-stage Selexol™ acid gas separation processes as the benchmark acid gas separation technology in their system analysis studies of coal gasification in integrated gasification combined cycle (IGCC) systems.

The Selexol™ process is based on mixtures of dimethyl ethers of polyethylene glycols as a solvent. The difference in H_2S (higher) and CO_2 selectivity of the solvent enables generation of an H_2S-rich byproduct stream and a relatively pure CO_2 byproduct steam. In a dual-stage process, the H_2S is first selectively removed in an absorption stage. The intermediate syngas stream is then treated in a second absorption stage to remove the CO_2. Pressure swing regeneration is used to regenerate the CO_2-rich solvent from the second absorption stage, generating a relatively pure CO_2 byproduct. Thermal regeneration of the rich solvent from the H_2S absorption stage generates an H_2S-rich stream that minimizes the CO_2 concentration in the stream. A schematic of the dual-stage process is shown in Figure 12-4.

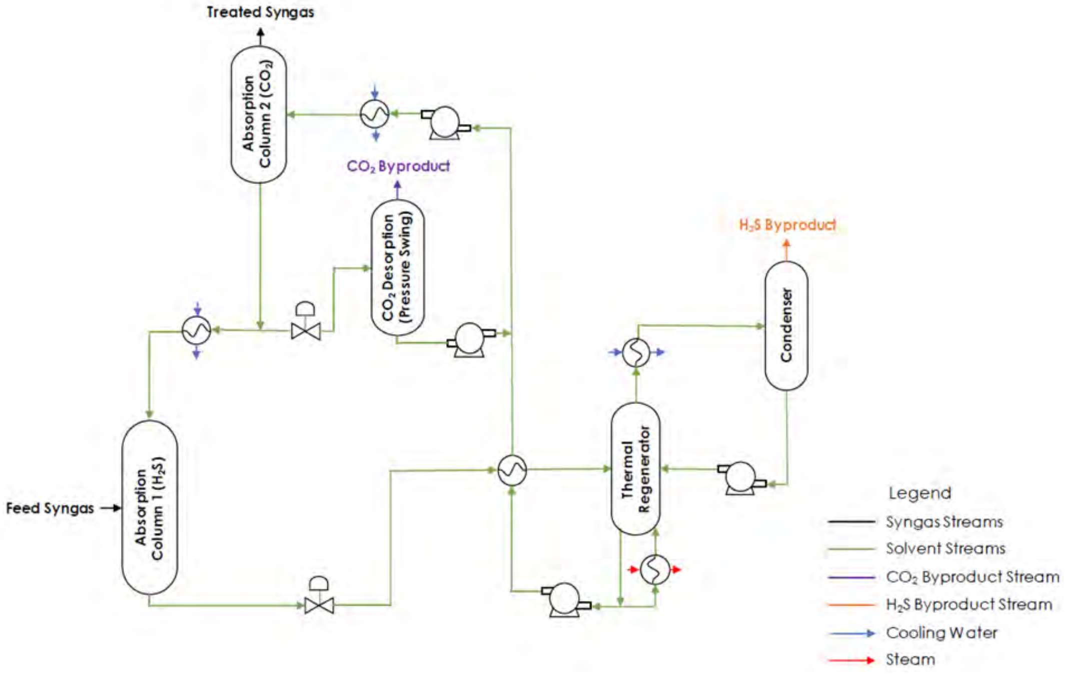


Figure 12-4. Schematic of dual-stage Selexol™ process

Dual-stage Selexol™ processes are preferred for IGCC applications with more than or equal to 90% CO_2 capture, because they operate at higher temperatures, which eliminates the need for refrigeration associated with Rectisol®. The elimination of this refrigeration requirement improves the thermal efficiency of the process in IGCC applications with more than or equal to 90% CO_2 capture, where the parasitic losses associated with CO_2 capture decrease thermal efficiency and net power output.[1]

12.2.2.4 ADSORPTION

Carbon dioxide capture by adsorption refers to the process where CO_2 is attached to a solid by means of a gas/solid interaction that separates the CO_2 from the process gas. These interactions between CO_2 and the solid can be based on either physisorption (weak Van der Waals forces) or chemisorption (strong covalent binding forces) (Figure 12-5). As a result of the fundamental interactions between CO_2 and the solids behaving like the interactions between CO_2 and the solvent in absorption, adsorption processes have many of the same features found in absorption processes. The similarities and differences between adsorption and absorption processes for CO_2 capture are included in Table 12-3.

Table 12-3. Comparison of adsorption and absorption processes for CO_2 capture

CATEGORY	ADSORPTION		ABSORPTION	
	PHYSICAL	CHEMICAL	PHYSICAL	CHEMICAL
Separation Mechanisms	Weak Van der Waals forces.	Covalent bonding.	Physical dissolution.	Chemical reaction.
Regeneration	Pressure swing is standard.	Temperature swing is standard.	Pressure swing is standard.	Temperature swing is standard.
Selection Criteria	High CO_2 partial pressures.	Low CO_2 partial pressure and/or extremely low effluent CO_2 concentrations are required.	High CO_2 partial pressures.	Low CO_2 partial pressure and/or extremely low effluent CO_2 concentrations are required.
Process Configuration	Adsorption processes are typically cyclic fixed-bed processes in which different process gases are fed during different (adsorption and regeneration) steps in the cycle, because solid adsorbents are more difficult to move.		Solvents can be easily pumped, allowing the solvent to be move between fixed process equipment (absorber and stripping columns).	
Energy Requirements (temperature swing processes)	The primary energy requirement is to break the bonds between CO_2 and solid. Additional energy is also needed for sensible heat for the solid.		The primary energy requirement is to break the bonds between CO_2 and solvent. Additional energy is also needed for sensible heat for the solvent, as well as latent heat of vaporization for aqueous solvents.	
Issues	Capacity loss due to irreversible chemical interactions with contaminants. Low CO_2 capacities requiring large adsorbent volumes and large process equipment. High production costs for promising materials. Competitive adsorption favors water vapor removal over CO_2.		Loss of capacity due to irreversible chemical interactions with contaminants. Thermal degradation. Solvent loss due evaporation. Solvent toxicity.	
Advantages	Wide range of potential operating conditions is possible. Lower material losses due to evaporation.		Solvents can generally be produced with low production costs in large volumes.	

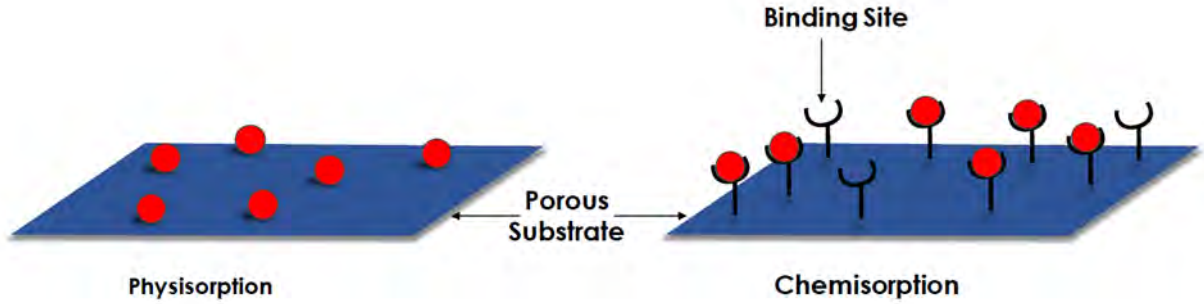


Figure 12-5. Schematic of physisorption and chemisorption

The fundamental interactions for separating the CO_2 from the process gas are similar for adsorption and absorption, so the regeneration to release the CO_2 for adsorption also employs pressure and temperature swing processes. Generic adsorption and regeneration processes occurring on the solid are shown in Figure 12-6. Physisorption processes are better suited for CO_2 removal from process gases with higher CO_2 partial pressures and favor regeneration by pressure swing; chemisorption processes are better suited to capture CO_2 from process gases where the CO_2 is present at low partial pressures and typically use heat to release the CO_2.

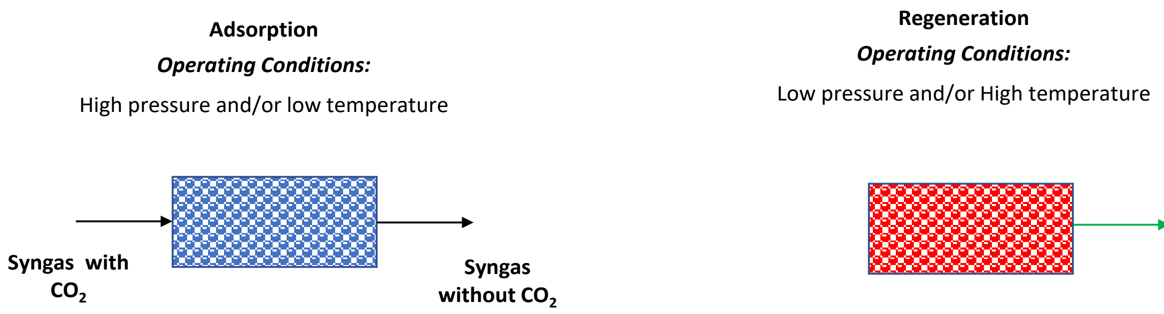


Figure 12-6. CO_2 adsorption on sorbent and regeneration of the adsorbent (release of CO_2) via pressure or temperature swing

Adsorption processes are based on solid adsorbents and the movement of the solids through the process, although not impossible, is challenging. Consequently, most adsorption processes use fixed beds of adsorbent and periodically switch the process gas to complete adsorption or regeneration in a specific fixed sorbent bed. Multiple fixed beds of sorbent are necessary to continuously treat the process gas. An example of an adsorption and regeneration cycle for a fixed-bed adsorbent process is shown in Figure 12-7.

Most of the R&D associated with CO_2 adsorption is focused on CO_2 capture from flue gas for post-combustion CO_2 capture applications. Physisorption research mainly focuses on increasing the sorbent specific surface area to maximize the absolute adsorbent capacity, while chemisorption research mainly focuses on minimizing the heat of reaction and the impact of impurities on the sorbent to decrease regeneration energy load.[4] The reasons for the focus on post-combustion CO_2 capture include the following: (1) CO_2 is the only compound that needs separation from the flue gas in post-combustion CO_2 capture applications and (2) reducing energy consumption for this separation despite the low driving force (i.e., low pressure, low temperature, and low CO_2 concentration) improves the performance of technologies for which pre-combustion CO_2 capture is not possible.

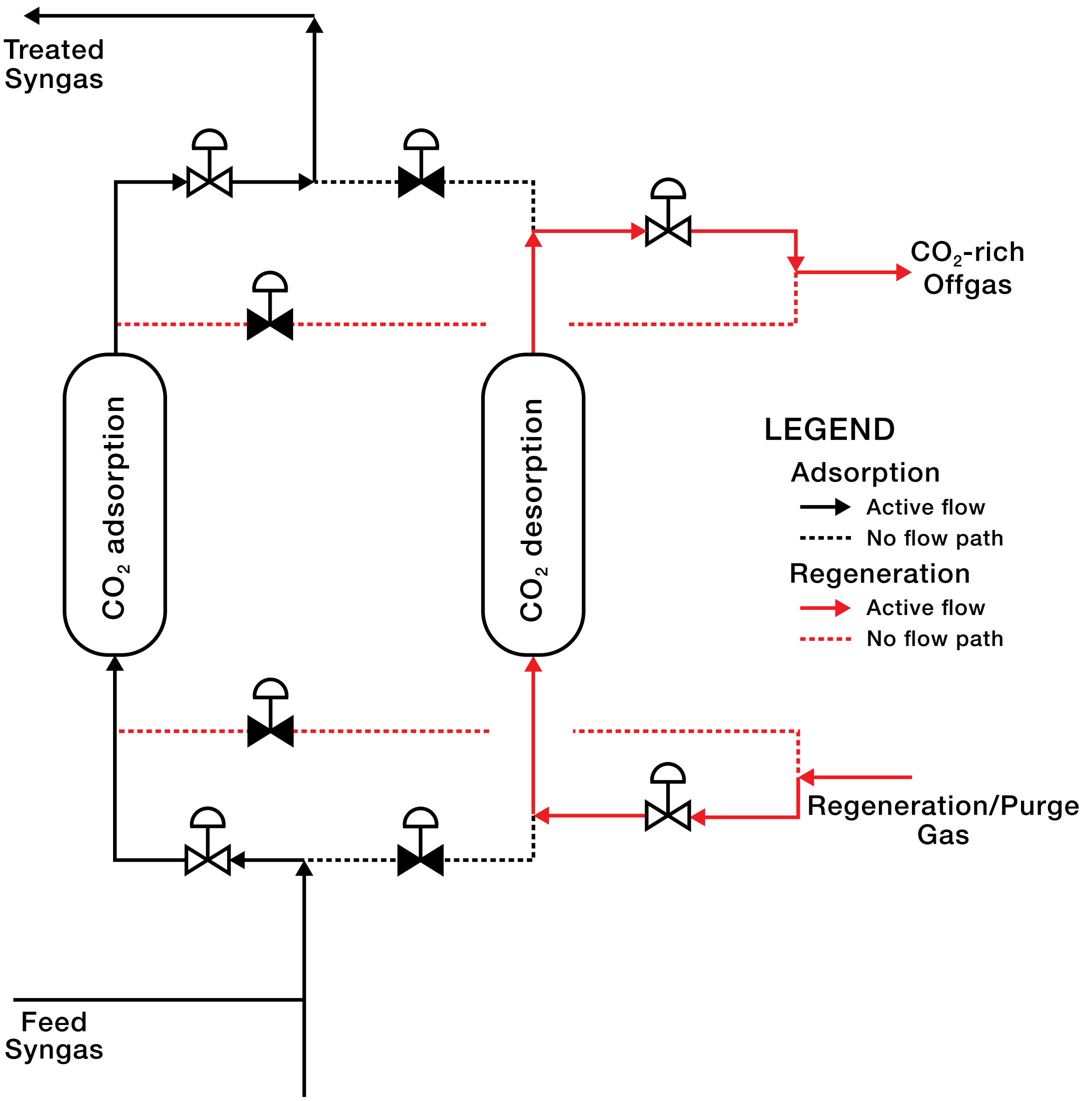


Figure 12-7. Adsorption-regeneration loop for CO_2 capture

In contrast, pre-combustion CO_2 capture is already practiced commercially for chemical production, where it is integrated with sulfur capture, which is required for chemical applications and power applications with CO_2 capture. The technical and funding challenges to identify and develop an adsorption-based process to compete with multiple existing commercial processes that can achieve high levels of both sulfur and CO_2 removal inhibit R&D efforts. However, RTI has developed a proprietary selective desulfurization system based on a regenerable sulfur sorbent. System analyses of the integration of RTI's proprietary desulfurization process and commercial amine solvent-based systems for CO_2 capture have demonstrated that improvements in capital cost and process efficiency can be achieved when compared to commercial solvent-based technologies (i.e., Selexol™ and Rectisol®) for both IGCC and chemical production applications.[5, 6] RTI's proprietary desulfurization process was successfully demonstrated in a 50-megawatt (MW) demonstration at the Tampa Electric IGCC plant in Polk County, Florida (USA) and is available for commercial deployment.[7]

One application that has used adsorption for the purification of a fuel stream is the use of pressure swing adsorption (PSA) for the final polishing of hydrogen-rich fuel gas to produce a high-pressure, ultra-high-purity hydrogen product. This approach has proven popular for retrofitting older coal gasifiers in China to extend their usefulness/life and provide valuable hydrogen.

12.2.2.5 MEMBRANES

Membrane separation processes employ selective permeation of gaseous components from one side of a membrane barrier to the other due to the concentration gradient across the membrane. The most practical concentration gradient for membrane applications results from high partial pressure of the target components on the feed side of the membrane and a low partial pressure of the component on the permeate (or downstream) side of the membrane. The gas flux through a membrane is a combination of the rate of penetrant diffusion and rate of interaction of the penetrant with the membrane surface.

For rigid materials including glassy polymers, gas flux is primarily based on diffusion and results in strong size selectivity. This size selectivity is exploited in commercial membrane applications for air separation, H_2 recovery from NH_3 purge gas, and CO_2 removal from natural gas. For syngas applications, the use of size selective membranes primarily separates the hydrogen from the other larger compounds, like CO_2, in the syngas. For applications that require relatively pure hydrogen, these size-selective glassy polymers can be used. However, for chemical production that requires syngas mixtures with specific H_2-to-CO ratios and production of high-purity CO_2 product streams, size-selective glassy polymers are not a viable separation option.

The interactions between the penetrant and the membrane provide an alternative mechanism for selective gas flux across the membrane. For polymeric membranes, the solubility of the penetrant in the polymer is a property that can be used for selective separation. Polymers where solubility selectivity dominates over diffusion selectivity are typically rubbery materials. The CO_2-to-H_2 selectivity for several glassy and rubbery polymer materials is included in Table 12-4.

Table 12-4. Comparison of relative CO_2 and H_2 permeation properties in different polymers

	POLYMER	SELECTIVITY*
TYPE	NAME	CO_2/H_2
Glassy	Polysulfone	0.40
	Polycarbonate	0.48
	Cellulose Acetate	0.40
Rubbery	Cis-Polyisoprene	2.7

**Based on pure gas data from Zolandz and Fleming*[8]

In general, membranes are preferentially suited for bulk separations and require additional polishing processes to achieve higher purity products. As the flux across the membrane is inversely proportional to the thickness of the membrane, most commercial membrane systems consist of a thin layer of selective membranes overlaid on a material that provides structural support while remaining sufficiently porous so that it does not impede gas flow. The cross section of a typical commercial polymeric membrane is shown in Figure 12-8.

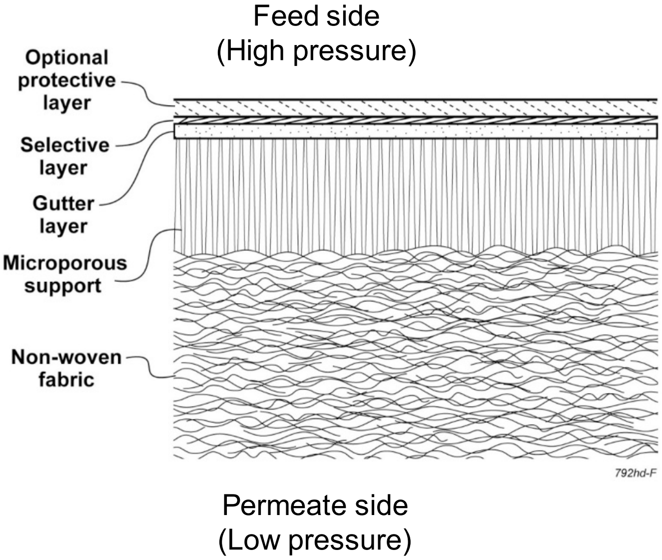


Figure 12-8. Typical structure of a commercial polymer

Membrane Technology and Research, Inc. (MTR) has developed the Polaris™ class of membranes for CO_2 separation, which possess high CO_2 selectivity over other gases. The Polaris membrane was initially developed for post-combustion CO_2 capture applications and has a CO_2/N_2 selectivity of 50 and a permeance up to 3,000 gas permeance units (GPU). For CO_2 separation in syngas applications, Polaris membranes have demonstrated a CO_2/H_2 selectivity of more than or equal to 6 with actual commercial spiral wound membrane modules tested at the National Carbon Capture Center (NCCC) in Wilsonville, Alabama (USA). During the NCCC testing, the Polaris membrane modules were coupled with refrigeration and permeate recycle (Figure 12-9) to capture more than 400 kg/day of high-purity, liquid CO_2 (more than 97% CO_2) from a feed syngas mixture with 12 vol% CO_2 with no evidence of membrane degradation over more than 800 hours of operation.

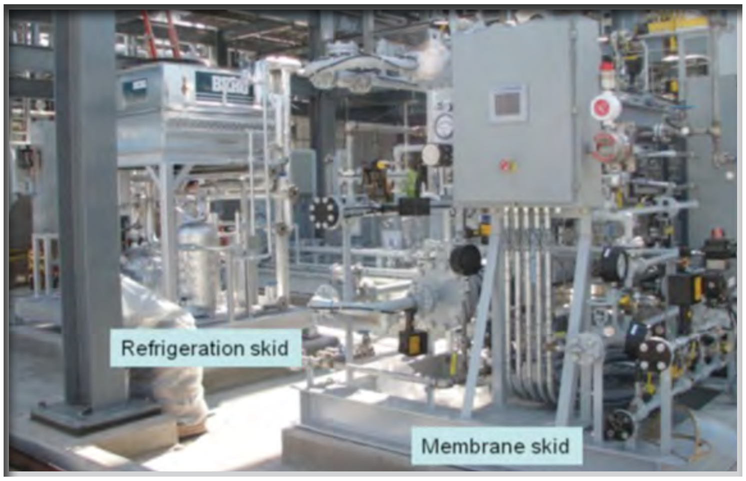


Figure 12-9. Pilot-scale liquid CO_2 skid at NCCC

12.2.2.6 CRYOGENIC PROCESS

Cryogenic CO_2 separation processes separate the CO_2 from the process gas via phase change of the CO_2 from a gas to a liquid. Carbon dioxide is liquid at any temperature between its triple point (-56.6°C [-69.9°F]) and its critical point (31.1°C [88°F]) by compression to the corresponding liquefaction pressure and removing the latent heat, which is 348 kJ/kg at its triple point. The key technical challenge is that the phase change from gas to liquid favors higher pressure and lower temperatures. Thus, cryogenic CO_2 separation is extremely energy intensive, as it requires compression of the entire process gas (i.e., CO_2 and other components) and refrigeration to cool the gas and remove the latent heat at suitable temperatures. Both higher pressures and lower temperatures are required for CO_2 condensation as the partial pressure of the CO_2 in the process gas decreases. Even syngas mixtures that have been partially to highly shifted to maximize CO conversion to CO_2 require cryogenic temperatures for the liquefication and removal of the CO_2. For cryogenic temperatures above -20°C (-4°F), a single-stage refrigeration loop can be used. When the cryogenic temperature required is below -25°C (-13°F), multiple refrigeration loops are required, increasing the capital and operating cost of the refrigeration process.

One of the advantages of cryogenic separation is the ability to recover a high-pressure CO_2 product, whereas most other separation techniques recover the CO_2 between atmospheric pressure and 150 pounds per square inch gauge (psig) for pre-combustion CO_2 capture applications. The liquid CO_2 product must be heated to ambient temperatures in all applications for further downstream processing. Although a portion of this heat can be recovered from cooling the process gas fed to the process, it is an additional processing step, which increases the amount of equipment needed and capital cost of the process.

12.2.2.7 CARBON DIOXIDE COMPRESSION

Although compression is technically not part of the CO_2 separation process, it is essentially a process requirement. All current CO_2 separation processes, except for cryogenic processes, generate the CO_2 byproduct at a pressure of less than or equal to 150 pounds per square inch absolute (psia) and more commonly near or below atmospheric pressure. However, most CO_2 utilization and storage applications require high pressure (more than 500 psia at a minimum; storage pressures regularly exceed the critical pressure of CO_2, which is 1,076 psia). Consequently, CO_2 compression is required by CO_2 separation processes and must be performed before or alongside some amount of cooling for effective integration with downstream CO_2 utilization or storage processes.

The high cost for the compression of contaminants in the CO_2 favors separation processes that generate a CO_2 byproduct with contaminants concentrations that meet specifications for downstream applications. One contaminant that is typically present in higher than allowable amounts is water vapor. This water vapor is typically removed as liquid water after condensation during the inter-stage cooling between compression stages. A final polishing step can be used to lower any remaining water vapor to the desired specification concentration. Similarly, final polishing treatment for any other contaminants to meet specifications for the CO_2 byproduct can be included as part of the compression process to meet specifications for the CO_2 product in the most cost-effective manner possible. Water vapor in CO_2 gas streams (so called wet CO_2) can lead to corrosion of carbon steel piping and equipment if a separate liquid water phase forms in the system, so dehydration below the chance for a liquid water phase to be present in most operating scenarios is recommended. Dehydration of CO_2 can be accomplished via several different technologies, including glycol dehydration, molecular sieve dehydration, and CO_2 subcooling. The extent of dehydration required depends upon the downstream use of the CO_2.

For gas compression, multiple stages of adiabatic compression with inter-stage cooling are typically employed. This approach not only enables the use of simpler compressor equipment, but also results in a closer approximation to isothermal compression, which results in higher compressor efficiency for converting energy into compressed CO_2. R&D is being conducted by both industry and academia into CO_2 compression to reduce the operating and capital costs.

Alternatively, the CO_2 byproduct can be cooled until the gas-phase CO_2 condenses into a liquid. Liquid compression requires simpler equipment (i.e., pump versus compressor) and less work than gas compression to reach the specified pressure due to the essentially incompressible nature and higher density of liquids. As the liquefaction temperature for near atmospheric CO_2 is below -20°C (-4°F), more complex refrigeration processes are required to reach the CO_2 liquefaction temperature. This increases the capital cost and energy intensity of this process. In addition, the resulting liquid will be very cold, requiring heat to reach the final temperature. Compression schemes, which integrate gas-phase compression, cooling, and condensation of the compressed gas followed by pumping of the condensed liquid to achieve the target pressure specification, have also been considered as part of optimization efforts to lower compression energy requirements and costs.

The project developer should be aware that the primary selection criteria for the CO_2 compression process is operating costs, which are primarily related to the energy costs. Most CO_2 separation processes generate a large fraction of the CO_2 byproduct at near-atmospheric conditions, so gas-phase compression is the primary process that has been evaluated. However, with MTR's Polaris membrane, the final cryogenic purification of the CO_2 byproduct favors liquid compression over gas-phase compression as the most economically competitive solution. Therefore, the cost and options for CO_2 compression should be evaluated as part of CO_2 capture. Other factors for compressor selection include, but are not limited to, capital costs, inlet mass flow rate, suction pressure, discharge pressure, turndown and sparing requirements, type of driver for the machine, and project duration.

12.2.3 SELECTION OF CARBON DIOXIDE CAPTURE TECHNOLOGY

The evaluation of CO_2 capture for any modular gasification process is critical due to the impact that CO_2 capture has on the process economics. Different end-use applications for CO_2 have different specifications for the utilized CO_2. Developers should ensure that the final CO_2 purity meets the end user's demands, which can affect both the choice of CO_2 capture technology and choice of any number of other post-processing equipment (see Chapter 10).

Although CO_2 capture can be achieved via pre-combustion, oxy-fuel combustion, and post-combustion approaches, pre-combustion enables CO_2 separation under the most favorable thermodynamic conditions, making pre-combustion capture the most promising approach for modular gasification systems. This advantage is strongest for high-pressure, oxygen-blown gasification technologies. As a result of this advantage, pre-combustion CO_2 capture from high-pressure, oxygen-blown gasification systems serves as the reference system for describing CO_2 separation design issues and selection criteria in this chapter.

Prior to the actual technology selection for CO_2 capture, the project developer must determine the amount of CO_2 to be captured. For chemical applications, the primary objective is to maximize the amount of carbon in the feed that is converted into chemical product, which sets the amount of CO_2 generated and the maximum CO_2 that can be captured. For power applications, the amount of CO_2 to be captured is a design specification that must be set by the project developer.

The most commercially developed CO_2 capture technologies are based on absorption. With commercial, high-pressure, oxygen-blown gasification for chemical production, Rectisol® is the gold standard solvent-based syngas cleanup process. For power generation applications requiring CO_2 capture, Selexol™ is competitive with Rectisol®, because it offers significant economic and thermal efficiency benefits.

For large-scale applications, the success of these existing commercial processes inhibits the entry of alternative and novel technologies. However, for smaller, modular approaches, the project developer should consider alternate technologies that may be more suitable for small-scale or modular applications and/or be more economically viable. As demonstrated with RTI's desulfurization sorbent, the decoupling of the sulfur and CO_2 removal processes can be economically viable and improve process efficiency.

The project developer should monitor post-combustion CO_2 capture R&D for potential disruptive developments that could be adapted to the potentially more favorable separation conditions (i.e., higher CO_2 partial pressure and temperature) available for the specific gasification technology utilized.

Finally, the project developer should include the capital and operating costs of CO_2 compression during the evaluation of the CO_2 capture process due to the energy intensity of CO_2 compression, especially when compression is required for downstream end use.

12.3 CARBON DIOXIDE UTILIZATION

There are many current commercial uses of CO_2, including use as a flame retardant and in refrigeration systems, welding systems, water treatment processes, and carbonated beverages. It is also used in the metals industry to enhance the hardness of casting molds and soldering agents (when dry). Carbon dioxide is also valuable for EOR operations, which increase the amount of oil that can be harvested. Most of these uses represent a small volume of usage compared with the net amounts of CO_2 generated for meeting energy demands using fossil fuels.

There are other promising CO_2 utilization opportunities for the CO_2 from fossil fuel usage, such as mineral carbonation to generate construction materials, chemical conversion to produce chemicals and fuels, and biological conversion to produce chemicals and fuels. Although considerable public and private R&D is dedicated to these opportunities, the key challenge remains the economic viability of the process compared with existing commercial technologies and processes. The main components affecting the economic viability of CO_2 utilization include a readily available CO_2 source, a low-cost energy supply that does not generate CO_2, a mismatch between supply and demand, and the large distances between production and end users. Project developers should consider that modular gasification systems could be sized and located to exploit available niche opportunities for the power and chemicals produced by gasification and CO_2 utilization opportunities. Consequently, project developers and potential plant owners should invest resources to evaluate potential niche CO_2 utilization opportunities during their evaluation of project feasibility. The following sections describe potential CO_2 utilization opportunities that could be exploited by modular gasification systems.

12.3.1 ENHANCED OIL RECOVERY

Oil extraction declines as the pressure in the reservoir drops with the production. EOR refers to the methods that restore pressure in oil wells using solvents and/or heat. Carbon dioxide injection is one of the preferred EOR methods that causes a portion of immobile oil to expand and flow to the wells for extraction due to CO_2's ability to mix well with crude oil. EOR is one of the feasible options for CCUS for CO_2 captured from power and chemical production facilities.

Conventional oil production is a relatively inefficient process, typically extracting approximately one-third of the oil present in the ground. "Secondary" recovery techniques, such as the addition of water or natural gas, can enhance recovery to 35 to 45% of the original oil-in-place (OOIP).[9] An additional 10 to 20% (40 to 65% of total OOIP) can be extracted by injecting CO_2 to increase reservoir pressure, decrease oil viscosity, and develop miscibility between the injected CO_2 and reservoir oil. Many U.S. industrial facilities, such as Enid Fertilizer Facility in Oklahoma (USA) and Shute Creek Natural Gas Processing Facility in Wyoming (USA), have been capturing and selling CO_2 for several decades, and Port Arthur Hydrogen Plant in Texas (USA) has been supplying CO_2 for EOR since 2013. In these examples, CO_2 separation is an existing process in the plant and the plant (i.e., CO_2 source) is near the EOR sites or existing pipelines connected to the EOR sites. These facilities demonstrate that modular gasification systems with CO_2 capture can exploit the favorable economic conditions of a local niche market.

12.3.2 MINERAL CARBONIZATION

The global demand for construction materials is huge. Each year, nearly 30 billion tons of concrete are produced with mineral carbonates as one of the key components.[10] The chemistry for making calcium (Ca) and magnesium (Mg) carbonates is well known and exploited in the commercial production of construction materials. During the production of mineral carbonates, CO_2 is chemically bound into stable, long-lived mineral carbonates, making construction materials a potential repository for CO_2. In addition, mineral carbonation reactions based on alkaline earth metals and CO_2 are thermodynamically favored at mild operating conditions near ambient temperatures and pressures. Based on global market size for construction materials, replacing existing products with mineral carbonates produced with captured CO_2 could enable annual consumption of up to 1 gigaton (Gt) of CO_2.[11]

The realization of the CO_2 utilization potential of mineral carbonation processes does not come without challenges (Table 12-5).[12] The challenges provided in the table are specific to the large-scale replacement of existing construction materials with mineral carbonates. Project developers should consider that local modular gasification processes offer an opportunity to provide a suitable CO_2 stream to capture niche markets for these purposes.

Table 12-5. Technical and market challenges for CO_2 utilization via mineral carbonization

CHALLENGE	DESCRIPTION
Chemical and Physical Barriers to Carbonation	Carbonation is impeded by competing hydration reactions, formation of surface passivating films, and liquid water acting as a gas-phase diffusion barrier for CO_2 into the pores.
Barriers to the Carbonation Process	Thermodynamically favorable carbonation required alkaline earth precursors, which include portlandite (Ca[OH]2) produced by roasting limestone ($CaCO_3$), generating CO_2 from the chemical reaction, and the fuel combustion for heat to drive the reaction. Solutions of dissolved CO_2 are highly acidic and carbonate precipitation occurs faster under basic conditions.
Availability and Suitability of CO_2	Any purification of the CO_2 increases production costs, decreasing the economic competitiveness of mineral carbonation products. Transportation costs are typically minimized by locating production as close to raw materials and the market as possible, which needs to be expanded to include CO_2 for mineral carbonation processes.
Cost and Abundance of Raw Materials for Mineral Carbonation	Portlandite is a key alkaline earth precursor for carbonation. Its primary source is the roasting of limestone, which adds significant cost. Alternative precursors for carbonation face cost issues like transportation costs, tradeoffs between cost and purity, and availability.
Construction Codes and Standards	A mineral carbonate-based construction product with superior performance will not be enough to ensure commercial application. Construction industry standards are slow to change and often established by jurisdictional means, creating fragmented compliance and acceptance standards. Construction standards/acceptance also generally prescribe the composition of materials that can be used, rather than performance standards.

12.3.3 CHEMICAL CONVERSION OF CARBON DIOXIDE

Fundamentally, CO_2 is one of the most stable known compounds. The bonds linking the carbon and oxygen atoms in CO_2 are strong, resulting in a molecule that is chemically and thermodynamically stable. Consequently, significant amounts of energy must be added to make CO_2 react. One means of introducing this energy is to use energy-intensive reactants. The first challenge with these energy-intensive reactants is selectivity associated with poor reaction control of a reactive reagent. The second challenge is these energy-intensive reactants are currently commercially produced from fossil fuels and possess a substantial CO_2 footprint. Successful CO_2 utilization via chemical conversion includes reducing and/or eliminating the use of nonrenewable energy and substantially reducing greenhouse gas (GHG) emissions compared to existing technologies in a process that is economically competitive. Current commercial processes that use CO_2 as a reagent are listed in Table 12-6. A list of chemical products that can be made using CO_2 and a summary of related R&D is provided in Table 12-7.[12] At this time, none of the CO_2-based process technologies for these products are competitive with existing non-CO_2 production technologies at commercial scale, but some are being demonstrated at small scale, which might be compatible with a modular gasification process.

Table 12-6. Current commercial production generated with CO_2

PRODUCT	ANNUAL PRODUCTION (KILOTONS/YEAR)
Salicylic Acid	90
Urea	171
Cyclic Carbonates	80
Polycarbonate	600
Polyether Carbonate	10

Table 12-7. Emerging chemical production with CO_2

CHEMICAL	R&D SUMMARY
Methanol	• Primary focus is on direct hydrogenation of CO_2 with H_2. • Catalytic routes are addressing low selectivity and water (co-produced) deactivation. • Electrochemical routes are at a more fundamental research stage.
Dimethyl Ether	• Commercial production will leverage advances in methanol production and convert methanol into dimethyl ether using conventional process technology.
Formic Acid	• Catalytic research is reducing base consumption, reducing catalyst costs, and finding effective separation techniques for formic product. • Electrochemical routes are working on catalyst stability and durability and an increase in energy efficiency of separation/purification of formic acid from aqueous solutions.
Methane	• Direct conversion of CO_2 and H_2 into methane, the Sabatier reaction, is catalyzed with nickel-based catalysts. • Electrochemical production routes have also been identified. • Current availability of cheap natural gas inhibits further development.
Carbon Monoxide	• Haldor Topsøe currently markets a high-temperature solid oxide electrolysis cell producing high-purity CO approximately 12 normal cubic meters (Nm^3)/hour. • Lower temperature electrochemical processes are at a fundamental research stage.

Table 12-7. Emerging chemical production with CO_2

CHEMICAL	R&D SUMMARY
Ethylene (and ethanol)	• Research has shown electrochemical production of ethylene and ethanol, but deactivation of the electrocatalyst is rapid (less than two hours currently) and CO_2 loss to the electrolyte is too high.
Diphenylcarbonate	• An indirect route for diphenylcarbonate using n-butanol and CO_2 to produce di-n-butylcarbonate, which is reacted with phenol to generate diphenylcarbonate and n-butanol (recycled) has been pilot plant tested for more than 1,000 hours producing 1,000 ton/day. • Direct synthesis of diphenylcarbonate from phenol and CO_2 has been demonstrated, but scaleup is limited by market for the propylene glycol byproduct.
Polymers (alternating, which includes one molecule of CO_2 followed by an epoxide)	• Empower Materials and Econic currently produce alternating polymers with CO_2. However, their low glass transition temperature limits their applications. Challenges include decomposition of cyclic carbonates, high reactivity with water, and a limited number of suitable epoxides.
Polyether Carbonates (ether linkages between two adjacent ring-opened epoxides)	• Relatively mature process technology and a wide range of applications already exist. • Novomer and Covestro produce and market polyether carbonates. • Current research is evaluating epoxides from renewable sources and copolymerization to form polycarbamates and oxetanes.
Carboxylic Acids	• Production challenges are poor reaction thermodynamics, which requires promotion with base, and overcoming the difficulty of activating the C-H bond using highly reactive and resource intensive reagents. • R&D is currently at a fundamental stage.
Hydrocarbons (fuels)	• Current Fischer-Tropsch, iron-based catalysts are being modified for production of heavier hydrocarbons. • Electrochemical approaches are investigating the fundamental chemical mechanism for carbon chain growth. • Reliable, inexpensive, carbon-free hydrogen generation would benefit this application.
Carbon Nanotubes	• Carbon nanotubes and carbon fibers currently produced from CO_2 do not have the properties required for commercially accepted products. • A breakthrough could result in a disruptive commercial technology.

12.3.4 BIOLOGICAL CONVERSION OF CARBON DIOXIDE

The final category of CO_2 utilization is the use of biological systems that consume CO_2. One of the advantages of biological CO_2 conversion is that these systems can accept gas mixtures with low CO_2 concentrations and tolerate/use impurities that may be present. One of the reasons these biological systems consume CO_2 is as a carbon source to support health and growth. As a result, a portion of the carbon in the CO_2 is converted into biomass. Partial conversion of the CO_2 to biomass reduces yield of the target product, which increases the size of the biological system needed to achieve a target production level, makes separation of the biomass from the product a critical downstream process, and makes finding/extracting additional products from the biomass important for improving process economics. Biological systems for CO_2 utilization are like current agricultural processes for producing crops and animals, but biological CO_2 systems are more efficient in land and water use and protein production. Biological systems are also extremely flexible with feedstocks and environments, allowing them to use less desirable land and reduce competition with crops. It also makes them more tolerant to the use of wastewater and waste process gases with no or limited purification. However, project developers should note that biological CO_2 conversion systems are less efficient than commercial chemical processes for the conversion of fossil fuels into value-added products, which results in a larger footprint for biological systems and poorer economic performance.

One of the primary groups of biological CO_2 conversion systems is based on photosynthesis, where solar energy provides the energy for chemical conversions. The challenge of distributing sunlight to the biological system and the low energy efficiency of photosynthesis increases land use requirements for photosynthetic systems. Photosynthetic biological systems for CO_2 utilization include algae and cyanobacteria; additional information on these photosynthetic CO_2 utilization processes in included in Table 12-8.[12]

Table 12-8. Summary of algae and cyanobacteria photosynthesis-based systems

ORGANISM	ADVANTAGES	DISADVANTAGES
Algae	• More productive than contemporary agricultural systems. • Potential product slates include a wide range of biofuels, specialized chemicals, dietary proteins, and food additives.	• Separation cost for the primary product. • Recovering additional value-added products improves process revenue, but also requires more separation processing. • Land-intensive due to low-light utilization efficiency. • Lack of hydrogen infrastructure to hydrotreat the algae's lipids into commercial biofuels. • Genetic modification technology is less advanced for algae than other organisms.
Cyanobacteria	• Two to four times more efficient at photosynthesis than plants. • More amenable to genetic manipulation than algae, allowing adaptation to a wider range of products. • Land use for cultivation does not compete with food crops.	• Low growth rate. • Their genetic manipulation is less advanced than for yeast and plants. • Limited number of species adapted for practical use. • Current productivity of products is too low to be commercially viable.

Non-photosynthetic biological systems are also capable of CO_2 utilization. These systems do not suffer from the inefficiencies of photosynthesis and offer a wide range of available organisms and potential chemicals for production.[12] The tradeoffs for these biological systems include:

- Some form of energy must be fed to the system, which comes primarily from the oxidation of reduced inorganic compounds.
  - Some biological systems can accept low-cost electrons from an electrochemical cell as a source of energy.
- Current production rates are limited as little or no genetic manipulation expertise has been developed for these organisms.
- Early levels of development compared to photosynthetic biological systems.

One benefit of non-photosynthetic biological systems is the ability to operate in the absence of sunlight, enabling continuous operation. Another potential advantage is the limited need for the expensive carbohydrate feed required by similar biological processes like fermentation. The realization of these benefits continues to drive R&D for these biological systems.

Integration opportunities for biological CO_2 utilization in modular gasification systems include the need for co-location of the CO_2 source and its utilization to avoid transportation of the CO_2 byproduct and the potential to use CO_2 with lower purity at near atmospheric operating pressures.

12.3.5 CARBON DIOXIDE STORAGE

In the absence of a value-added, commercial use of the captured CO_2, any captured CO_2 will need to be safely and permanently stored underground inside a compatible sedimentary rock (e.g., sandstone or limestone)[13] or a volcanic basalt[14] formation. Project developers can reference NETL's Carbon Storage Program website for more information.* The website includes technology area/technology component descriptions, Best Practices Manuals (BPMs), and project information.

12.4 SUMMARY

The selection and design criteria for CO_2 capture for modular gasification systems are dependent on target application and products being produced, the regulatory environment, economic constraints, and local niche economics and markets.

The most influential factor on selection and design criteria is the targeted product of the modular plant: chemicals or power. For chemical production, operating constraints for the downstream chemical conversion processes require shifting of the CO-to-H_2 ratio in the syngas and effective CO_2 removal to optimize the performance of the chemical conversion process. For power production, turbine specifications and performance goals dictate the design constraints for any pre-combustion CO_2 capture process. A summary of the differences in the selection and design criteria used for chemical and power applications is included in Table 12-9. Compression of the CO_2 byproduct is included as a selection criterion, as most applications for CO_2 utilization and/or storage require a high-pressure CO_2 stream and most conventional CO_2 separation technologies generate CO_2 at less than or equal to 150 psia.

Table 12-9. Selection and design criteria for chemical and power applications

CRITERIA	APPLICATION	
	CHEMICALS	POWER
Amount of CO_2 Capture	Maximizing chemical production also maximizes the conversion of carbon in the feed into carbon in the product. For products with higher H-to-C ratios, more of the carbon in the syngas must be removed as CO_2, which affects size and complexity of WGS process.	The addition of CO_2 capture reduces energy efficiency and increases cost of power production.
CO_2 Capture Technology	Rectisol® is the standard commercial option for CO_2 capture, as it is capable of simultaneous removal of sulfur (H_2S and COS), CO_2, and other contaminants like hydrocarbons, NH_3, and HCN at high efficiency. Smaller-scale applications may allow for alternative technologies. Design criteria include optimizing the integration of the CO_2 capture for energy efficiency, capital and operating costs, and subsequent processing of the CO_2 product.	The standard commercial approach for simultaneous syngas cleanup and CO_2 separation is the Selexol™ process. Selexol™ is preferred over Rectisol®, because Selexol™ can achieve suitable effluent concentrations more efficiently than Rectisol®. The design criteria discussed for chemical applications are important to make up for the increase in power production cost with CO_2 capture.
CO_2 Compression	The amount of compression for the CO_2 product will be set by the end use (e.g., CO_2 utilization process or storage specifications).	The same criteria apply as described for chemical applications.
CO_2 Utilization	The largest current use for captured CO_2 is EOR. Other smaller commercial uses include beverage carbonation, pneumatic machinery, metallurgical applications, and as a flame retardant. Emerging technologies, which may be extremely size- and risk-compatible with modular plants, include mineral carbonization and chemical/fuel production via chemical or biological routes.	The same criteria apply as described for chemical applications.

---

* Available at *https://netl.doe.gov/coal/carbon-storage*.

After capturing the CO_2, the project developer should identify a suitable application for the captured CO_2. The most promising, current use of CO_2 is EOR. Most existing commercial plants that already include CO_2 capture and are located near an EOR site, or a pipeline connected to EOR sites, have taken advantage of this opportunity. The wide-scale use of CO_2 captured from power plants for EOR is dependent on the capital investment and pipeline infrastructure needed to facilitate transportation. Although there are large R&D efforts focused on developing commercial processes for CO_2 utilization, current commercial CO_2 utilization processes are limited to consuming less than 900 kilotons of CO_2 per year.

With these conditions, CO_2 storage in geologic formations seem to be the most viable solution for CO_2 capture from large stationary CO_2 sources. However, one of the inherent advantages of modular plants is the ability to be sized to take advantage of local niche opportunities and markets. Similarly, some CO_2 utilization R&D efforts are being tested at pilot-scale. Modular plants can take advantage of this R&D development and employ copies of a similar size and/or provide opportunities for pilot-scale testing of emerging CO_2 capture and/or utilization technologies. Project developers should evaluate potential local niche opportunities for CO_2 utilization as part of the initial feasibility analysis of any modular plant design, which could add significant economic incentive for the modular plant.

Endnotes

1 NETL (May 2011) "Cost and Performance Baseline for Fossil Energy Plants Volume 3a: Low Rank Coal to Electricity: IGCC Cases", DOE/NETL-2010/1399.

2 DOE/NETL (August 2015) Carbon Dioxide Capture Handbook, DOE/NETL, *https://www.netl.doe.gov/sites/default/files/netl-file/Carbon-Dioxide-Capture-Handbook-2015.pdf*.

3 G.T. Rochelle (2020) "PZAS at NGCC Conditions", National Carbon Capture Center – Carbon Capture Project 4.

4 C. Dhoke, S. Cloete, S. Krishnamurthy, H. Seo, I. Luz, M. Soukri, Y. Park, R. Blom, S. Amini, A. Zaabout (2019) "Sorbents screening for post-combustion CO_2 capture via combined temperature and pressure swing adsorption", Chemical Engineering Journal 380.

5 Nexant, RTI International, and Aerojet Rocketdyne (2016) "Integration of Aerojet Rocketdyne and RTI Advanced Gasification Technologies for Hydrogen-Rich Syngas Production Task 2: Techno-Economic Analysis for IGCC Application", Technical Report submitted under DOE/NETL Cooperative Agreement DE-FE0012066 RD&D.

6 V. Gupta, P. Sharma, D. Denton, and B. Turk (2016) "Benefits of Integration of Aerojet Rocketdyne and RTI Advanced Gasification Technologies for Hydrogen-Rich Syngas Production Task 5: Techno-Economic Analysis for Coal-to-Liquids via Fischer-Tropsch Synthesis", Technical Report submitted under DOE/NETL Cooperative Agreement DE-FE0012066.

7 G. Howe, J. Albritton, D. Denton, R. Gupta, and B. Turk (2018) "RTI Warm Syngas Cleanup Operational Testing at Tampa Electric Company's Polk 1 IGCC Site", Final Technical Report submitted under DOE/NETL Cooperative Agreement DE-FE0026622/.

8 R. Zolandz and G. K. Fleming (1992) Gas Permeation in Membrane Handbook, W. S. W. Ho and K. K. Sirkar, eds., New York: Springer US, p. 17.

9 M.K. Verma (2015) "Fundamentals of Carbon Dioxide-Enhanced Oil Recovery (CO_2-EOR)—A Supporting Document of the Assessment Methodology for Hydrocarbon Recovery Using CO_2-EOR Associated with Carbon Sequestration", U.S. Department of the Interior: U.S. Geological Survey, Reston, VA, Report No. 2015-1071, *https://pubs.usgs.gov/of/2015/1071/pdf/ofr2015-1071.pdf*.

10 World Business Council for Sustainable Development (2009) "The Cement Sustainability Initiative", *https://www.wbcsd.org/Sector-Projects/Cement-Sustainability-Initiative*. Accessed October 10, 2018.

11 The Global CO_2 Initiative (November 2016) "Carbon Dioxide Utilization (CO_2U): ICEF Roadmap 1.0", *https://www.icef-forum.org/platform/upload/CO_2U_Roadmap_ICEF2016.pdf*.

12 National Academies of Science, Engineering, and Medicine (2018) Gaseous Carbon Waste Streams Utilization: Status and Research Needs, Washington, DC: The National Academies Press. doi: *https://doi.org/10.17226/25232*.

13 D. Daly (2018) "Energy and CO_2 Management: Carbon Capture and Storage", 2018 Lignite Education Seminar. Plains CO_2 Reduction (PCOR) Partnership: Bismarck, ND. *https://cdn.lignite.com/wp-content/uploads/2018/06/12-CO_2-Daly-FINAL-DA.pdf*.

14 D. Nield (November 23, 2016) "Scientists are Trapping CO_2 and Turning It into Stone." Science Alert. *https://www.sciencealert.com/scientists-are-again-turning-co2-into-solid-rock-to-fight-climate-change*.

13.0 GUIDELINES FOR THE DESIGN OF GASIFICATION-BASED ANCILLARY SYSTEMS

13.1 INTRODUCTION

Chapter 13 provides guidelines for the design of gasification ancillary systems and insights into other key supporting elements for consideration during the design of a gasification-based system.

13.2 ANCILLARY SYSTEMS EQUIPMENT AND SOFTWARE REQUIREMENTS

In addition to the gasification and power generation equipment, there are several ancillary systems that are part of the plant design. Without these systems, such as instrument air, inerting systems, flares, cooling water, water treatment, plant controls, material and product handling, the gasification plant would not operate. Although these ancillary systems are critical to normal operation of the gasification plant, the operation and required capacity of these ancillary systems must also support non-standard operations like startups and shutdowns, which may exceed the typical requirements for normal operation.

13.2.1 STARTUP AND SHUTDOWN

Startup and shutdown are the processes of bringing equipment into and out of service. Industry studies have been conducted that determine that many process safety incidents occur during startups, shutdowns, and other cases that infrequently occur. Periods of unusual operation, such as startup and shutdown, involve procedures that employees are not as familiar with, as well as operating conditions that may rapidly change. Facilities should employ effective communication; provide workers with appropriate training; and have strong, up-to-date policies and procedures in place for hazardous operations like startups and shutdowns to prevent accidents.[1]

Startup is the initial procedure of a plant that allows it to reach normal operating conditions. The two types of startup are dry/cold startup of the empty equipment and startup of the plant after some maintenance due to a partial shutdown. Shutdown is the final procedure of a plant that ceases the production and takes the plant to safe atmospheric conditions for inspection and maintenance. The three types of shutdown are total, partial, and critical (also known as emergency) shutdown. Total or partial shutdown may be necessary to perform routine maintenance, while critical shutdown can occur due to malfunction, trouble, fault, or an abnormal situation or accident.

Startup and shutdown are more labor intensive than normal operation and pose a greater risk to operating staff and equipment. Activities must be carefully planned to allow for adequate staffing and safe operation. Maintenance activities performed during a plant shutdown should also be carefully planned to minimize down time.

Prior to system startup, plant personnel should walk down the system to verify that the equipment is ready for operation. The position of the system's valves and breakers should also be verified prior to startup. All Lock Out/Tag Out (LOTO) tickets should be resolved per the plant's startup and commissioning procedures.

13.2.1.1 ANCILLARY SYSTEMS STARTUP

Several ancillary systems must be available prior to starting the plant. These ancillary systems include the plant control system, electrical system, instrument air system, inerting system, flare system, and cooling water systems.

13.2.1.2 PLANT CONTROL SYSTEMS

The plant control system uses electronic means to manage the level of automation of plant startup processes. The level of automation during startup varies from facility to facility and may require more operator input than normal operation. Adequate field personnel must be provided during startup for safety. When startup is fully automated, the distributed control system (DCS) ensures that all startup/shutdown operating procedure steps are followed with proper permissive hold points for operator intervention. Process control systems are discussed further in Section 13.2.2.

13.2.1.3 ELECTRICAL SYSTEM ENERGIZATION

Electric power must be available for all system equipment. Proper personal protective equipment (PPE) should be used and energization should be performed in accordance with the arc flash hazard assessment recommendations, along with the affixed safety labels generated from the study. An arc flash is an electrical explosion due to a fault condition or short circuit and can occur when workers are working with live electrical equipment.

Safety related work practices should be followed in accordance with the National Fire Protection Agency (NFPA) 70E and the plant safety policy, both of which are necessary for practical safeguards of employees and equipment. Field staff should not work on energized electrical equipment within a limited approach or arc flash boundary unless it falls into one of the following categories:

- When de-energizing the equipment introduces additional hazards or increased risk.
- When it is not feasible to de-energize the equipment due to equipment design or operational limitations, including some startup, troubleshooting, or performance testing.
- When the supply voltage is less than 50 volts and it is determined that there will be no increased exposure to electrical burns or to explosions due to electrical arcs.

Only qualified personnel should perform tasks like testing, troubleshooting, and voltage/current measuring within a limited approach boundary. The operations team should ensure the establishment of electrically safe working conditions wherever necessary. They should also identify, document, and implement the LOTO procedure to safeguard employees from exposure to electrical hazard.

Risk assessment plans should identify the hazard, assess the risk, and implement the risk control measures. The safety program should include a job briefing, job hazard analysis (JHA), and training to establish safe working conditions.

Temporary protective grounds, discharging devices, wireless-non-contact phasing sticks, wireless remote clamp-on meters, remote racking devices, and other necessary safety tools and test instruments should be utilized to protect employees from shock hazards.

13.2.1.4 INSTRUMENT AIR SYSTEMS

Instrument air is needed during startup and shutdown for the operation of control valves and other process equipment. The instrument air system should be in operation and at adequate pressure prior to startup of the systems, as it is used to control automated valves. Dryers should be fully regenerated and the wet air tank condensate should be purged. Instrument air systems are further explained in Section 13.2.3.

13.2.1.5 INERTING SYSTEMS

Inerting systems (see Section 13.2.5) are needed to purge out unsafe gases during shutdown (for maintenance) and before startup. Nitrogen, carbon dioxide (CO_2), and/or steam are mostly used in plants for inerting.

13.2.1.6 COOLING WATER SYSTEMS

Cooling water must be available prior to system startup to prevent damage to the equipment. Adequate corrosion inhibitor and glycol (if required) levels should be established prior to startup. Cooling water systems are further explained in Section 13.2.4.

13.2.1.7 COOLDOWN

Processes that operate at elevated temperature must be allowed to cool after shutdown if maintenance activities are to be performed for personnel safety and thermal destressing. This process is referred to as cooldown. Cooling water should be maintained until any required cooldown is completed.

13.2.1.8 VENTING AND DRAINING

If equipment requires maintenance during shutdown, it may need to be vented or drained. Venting (i.e., the release of gas pressure) or draining (i.e., the release of trapped liquid) is performed according to the equipment manufacturer’s recommendations prior to maintenance activities.

13.2.1.9 EQUIPMENT LINEOUT AND LOCK-OUT/TAG-OUT

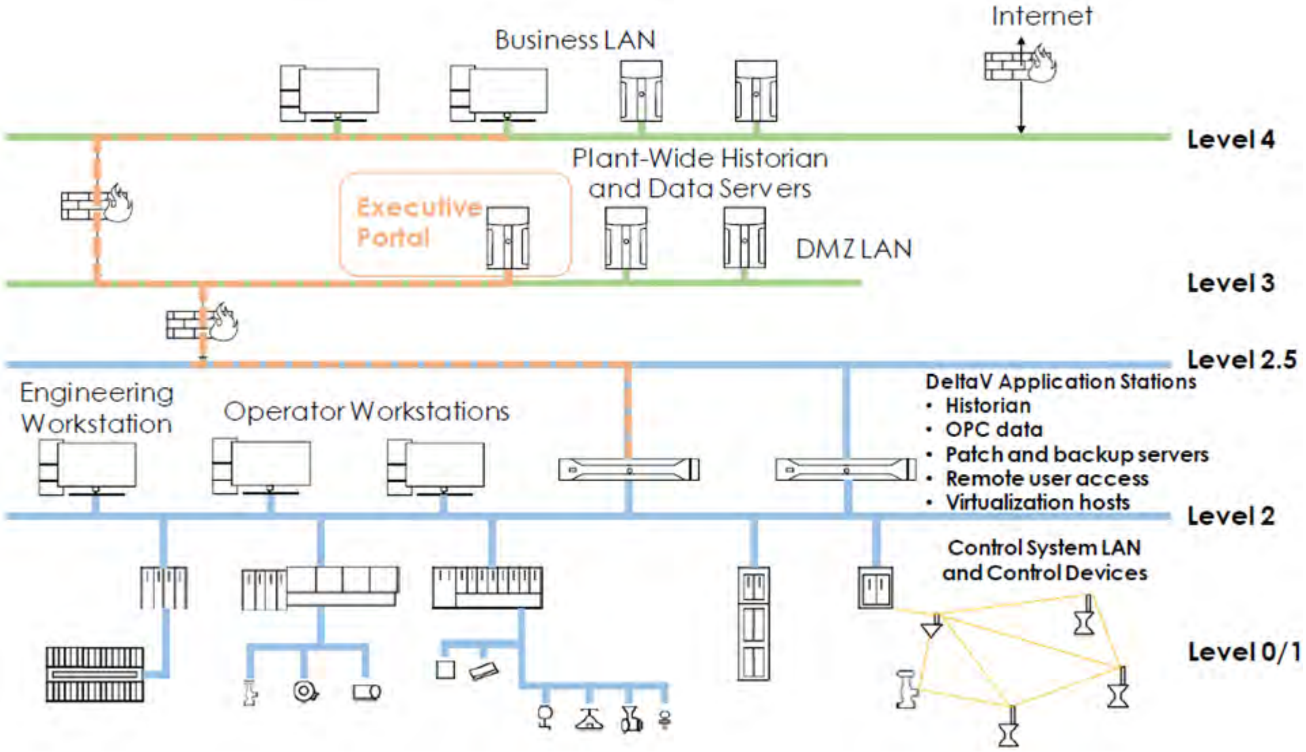
Prior to any shutdown maintenance activities, plant personnel should walk down the system to verify that the equipment has properly shutdown. The position of the system’s valves and breakers should also be verified after shutdown. Isolation of any equipment, valves, or breakers should be addressed using the LOTO procedures.

13.2.1.10 GENERAL ANCILLARY SYSTEMS

Beyond the ancillary systems associated with startup and shutdowns, there are ancillary systems that are required to support operation of the plant or plant personnel. These ancillary systems include water and wastewater treatment; laboratory facilities; and inventory, material, and product storage facilities. In addition to these systems, the project developer will have to include general safety and support infrastructure, like fire and alarm system, site drainage system, and required facilities to support personnel operation, including a control room and offices. This last group of systems is not discussed in detail as these systems are typically required for any business or plant.

13.2.2 PROCESS CONTROL SYSTEMS

Process control systems include a programmable logic controller (PLC), a DCS, and a supervisory control and data acquisition (SCADA) system. A typical schematic architecture of a modern process control system is shown in Figure 13-1.



Used with permission from Emerson[2]

Figure 13-1. Representative process control system architecture

A PLC is a remote, computerized system adapted for use in harsh, industrial environments for the purpose of monitoring and controlling a specific system, such as a wastewater treatment plant, burner control system, or emissions monitoring system. PLCs are advantageous, because they can be easily reprogrammed and quickly react to changing process conditions.

A DCS is a system in which multiple controllers are distributed throughout the plant and linked to a central, integrated, supervisory control center. Each individual DCS controller (or pair of redundant controllers) monitors the performance parameters of an individual section of the plant and reports findings to a remote, centralized computer controller. Each DCS module independently operates, making for an easier, less-computationally-intensive, decision-making process. A DCS can be used to enhance production efficiency by improving response times. The decentralized nature of the DCS also improves plant reliability (i.e., a failure in a single central digital controller would result in a complete shutdown of all operations, while an individual controller in a DCS only has dominion over a single plant area, so the rest of the plant may continue to function during repair of the one control unit). A DCS is best suited to larger, more complex tasks due to its distributed nature.

A SCADA system is like a DCS in that it uses discrete control units to monitor individual plant components. Unlike a DCS, SCADA systems (which can also include PLCs) are mainly used for data gathering and analysis, while DCSs focus on the real-time control over the whole plant. Most industrial plants utilize some combination of these systems as part of an overall control strategy. The plant manager is responsible for choosing and implementing an overall monitoring and control structure that best suits the plants' needs.

The human machine interface (HMI) is the interface between the process control system and the plant operators that allows operators to control and monitor plant processes from the control room. A typical control room has multiple plant operator HMI high-resolution displays located on one or more operator consoles, with large wall or ceiling-mounted overview displays for at-a-glance status and alarm indication. HMI displays are the interface to the main DCS and allow monitoring, control, and adjustment of facility operating conditions, utilizing state-of-the-art optimization of information for efficient presentation of critical information. HMI displays are networked using virtual workstations from main servers, requiring less maintenance and upkeep for software configuration.

Alarms from the process control system are conditioned with software logic to avoid alarm flooding (i.e., multiple alarms occurring at the same instant, causing stress and confusion to the operator as to the actual source of the main problem), ensure that non-critical alarms are grouped to cover an area of the plant for display on the HMI, and alert the operator to changes in operating conditions only when required. The alarm management system should follow American National Standards Institute (ANSI)/International Society of Automation (ISA) 18.2[3] alarm management guidelines. All critical alarms for each area of the plant that require immediate operator action are individually displayed both on the HMI screens and locally at the equipment requiring attention. Alarms should also be provided to meet the requirements of all applicable fire protection codes.

A modern control room with HMI stations and large information displays is depicted in Figure 13-2. The control room operator consoles also house hardwired emergency stop buttons for critical pieces of equipment, which allows immediate shutdown in the case of equipment failure or safety concerns. The emergency buttons act independently of any control system or logic to ensure dependable shutdown of turbines, pumps and/or motors, fired burner systems, or supply valves.

Used with permission from Data Resources[4]

Figure 13-2. Modern DCS control room

The process control system includes signal input/output (I/O) cabinets throughout the facility for collecting and distributing signals to and from equipment (e.g., motors, pumps), control actuators (e.g., valves), and instrumentation. All cabinets should be in locked areas to ensure safety and security. The cabinets communicate with processors and servers in the control room via a redundant fault tolerant communications network comprised of fiberoptic cables to ensure fast and reliable communications while maintaining security against external viruses, malware, and intrusion attacks. System servers archive data for record keeping, troubleshooting, and data analysis.

The process control system also communicates with third-party control systems of packaged equipment, such as air compressors, motor controls, electrical distribution equipment, water treatment systems, auxiliary boilers, and code compliant gas burners. These independent systems are typically controlled by PLCs. While the PLCs operate in a standalone manner, supervisory control parameters are shared with the main process control system so that the equipment can be started and stopped and major control parameters (e.g., speeds, output demands, process temperatures, pressures, flow rates, chemical analysis readings, and other critical parameters) can be monitored and alarmed. All critical interfaces should be redundant to ensure reliability.

Modern process control systems often communicate with field instruments and devices using FOUNDATION™ Fieldbus[5], Modbus[6], Profibus[7], or Hart[8] communication protocols. In addition to transmitting the primary process measurement, these protocols also provide diagnostic information about the devices and can alert operators about predictive maintenance requirements.

The process control system accepts a wide variety of signal types at multiple voltage levels, ensuring that any necessary instrumentation signals or interfacing signals can be accepted. It can also interface with business systems to share data for accounting, recordkeeping, data report generation, and economic analysis (i.e., SCADA functions). The system is widely expandable, allowing for future growth and modification.

The process control system is programmed to monitor and control plant processes utilizing the software provided by the manufacturer at the direction of the design engineers. All software is fully tested at the factory before shipment to the facility and easily editable with proper credentials for future changes, growth, and enhancements. The programming includes all logic and display configurations necessary for the control of the systems, along with the supervisory parameters and interface to the third-party control systems mentioned above. Plant data can be trended, plotted, formatted into reports, and archived as required. Most modern process control systems are self-documenting, meaning that the graphical configuration of the control algorithms via an engineering workstation depict the actual process control algorithms running in the programmed logic.

13.2.3 INSTRUMENT AIR SUPPLY

The purpose of the Instrument Air Supply (INA) system is to provide clean, dry, oil-free, instrument-quality compressed air to components and processes throughout the plant. The INA system provides the motive force within a modern process facility. Components may include control valves, pneumatically actuated or driven equipment, instrumentation, or general service applications. The INA system can also provide compressed air to a plant Service Air (SVA) system, which consists of utility stations for the maintenance of outage components (e.g., pneumatic tools) or to blowdown/clean equipment components. SVA system air is not conditioned to the same extent as INA system air. Connecting equipment requiring clean air to an alternative air source like the SVA system can result in additional equipment maintenance or premature equipment failure. Also, the SVA supply is typically fed from the same compressors as the INA, so it is recommended to have a pressure regulator shutoff on the SVA feed to prevent starving critical INA systems before the INA pressure falls below a minimum level. A typical flow path for an INA system is shown in Figure 13-3.

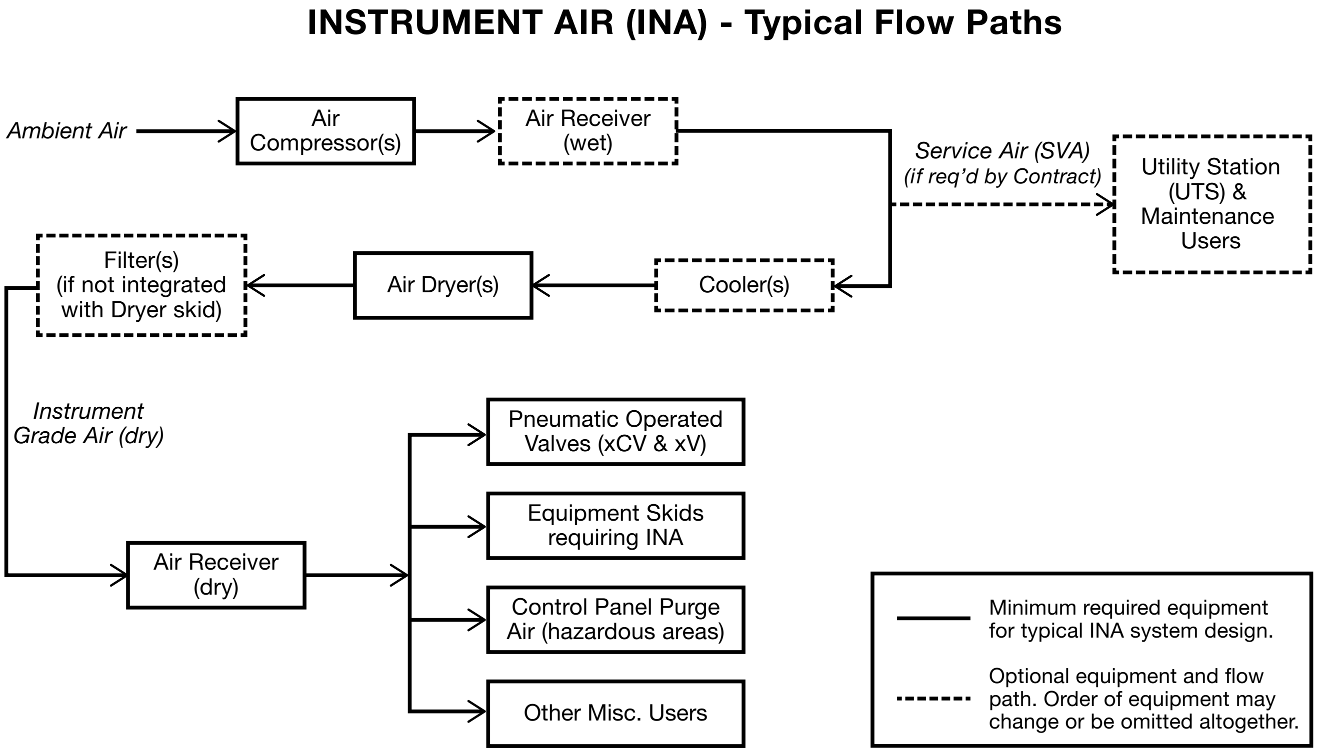


Figure 13-3. Example layout of instrument air system

The INA system includes four primary types of equipment: oil-free air compressors, air receivers, desiccant air dryers, and air filters. The primary process flow path begins with the air compressors. The air compressors take in ambient air and pressurize it, sending it to a wet air receiver. The wet air receiver acts as a pulsation dampener and condensate knockout drum. From the wet air receiver, the compressed air is filtered for particulates and oil and forwarded to the desiccant air dryer skid. Once the air has been dried, it is sent to the dry air receiver, which is a large storage reserve tank for compressed air. From the dry air receiver, the air is distributed throughout the plant through the INA piping system. Proper instrument air dryness is required to prevent damage to instruments or valves. The system should ensure redundancy of the desiccant air dryers and have proper alarms for when air moisture content approaches the dew point limit. An example compressed air system is shown in Figure 13-4.

Used with permission from Sullair[9]

Figure 13-4. Skid-mounted compressed air system

Typical piping design includes a distribution header or a loop where sections may be isolated. A distribution header often requires less material, but a distribution loop ensures compressed air is available when parts of the piping system are shutoff or leaking. From the main header loop, compressed air is routed to the necessary components. Critical systems may justify redundant INA supply lines.

When designing the INA and SVA systems, the project developer should consider the cost/benefit of INA loop versus headers, INA protection from SVA drawdown, and component redundancy and monitoring alarms.

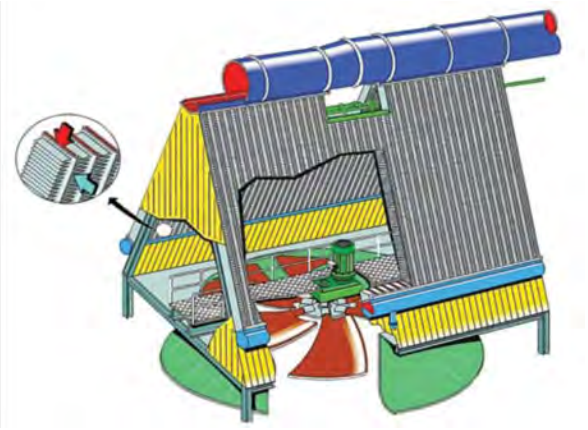
13.2.4 COOLING WATER SYSTEMS

Cooling water is required to reject heat during the gasification and power generation processes. A typical power plant has a circulating water system (CWS) and closed cooling water system (CCW). The CWS is an open loop system that uses evaporation to reject heat to the atmosphere. The CCW system is a closed loop that uses a recirculated cooling water mixture (usually made up of a mixture of propylene glycol and demineralized water treated with a corrosion inhibitor) to reject heat to either the CWS or air using heat exchangers. The cooling water process for a specific piece of equipment is dictated by cooling water quality requirements for the equipment.

In the CWS, heat is rejected to the atmosphere by evaporation using a cooling tower, as depicted in Figure 13-5. Low-pressure steam is condensed in a surface condenser with water as the cooling medium from the CWS.

Figure 13-5. Mechanical draft cooling tower

In locations where sufficient water is not available for evaporative cooling, dry cooling must be employed. For dry-cooled plants, air is used as the heat transfer medium.[10, 11, 12] After exiting the turbine, exhaust steam heat is extracted and sent to the atmosphere by an air-cooled condenser (ACC), as shown in Figure 13-6.



Permission pending from Wurz[10]

Figure 13-6. Air-cooled condenser

The CCW system supplies a cooling water mixture to various equipment components in the plant, such as the rotating equipment, lube oil skids, and turbine generators. The CCW is the fundamental means of rejecting waste heat from the primary process equipment within a modern gasifier facility. The manner of how the heat is rejected is dependent on whether the plant employs wet or dry cooling. For a wet-cooled plant, heat exchangers are installed to cool the CCW with water from the CWS; heat is rejected through a cooling tower. For a dry-cooled plant, heat from the CCW system is directly rejected to the atmosphere through an air-cooled heat exchanger (ACHE).

13.2.5 INERTING SYSTEMS

Inerting systems are used in gasification plants for pre- and post- startup fire protection and passivation of equipment during outages. Inerting is a process where oxygen is purged from a confined space using an inert gas (e.g., nitrogen, CO_2, argon), water, or steam to a level too low to support combustion. Spaces that have been inerted are considered a confined space and subject to additional safety precautions.

Feedstock handling equipment requires inerting at certain points of operation. NFPA 85, Chapter 9.5.4 states, "A pulverizer that is tripped under load shall be inerted and maintained under an inert atmosphere until confirmation that no burning or smoldering fuel exists in the pulverizer, or the fuel is removed." Mill systems are typically inerted during startup and shutdown. Steam is the most used inerting medium for pulverizer systems.[13]

Inerting fire protection systems are used to prevent and stop fires and explosions in feedstock handling equipment using water, steam, or CO_2 as the inerting agent. Low-pressure CO_2 or steam used before pulverizer startup and after each shutdown may extinguish undetected fires and reduce the risk of explosion. The injection of water, steam, or CO_2 during pulverizer operation is not recommended, as it may lead to plugging in the outlet conduits of the pulverizer. Steam or CO_2 injection is recommended for feedstock bunker fire protection to prevent the level of combustible gas from exceeding the lower combustion limit (LCL).[14] The feedstock bunker or silo design must incorporate any required fire protection systems, including bulk storage of CO_2 and any required injection equipment and piping.[15]

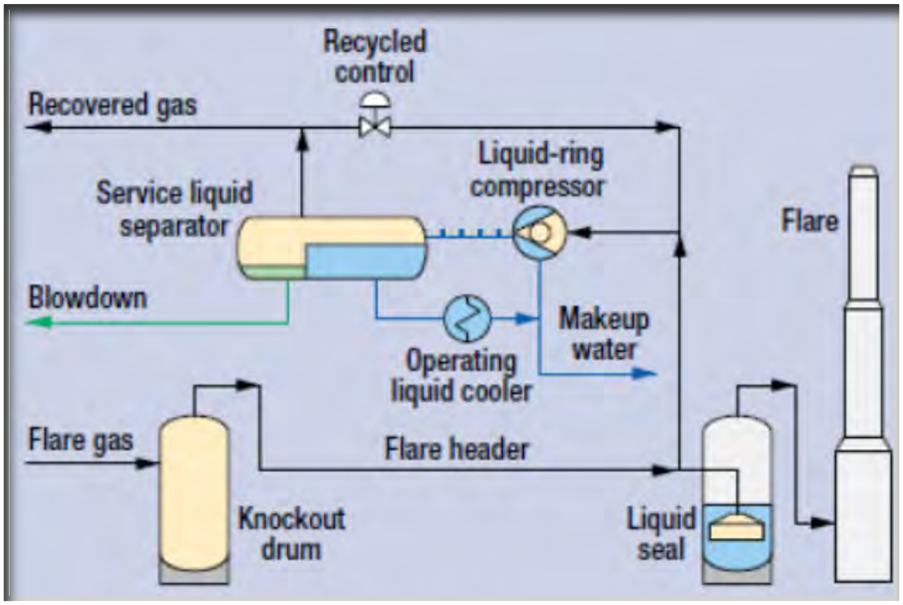
Inerting is also used to prevent corrosion damage during extended shutdowns. Nitrogen is added to a vessel's head space to protect against tank corrosion, chemical oxidation, and fire or explosions. Nitrogen purge systems rely on the availability of nitrogen, which can be supplied from high-pressure tanks, delivered as a liquid from cylinders or bulk storage tanks, or produced from a nitrogen generator.[16] Both membrane and pressure swing adsorption (PSA) nitrogen generators are available to produce nitrogen at up to 99.9% purity. Nitrogen can also be provided by the gasification plant's air separation unit (ASU), if one is present.

When designing the inerting system, the project developer should consider facility fire protection needs; inerting gas availability, cost, safety, and environmental impact; shutdown and outage corrosion prevention needs; and system complexity, monitoring, and control.

13.2.6 FLARE SYSTEMS

The U.S. Department of Energy (DOE) defines natural gas flaring as the controlled combustion of natural gas for operational, safety, or economic reasons.[17] Flaring for operational and safety reasons is done to prevent overpressure in piping and vessels, frequently by burning off flammable or combustible gases released from pressure safety valves. Flaring often occurs during startup and shutdown. Flaring for economic reasons is done to burn off gas or liquid in the absence of demand or adequate infrastructure to gather and distribute.

The American Petroleum Institute's (API) Standard 521 outlines the requirements for the design of flare systems. A flare system typically contains a flare stack, piping that delivers the gases to be flared and the fuel to support the flare, and a knockout drum. The flare stack contains the flame tip, which controls air flow and has seals to prevent flashback. The knockout drum is located at the base of the flare stack to collect condensed liquids from the gas passing to the flare. One or more flares may be required at a process location. Reducing the amount of gas flared has both economic and environmental benefits. A variety of strategies exist to minimize flaring, including understanding the conditions when the gas produced must be flared, proper maintenance of equipment and piping, collection and sale of waste gases, and recycling waste gases back into the process. Flare gas recovery units (FGRUs) are used to collect and use gas that would otherwise be flared. These are typically found in refinery units or other large-scale chemical production facilities where the gas that would be flared can be used as fuel instead. A typical flare configuration with recovery unit is shown in Figure 13-7.



Permission pending from Chemical Engineering Processing

Figure 13-7. Overall flare stack system with recovery unit system

In a typical solids-fed gasification facility, flaring occurs downstream of the gasifier reactor, where syngas is sent to the flare. This could occur in the syngas cooling system or any of the downstream units where syngas would be utilized to generate chemicals.

Once the syngas is out of the reactor and scrubber area, it should be free of dust. However, the main emissions of concern for a gasification facility discharging to a flare system are particulate matter, nitrogen oxide (NO_X), sulfur dioxide (SO_2), carbon monoxide (CO), and volatile organic compounds (VOCs). A flare designed to combust syngas should be of a low velocity type due to the low heating value of syngas of around 200 British thermal units (Btu)/scf. Flare stacks are often distanced from occupied spaces to minimize the risk of radiant heat transfer to personnel and equipment when the flare is in operation.

13.2.7 WATER AND WASTEWATER TREATMENT SYSTEMS

Water and wastewater treatment requirements are defined by the inlet water quality available and the discharge water quality needed.

The inlet water varies based on the original source of the water (e.g., river, aquifer) and how it was previously used. These factors impact the content and volume of dissolved and suspended solids in the water source. All contaminants in a water or wastewater source fall into one of two types of constituent, dissolved or suspended solids, and can be subclassified into other categories (e.g., organic, inorganic, volatile, non-volatile).

Water is a universal solvent; it gains constituents through contact with other sources and all constituents concentrate due to evaporation if the water is used for cooling. The high latent heat and availability of water make it an ideal heat transfer fluid, but when pure water evaporates, the remaining constituents in the water concentrate. The constituents originally in the water will eventually precipitate out of the water if pure water is continually evaporated and removed. The first constituent to precipitate is the first limiting constituent and—if that limiting constituent is inhibited or removed—there can be a second, a third, and so on. Cooling towers are operated so that the circulating water is below the water's saturation point for its limiting constituents. The concentration and precipitation of dissolved solids can cause scaling or fouling of equipment and piping.

The limiting constituents should be removed from the water source to prevent scaling and fouling; the ions that need to be removed are specific to the end use of the water. Water purity requirements may arise from various sources, including equipment specifications, limitations placed on a discharge to a body of water, limits set by a municipality or National Pollutant Discharge Elimination System (NPDES) permit, and/or general scaling and fouling limitations. Regardless of the source of the water purity requirements, there is a limitation placed on the effluent quality of the water, which can be managed through the proper selection of water treatment equipment.

A qualified engineering firm should work with qualified vendors to evaluate the site-specific water source and water users to determine the actual site water treatment requirements and optimal treatment scheme for removal of limiting constituents and appropriate water discharge.

13.2.8 LABORATORY FACILITIES

Modern process facility operation relies heavily on accurate monitoring of the physical and chemical properties within the facility's processes. A modern power plant chemical lab is shown in Figure 13-8; a typical chemical sample panel often incorporated within the chemical laboratory is shown in Figure 13-9. Most facilities have an advanced chemical laboratory with an integrated sample panel that imports samples from the facility and continuously analyzes the physical and chemical properties of the various constituents being monitored.

Used with permission from Savant Instruments[18]

Figure 13-8. Modern power plant chemical laboratory

13.2.8.1 WATER AND STEAM MONITORING

Water and steam quality should be continually measured to prevent corrosion and scaling in process equipment. Parameters measured include pH, conductivity, sodium, silica, phosphate, oxygen reduction potential (ORP), and dissolved oxygen. The Electric Power Research Institute (EPRI), American Society of Mechanical Engineers (ASME), and International Association of the Properties of Water and Steam (IAPWS) have published recommendations that further outline water quality and analysis equipment accuracy requirements.[19] Recent IAPWS recommendations are outlined in Table 13-1.

Used with permission from Solutions Incorporated[20]

Figure 13-9. Chemical sample panel

Table 13-1. Summary of minimum key monitored parameter requirements for power plant water analysis[21]

SAMPLING LOCATION		MINIMUM KEY INSTRUMENTATION	CAVEAT
Condensate pump discharge		Conductivity after cation exchange	Not required for plants with air-cooled condensers
		Dissolved Oxygen	
		(Sodium) seawater-cooled plants	
Feedwater (drum and once-through boiler circuits)	Condensate pump discharge	Conductivity after cation exchange	
	Economizer inlet or main feed pump (HRSGs)	Conductivity	
		Conductivity after cation exchange	
		pH	
		Dissolved oxygen	
	Deaerator inlet	ORP	Copper alloys in feedwater circuit
Boiler drum downcomer (preferable) or blowdown	Plants on all-volatile treatment (AVT)	Conductivity	
	Plants on caustic treatment	Conductivity after cation exchange	
		pH	
	Plants on oxygenated treatment	Conductivity	Sample from downcomer
		Conductivity after cation exchange	
		pH	
		Dissolved oxygen	
	Plants on phosphate treatment	Conductivity	$[Na]:[PO_4]$ molar ratio measurement may require additional monitoring
		Conductivity after cation exchange	
		pH	
		Phosphate	
Steam	Saturated	Conductivity after cation exchange	Iso-kinetic sampling is necessary
		Sodium	
	Superheated/ reheated	Conductivity after cation exchange	
		Sodium	
Make-up water plant outlet		Conductivity	Plants with storage tank exposed to atmosphere
		Conductivity after cation exchange	
		Silica	Plants where there is a risk of non-reactive silica or organic contamination of raw water

13.2.8.2 FUEL GAS MONITORING

The cleanliness of fuel gas to a power plant or process should be monitored. Liquid or solid contamination in the fuel gas stream can foul burners or gas turbine combustors. The most common contaminants in pipeline natural gas are moisture, carryover from oil-flooded fuel gas compressors, and pipe scale. Filter and separator vessels with specifically designed internals can be installed to remove entrained liquids and solids in a gas stream.

Gas chromatography can be used to measure fuel gas quality. The gas chromatograph will be installed to measure a small side stream from the main fuel gas line. The chromatograph can be designed to monitor the percentage of each of the gaseous components, as well as the presence of gaseous contaminants like hydrogen sulfide (H_2S), CO_2, and nitrogen.[22]

Fuel gas parameters are also typically measured in real time, including pressure, temperature, flow rate; higher heating value (HHV); lower heating value (LHV); Wobbe Index (calculated interchangeability of fuel gases); and constituent analysis.

13.2.8.3 FUEL ANALYSIS

Fuel properties vary by region, species, and various other factors. Real-time measurement of fuel quality is necessary to verify that the feedstock meets the plant's specifications. Advanced real-time online analyzers are available to provide stable and accurate analysis of calorific value, total ash, moisture, volatile matters, and carbon content.[23] An online laser fuel analyzer is depicted in Figure 13-10.

Used with permission from Lyncis[24]

Figure 13-10. Online laser fuel analyzer

13.2.8.4 FLUE/EXHAUST GAS MONITORING

A continuous emissions monitoring system (CEMS) is a packaged system of gas analyzers, including a gas sampling system and temperature, flow rate, and opacity monitors that are integrated with a data acquisition system to demonstrate environmental regulatory compliance of various industrial sources of air pollutants in an exhaust stream or stack. A typical extraction CEMS cabinet is shown in Figure 13-11.

Used with permission from Monitoring Solutions[25]

Figure 13-11. Extraction CEMS

CCEMS are required on almost all major emission sources in the United States and they monitor primary regulated constituents, including: NO_X, SOX, Hg, CO, CO_2, O_2, unreacted hydrocarbons (UHCs)/VOCs, particulates,and opacity.

13.2.9 MATERIAL AND PRODUCT STORAGE

Making use of available space to store raw material ensures that the needed amount of material is always available onsite when needed, reducing down time and ensuring continuous operation. The two most important material storage needs for a gasification facility are feedstock and catalysts, which are detailed in the following subsections.

13.2.9.1 STORAGE

Small particles (dust) from some fuels can be combustible, can cause flash fires and explosions, and are generally harmful to human health from inhalation, regardless of fuel species. U.S. federal environmental regulation 40 CFR Part 60.254[26] outlines the standards for fugitive dust, opacity, and particulate emissions. Open storage piles require a fugitive dust emissions control plan using one or more of the following control measures to minimize fugitive dust to the greatest practicable extent. The dust control plan should consider the following for individual site conditions:

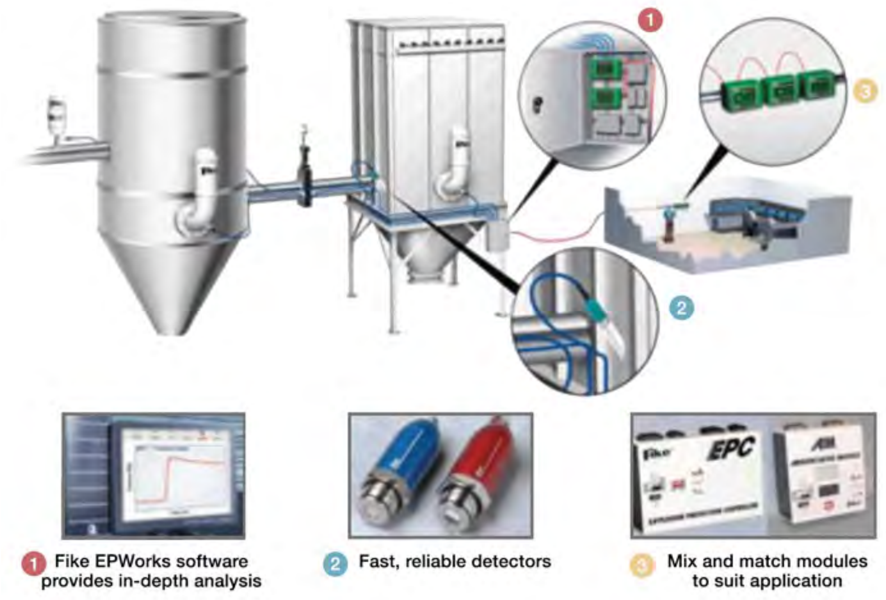
- Locating the source inside a partial enclosure.
- Installation and operation of a water spray or fogging system.
- Application of appropriate chemical dust suppression agents on the source.
- Use of a wind barrier.
- Compaction.
- Use of a vegetative cover.

Ideally, feedstock should be stored in covered structures for protection against spontaneous combustion and cycled or blended (as needed). A dome-covered storage facility with a stacker-reclaimer system at a modern coal power plant in a location subject to high winds and hurricanes is shown in Figure 13-12. The project developer must consider safety, environment, and costs when determining the plant's fuel storage design.[27]

Used with permission from SNPower Dinginin Ltd Co.[27]

Figure 13-12. Storage stacker reclaimer

NFPA Standard 652, the "Standard of the Fundamentals of Combustible Dust," provides the basic principles of, and requirements for, identifying and managing the fire and explosion hazards of combustible dusts and particulate solids.[28] Smaller silos and bunkers can use inerting systems and explosion venting to mitigate risk; alternatively, a more sophisticated explosion suppression system can be installed to stop a deflagration event and avoid endangering equipment or personnel (by circumventing the need to vent combustible dust into adjacent occupied areas). A Fike® explosion protection and control system is depicted in Figure 13-13.



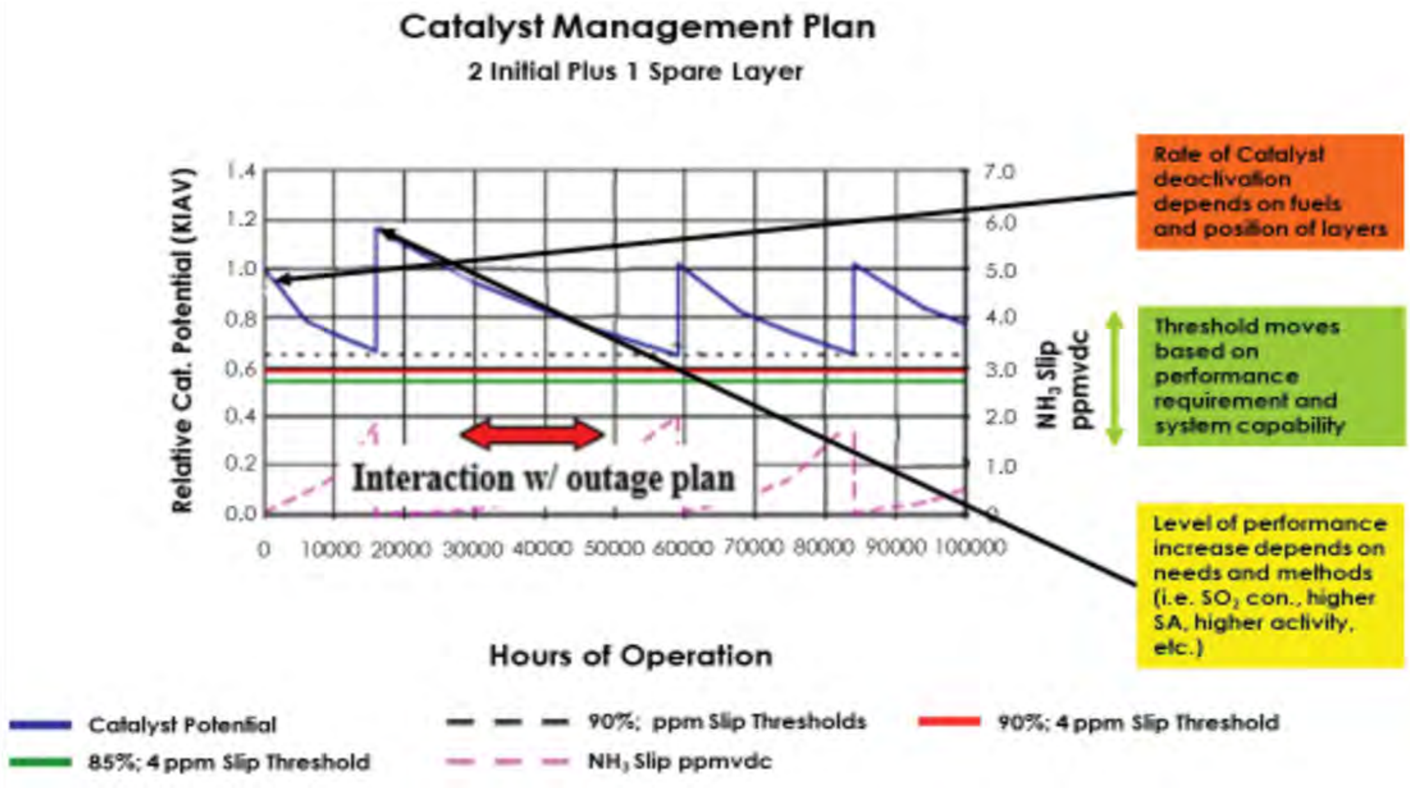
Used with permission from Fike[29]

Figure 13-13. Fike® explosion suppression system

13.2.9.2 CATALYST STORAGE AND MANAGEMENT

Emissions control systems are required when a plant is operating, which emphasizes the importance of outage planning and catalyst replacement.[30] An example is provided for a selective catalytic reduction (SCR) catalyst management plan in Figure 13-14. Operators should work closely with their catalyst provider to implement a proper catalyst management plan that considers the following aspects:

- Projected plant outage schedules and demands.
- Boiler/SCR operations assessment.
- System inspections and field sample data analysis.
- Fuel management and flexibility.
- NO_X performance and SO_2 conversion objectives.
- Cost of NO_X credits and capital budgeting.
- Mercury (Hg) oxidation.
- Catalyst technology advancements.
- Integration with multi-pollutant control (e.g., NO_X, Hg, sulfur trioxide [SO_3]) programs.



Used with permission from Cormetech[30]

Figure 13-14. Typical catalyst management plan

Catalyst vendors typically provide detailed storage and handling procedures for SCR catalysts. For long-term storage (i.e., beyond six months), the relative humidity around the catalyst must be maintained to less than 70%. Failure to meet this condition may void the catalyst warranty.[31] Under no circumstance should the catalyst installed within the reactor be (1) exposed to ambient conditions without dehumidification for a total accumulated period of four weeks per year or longer or (2) subject to water leakage or condensation of very high humidity.

13.3 OTHER SUPPORTING DESIGN ELEMENTS FOR CONSIDERATION

There are additional elements to consider in plant design beyond the plant equipment and systems design. These items include the contracting method, transportation, and process safety.

13.3.1 ENGINEERING, PROCUREMENT, AND CONSTRUCTION CONTRACT

An engineering, procurement, and construction (EPC) contract provides the legal, contractual, and technical specification requirements for the construction of a facility. It is the standard method of contracting larger projects. A major benefit of using an EPC project delivery method is the single integrated schedule for all work to be executed, from design through commissioning. Work can be organized to allow procurement and construction to begin before the design is fully complete. EPC employs a "just in time" supply of engineering for the project, which allows schedules to be reduced and considerably reduces interest during construction (IDC). The risks of any engineering errors or delays fall on the contractor, so instead of delay or cost claims the contractor is obligated to resolve any issues internally without cost to the owner. However, the owner cedes control of the decision making for vendors and defining "perfect" versus "good enough." Lump sum turnkey (LSTK) EPC contracting works best for "fit-for-purpose" plants. The EPC contract is structured with a clear hierarchy, as shown in Figure 13-15.

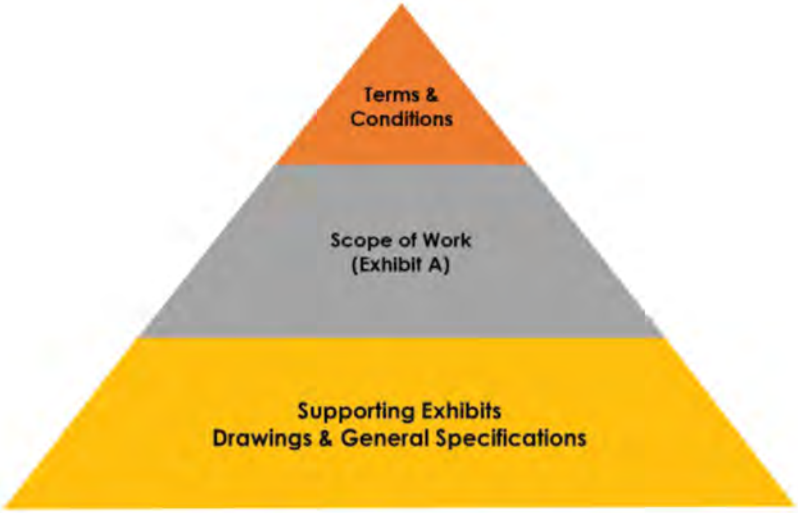


Figure 13-15. EPC governing document hierarchy

The “Terms & Conditions” (T&Cs) of the contract define the respective legal obligations between the owner and the contractor. It is the highest-level governing document in the contract and takes precedence over all subsequent sections.

The “Scope of Work” is the technical specification for the project. It is intended to fully define what the contractor is to provide, along with any other requirements, such as owner specific standards or production capacity, efficiency, or reliability guarantees. Owners sometimes use front-end engineering and design (FEED) efforts to fully develop the scope of work. A well-developed scope of work is essential for obtaining a plant that meets owner requirements.

The “Supporting Exhibits” are typically used to provide additional details. “Drawings and General Specifications” are not considered to be actual design documents, but rather a method to more effectively communicate the scope of work details.

Other contract considerations include:

- Owner/contractor obligations: defines the party responsible for aspects of the project like site access and permitting.
- Limits of liability: impacts the contractor’s risk and, therefore, the risk premium.
- Liquidated damages: establishes compensation to the owner for the contractor’s failure to meet certain obligations, such as capacity, efficiency, or schedule.
- Incentives/penalties: a balanced system of incentives and penalties tied to achieving desired key objectives (budget, schedule, quality/rework, safety, etc.) can help ensure a successful EPC outcome, because it establishes common shared priorities between the EPC contractor and the project developer/owner.
- Risk sharing versus LSTK EPC: overall EPC costs can be reduced if the project developers/owners and/or various sub-component suppliers share some of the overall risk versus an LSTK EPC approach where the EPC contractor assumes all the risk and thus adds substantial contingency costs to the contract.

13.3.2 ENGINEERING, PROCUREMENT, AND CONSTRUCTION AND MODULE SUPPLIER SELECTION

EPC contractor selection should be based upon the reputation, experience, and demonstrated skills of the company. The proposed project management team may be a consideration, as well as reference projects, and references from previous clients. It is recommended that the owner visit the potential EPC contractor’s previous or current projects and converse with the other owners. An EPC contractor with a demonstrated record of on-time and on-budget performance with a stellar safety record is preferred.

Additional considerations should be compliance to EPC contract T&C, scope of work, price, and schedule. It is also important to establish that the EPC contractor is familiar with the project site jurisdictional requirements and labor market. The owner should review the published safety records and financial statement of the EPC contractor and inquire on contractor-initiated change orders on previous projects. It is also important to review the EPC contractor’s quality management plan, document control system, and project management systems.

An EPC contractor that has its own engineering capability and directly hires construction labor may be preferable to one that subcontracts significant portions of the work, because of the additional control over costs and supervision.

A single EPC contracting entity is preferred over a joint venture (JV) arrangement unless the JV parties agree to joint and several liability risk (i.e., each JV member can be held independently liable for the entire amount). If there is a JV arrangement between the engineer and construction entities, the construction entity should be the lead entity that controls the schedule, because construction represents the larger portion of the EPC project cost.

EPC costs are highly contingent on project specifics. A rule-of-thumb rough breakdown for projects costs is shown in Figure 13-16.

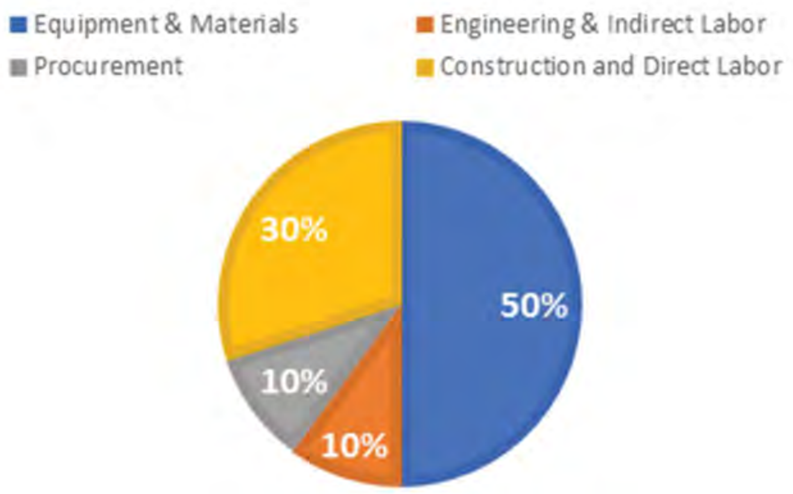


Figure 13-16. Indicative EPC cost breakdown

Module supplier selection is based on reputation and relevant similar module fabrication and client references. Ideally, the primary owner will visit the sites of previously supplied module equipment and confer with the other owners to assess the reliability of delivered equipment, maintainability, and availability of service and spare parts. The owner should visit the factory and assess the quality standards. Additional considerations include schedule, proximity to the site, factory backlog, and quality control (International Organization for Standardization [ISO] 9001 certification is preferred). If the module supplier is integrating components from other suppliers, it is preferred that it has previous experience with those vendors.

The owner should require that module equipment is tested to the fullest extent possible at the factory prior to shipment. Instruments should be calibrated, motors should be bumped, all valves should be actuated, and control valves tested in advance. The owner should also review the shipping and packaging preparation standards to ensure that the modules will not arrive at site damaged or deteriorated.

13.3.3 MODULE TRANSPORT CONSTRAINTS

Component and system module size and weight limitations are constrained by fabrication shop capability, method of transport to the site, site access, and crane capability. The primary transport methods employed are truck/roadway, rail, barge (waterways), ocean marine (including container ships) for globally sourced equipment, and air.

In most cases, some amount of truck/roadway transport will be required and is generally combined with one or more of the alternative transportation modes. When more than one type of transport is used, it is known as intermodal or co-modal and typically involves standard size shipping containers. Each transport method has its own inherent physical and regulatory constraints that must be considered before module procurement. The modules must be designed for the method of transport to prevent damage to the module or failure in the shipment.[32]

13.3.3.1 TRUCK/ROADWAY TRANSPORT

Roadway transportation is the most available form of transportation, providing for door-to-door service. It is easy to track the location of truck shipments and can be used for diverse types of loads. For longer distances, truck transportation becomes more expensive compared to other means. Unexpected delays can be caused by weather, accidents, and traffic conditions. Truck transportation is subject to size and weight limit regulations. For oversize/overweight transportation, special permits must be obtained.

13.3.3.2 RAIL TRANSPORT

Rail transport is economical if there is adequate time to schedule and reasonable access to the rail head. Rail transportation can consolidate shipments requiring multiple trucks onto a single train, reducing transportation costs.[33] Trains also burn less fuel than road vehicles on a weight-per-mile basis. Most module rail transport utilize standard shipping containers that can be transferred between barges, rail, and trucks (Figure 13-17).

Used with permission from Inwasty[34]

Figure 13-17. Shipping containers on rail passing through the San Marco area of Jacksonville, Florida

Twenty-foot and 40-foot containers are most common in U.S. shipping. Common containers are 8.5 feet tall; high cube containers are 9.5 feet tall

13.3.3.3 BARGE TRANSPORT (WATERWAYS)

The United States has a large inland waterway transportation system with a network of interconnected rivers and canals. More than 1 billion tons of cargo are annually moved via barge[35]. Barge transport is convenient for larger modules where there is convenient shipping dock access.

13.3.3.4 OCEAN TRANSPORT (CONTAINER SHIPS)

Container ships are a type of cargo ship with all the load comprised of truck-size intermodal containers. They are a common means of commercial intermodal freight transport and now carry most seagoing non-bulk cargo. A cargo ship sailing from a European port to a U.S. location will typically take 10 to 12 days.[36] Container ship capacity is measured in 20-foot equivalent units (TEU). Typical loads are a mix of 20-foot and 40-foot (2-TEU) ISO-standard containers, the latter being more predominant.

13.3.3.5 AIR TRANSPORT

It may be necessary to consider air shipment for some components, depending on schedule or other constraints. Air shipment of large equipment is usually cost prohibitive and limited by the size and weight of the module. Air transport may be justified in cases of remote site access or need for immediate delivery.

The three major benefits to air freight are speed (overseas flights are less than a day), reliability (allows better tracking), and protection (goods are less likely to be damaged).[37]

13.3.4 MODULARIZATION VERSUS FIELD FABRICATION CONSIDERATIONS

In general, modularization of system components is preferred over field fabrication. However, there are mitigating circumstances where field fabrication may be required or preferred. Typical modularization advantages over field fabrication and erection include:

- Reduced project cost.
- Improved project schedule (e.g., less field disruption, quicker installation).
- Improved quality.
- Reduced safety risk.
- Improved construction efficiency.
- Reduction in site waste.
- Ease of maintenance/equipment replacement.

Modularization of standard ancillary components is the industry norm and pre-configured modules are readily available from a broad list of vendors. Some examples of standard modular ancillary systems are

- Feedwater pumps.
- Service water pumps.
- Firewater pumps.

- Demin/reverse osmosis (RO) water systems.
- Chemical treatment skids.
- Standby diesel generator sets.
- Instrument air skid.
- Electrical motor control centers (MCCs).
- Gas compressors.
- Package boilers (Figure 13-18).
- Pipe Rack assemblies (Figure 13-19).

Used with permission from B&W[38]

Figure 13-18. Package boiler

Used with permission from Flour[39]

Figure 13-19. Pipe rack modules

Specialty systems can be modularized if there is sufficient design definition, time, and engineering resources. However, for single components with small bore piping and/or low voltage electrical requirements, it is typically more cost-effective to field erect.[40]

13.3.4.1 SAFETY

Higher safety factors can be obtained through modularization from the reduction of field labor hours. Other safety benefits may arise from less exposure to difficult access, work at heights, or confined spaces. However, larger modules have their own set of safety concerns related to transport and crane lifts.

13.3.4.2 QUALITY CONTROL

Quality control is typically much easier to maintain with shop fabrication rather than in the field. Shop fabrication is well monitored and typically performed in accordance with ISO 9001 standards. Conversely, quality in the field may be affected by inclement environmental conditions, inadequate labor skill, and/or lack of a solid quality control program. Sometimes the project may have a higher standard for a component or instrument than what is available on a standard modular system and may be forced to accept a lesser quality part as supplied on the standard module.

13.3.4.3 TRANSPORT AND SITE LIMITATIONS

Due to transportation or site constraints, module size and weight may limit the degree of modularization and/or necessitate multiple modules with field interconnections. The project manager should select a shop fabricator as close to the jobsite as possible and consider the limitations of rail, road, and water transportation, in addition to seasonal limits that may impact schedule delivery. Site access and existing building interferences may dictate field erection over modular components or vice versa.

13.3.4.4 LABOR RULES AND CAPABILITIES

Project labor rules and capabilities may influence the decision of whether to modularize or field fabricate. Union labor agreements may dictate that certain component systems (e.g., boiler components) be field fabricated and code stamped, rather than brought to the field already assembled and stamped. The lack of qualified field labor or high field labor costs may dictate that certain component systems be shop fabricated in cases where more exacting quality control or fabrication techniques are required.

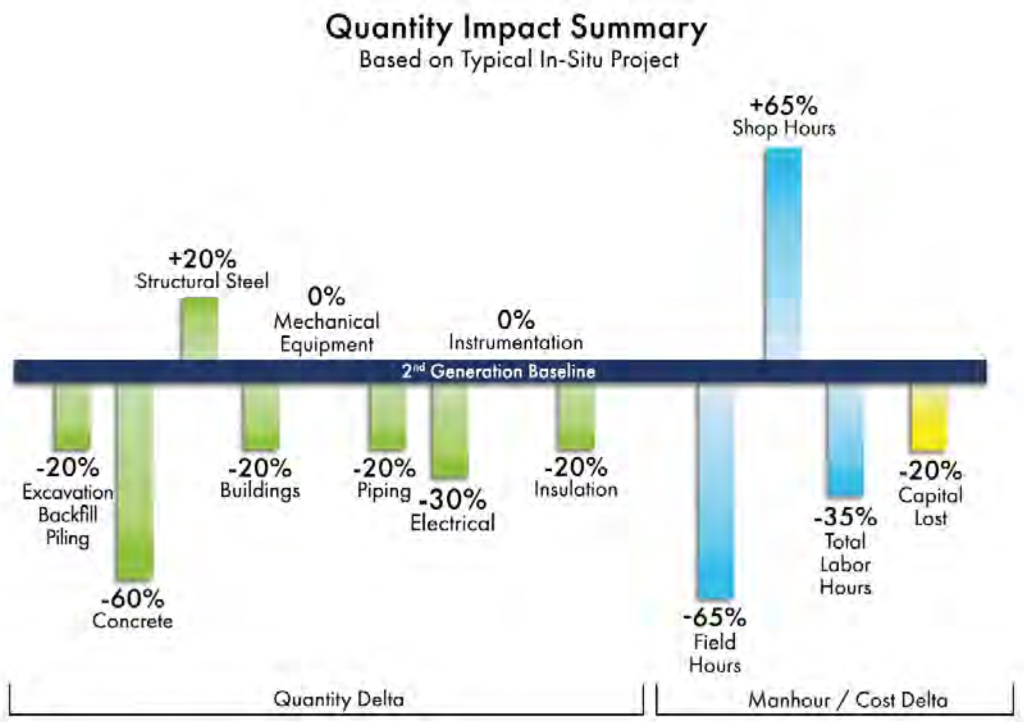
13.3.4.5 SCHEDULE CONSIDERATIONS

Modularization usually improves the overall schedule, allowing component systems to be fabricated offsite and installed onsite with less utilization of the onsite field labor pool. The startup of modularized systems is usually faster, because the system may have already been factory tested and calibrated prior to shipment to the site.

Under some circumstances (e.g., factory delays, part supply shortages) it may be beneficial to partially ship completed modules to the site and field fabricate the remaining work to avoid overall schedule delays. Larger skid-mounted modules may delay early civil foundation design and delay the overall project if the foundation design drawings are not readily available. Ocean marine cargo inspection at the port of export for oversized (i.e., too large to fit in container) modules helps ensure modules are properly secured to minimize the risk of damage in transit.

13.3.4.6 COST CONSIDERATIONS

The project should assess the overall cost/benefit of shop-fabricated system modules versus field erection early in the planning stages. Where labor costs are high and/or site working conditions are adverse, it is almost always cost effective to modularize as many component systems as possible. For specialty systems that have not previously been modularized, the cost of design, engineering, and the risk of shop fabrication or transport delays may justify more field erection. The potential cost savings of modularization are shown in Figure 13-20.



Used with permission from Flour[40]

Figure 13-20. Potential cost savings with modularization

13.3.5 COMMISSIONING PLANNING

Commissioning any facility requires significant pre-planning and methodical implementation to avoid damaging equipment or risking the safety of the commissioning personnel. Commissioning should proceed on a system-by-system basis, starting with the fundamental ancillary systems and finishing with the primary process systems dependent upon the ancillary systems. The EPC contractor usually has a dedicated commissioning team to transition each system from the EPC construction team once the system is deemed mechanically complete. Upon successful commissioning of all systems, the plant is ready for functional and performance testing prior to substantial completion.

All startup and commissioning activities should use available tools and programs to perform safe work prior to starting each task by conducting the following:

- Thorough review and approval of work authorization.
- Follow all LOTO safety checks and balances.
- Pre-activity meeting with vendors and subcontractors.
- Job briefing with craft labor and staff prior to any activity.
- Detailed JHA.
- Engineering review of all previously identified high-risk activities.

13.3.5.1 MECHANICAL COMPLETION

No system should be commissioned until it is deemed mechanically complete. Mechanical completion means that the system is installed and complete with respect to all primary disciplines (i.e., mechanical, electrical, and instrument and controls) and all requisite pre-commissioning installation tests have been successfully completed.

13.3.5.2 PUNCHLIST

The punchlist is a database tool used to maintain the status of all construction defects and deficiencies in the plant that need to be resolved. It is typically developed by the EPC contractor as a spreadsheet database with review and input from the owner. If the EPC contractor is unwilling or fails to provide an objective punchlist, the owner may initiate the task to protect his or her interests.

The punchlist should be developed on a system-by-system basis prior to mechanical completion and commencement of system commissioning. Defects and deficiencies in construction and installation should be identified and prioritized in the punchlist and a resolution should be defined. Priority weightings should be assigned to each item, usually based on some variation of the criterion shown in Table 13-2.

Table 13-2. Example of punchlist priority criterion

PRIORITY	CRITERION	DESCRIPTION	EXAMPLE
1	Critical – Pre-commissioning	Defects or deficiencies that are unsafe for personnel or equipment and prevent commissioning.	Damaged electrical or mechanical connections or lack of proper safety equipment, incorrect equipment or materials, improper installation requiring significant rework to correct.
2	Important – Pre-testing	Defect or deficiencies that would prevent long term operation or cannot be easily resolved once the plant is operational – must be resolved before performance testing.	Lack of maintenance access, lack of proper equipment or instrument supports, missing local instruments, replacement of initial run-in lubricants, filters, or components.
3	Final Completion	Minor items that can be resolved post-testing and substantial completion, but must be resolved before final completion.	General housekeeping and cleanup, final painting, non-critical insulation or heat tracing, and final paving.
4	Commercial/Optional	Items identified by owner that contractor disputes as a scope requirement.	Items that owner believes are specified in the contract, but did not prevent the plant from starting up.

13.3.6 PROCESS SAFETY/HAZARD AND OPERABILITY STUDY

Process safety focuses on the prevention and mitigation of fires, explosions, and accidental chemical releases at process facilities according to the CCPS. It excludes worker health and safety issues, which are usually covered under personnel safety. Process safety hazards can cause major accidents involving the release of dangerous materials, fires, or explosions, which can result in multiple injuries and fatalities, as well as economic, property, and environmental damage. Process safety includes the prevention of leaks and spills, equipment malfunctions, overpressures, excessive temperatures, corrosion, metal fatigue, and other similar conditions. The plant's safety culture should address both personnel and process safety.

13.3.6.1 HAZARD EVALUATION

Hazard evaluation is an organized effort/study to identify and analyze the significance of hazardous situations associated with a process or activity. Specifically, hazard evaluation is used to pinpoint a weakness in the design and operation of a facility that can lead to process safety incidents, such as chemical releases, fires, or explosions. Hazard evaluation techniques can also be used to investigate operability, economic, and environmental concerns. Hazard evaluations should be an integral part of the plant's process safety management (PSM) program. For additional information, refer to the 2008 CCPS Guidelines for Hazard Evaluation Procedure[41] and 29 CFR 1910 Part 119.[42]

13.3.6.2 RISK-BASED APPROACH

Risk, as defined for purposes of process safety, is the combination of the expected frequency and impact of a single potential scenario/incident/event. The scenario risk is determined by the product of the scenario frequency and its total impact:

Scenario Frequency x Scenario Impact = Scenario Risk

(Loss Events/Year) x (Impact/Loss Event) = (Impact/Year)

Scenario risk can be measured in injuries per year, fatalities per year, and total monetary losses per year. Risk is estimated for each scenario and then the scenario risk is ranked as low, medium, or high. High scenario risk implies that the existing safeguards are inadequate. New safeguards should be added and/or the existing safeguards should be modified to reduce the risk to within tolerance limits. The risk may also be determined to be as low as reasonably practicable (ALARP) if additional safeguards are not possible.[41]

13.3.6.3 HAZARD AND OPERABILITY STUDY AND LAYER OF PROTECTION ANALYSIS

A hazard and operability (HAZOP) study is defined by CCPS as a systematic qualitative technique to identify process hazards and potential operating problems using a series of guide words to study process deviations. A HAZOP study is used to question each part of a process to discover what deviations from the intention of the design can occur and their causes and consequences. This is done systematically by applying suitable guide words. This systematic detailed review technique, for both batch and continuous plants, can be applied to new or existing processes to identify hazards.[41]

A layer of protection analysis (LOPA) is defined by CCPS as an approach that analyzes one incident scenario (cause-consequence pair) at a time using predefined values for the initiating event frequency, independent protection layer failure probabilities, and consequence severity to compare a scenario risk estimate to risk criteria for determining where additional risk reduction or more detailed analysis is needed. Scenarios are identified elsewhere, typically using a scenario-based hazard evaluation procedure, such as a HAZOP study.[41]

HAZOP studies and LOPA can be performed in separate steps, but it is recommended to perform these studies together with the same team for consistency in developing scenarios and safeguards.

13.3.6.4 MANAGEMENT OF CHANGE

Occupational Safety and Health Administration (OSHA) 3132 (2000) states that written procedures to manage changes (except for "replacements in kind") to process chemicals, technology, equipment, and procedures, and change to facilities that affect a covered process must be established and implemented[43]. These written procedures must ensure the technical basis for the proposed change, impact of the change on employee safety and health, modifications to operating procedures, necessary period for the change, and authorization requirements for the proposed change. Employees who operate a process must be informed of and trained in the change prior to startup of the process or startup of the affected part of the process.

13.4 SUMMARY

This chapter describes the various ancillary plant systems and their purpose. These ancillary systems are an integral part of the overall gasification and power generation plant. Additionally, precautions required during startup and shutdown operation are addressed.

The chapter presents the supporting elements for project developers to consider during design, including EPC contracts, transportation, modularization, and commissioning. These items can greatly affect the project schedule, overall cost, and risk to the project developer.

The chapter also presents process safety principles and procedures, including HAZOP studies, LOPA, and change management. These are all part of a plant's process safety management system. Process safety is governed by OSHA and there may also be additional state requirements that project developers should review.

Endnotes

1 ISS Source (July 18, 2018) “Lessons Learned: Startup, Shutdown Incidents”, *https://isssource.com/lessons-learned-startup-shutdown-incidents/*.

2 J.C. Bravo (July 2015) “Cyber-Security in Metals and Mining”, Emerson. *https://www.emersonautomationexperts.com/2015/cybersecurity/cyber-security-in-metals-and-mining/*.

3 International Society of Automation (2020) “ISA Standards: Practical Solutions from Industry Experts”, *https://www.isa.org/standards-and-publications/isa-standards/?utm_campaign=smm-ISA-Standards-landingpage&utm_medium=referral&utm_source=blog*.

4 Data Resources (2020) “Modern DCS Control Room”, *https://datacenterresources.com/*.

5 Fieldbus Foundation (2006) “Fieldbus Foundation”, *https://fieldbus.org*.

6 Modbus Organization, Inc. (2020) “Modbus”, *https://modbus.org/*.

7 Profibus & Profinet International (2020) “Profibus”, *https://profibus.com*.

8 Allen Bradley by Rockwell Automation (2020) “HART Communication Protocol”, *https://ab.rockwellautomation.com/Networks-and-Communications/Process/HART-Communication-Protocol*.

9 Sullair (2020) “Typical Skid Mounted Compressed Air System”, *https://sullairhouston.com/products/*.

10 W. Wurtz (December 4, 2015) “Air-Cooled Condensers Eliminate Plant Water Use”, POWER Magazine, *www.powermag.com/air-cooled-condensers-eliminate-plant-water-use/*.

11 B. Buecker (November 19, 2018) “Understanding the Basics of Cooling Tower Heat Transfer”, Power Engineering, *www.power-eng.com/2018/11/19/understanding-the-basics-of-cooling-tower-heat-transfer/*.

12 Watco Group (May 13, 2020) “Surface Condenser and Its Function in Thermal Power Plant”, *www.watco-group.co/surface-condenser-in-thermal-power-plant/*.

13 Q. Lin and T. Collins (July 27-31, 2014) “Mill Steam Inerting System Review and Performance Validation”, *https://www.babcockpower.com/wp-content/uploads/2018/02/mill-steam-inerting-system-review-and-performance-validation.pdf*.

14 J.E. Fisher (January 1, 1978) “Development and Operating Experience with Inerting Systems on Coal Pulverizes and Bunkers”, American Power Conference Journal Volume 40, *https://www.osti.gov/biblio/5922349-development-operating-experience-inerting-systems-coal-pulverizers-bunkers*.

15 J. Benson, V. Dholakia, and X. He (December 1, 2018) Biomass Storage: “Fire Suppression Using Nitrogen Inerting”, *http://www.biomassmagazine.com/articles/15759/biomass-storage-fire-suppression-using-nitrogen-inerting*.

16 Parker (March 26, 2016) “Parker Nitrogen Generations Systems”, *https://promo.parker.com/promotionsite/nitrogen-generators/us/en/home*.

17 DOE (June 2019) “Natural Gas Flaring and Venting: State and Federal Regulatory Overview, Trends, and Impacts”, *https://www.energy.gov/sites/prod/files/2019/08/f65/Natural%20Gas%20Flaring%20and%20Venting%20Report.pdf*.

18 Savant (2016) “Complete Laboratory Setup for Power Plants, DM Plant and ETP’s Along”, *https://savantindia.in/strategic-and-commercial-approach-with-technical-understanding/*.

19 D.G. Daniels (August 31, 2011) “The New Water Lab”, Power, *https://www.powermag.com/the-new-water-lab/*.

20 Solutions Incorporate (2020) “Combined Cycle Plant- Steam and Water Samples”, *http://www.samplepanels.com/project-1.html*.

21 IAPWS (September 2009) IAPWS Technical Guidance Document: Instrumentation for Monitoring and Control of Cycle Chemistry for the Steam-Water Circuits of Fossil-Fired and Combined-Cycle Power Plants.

22 D. Burns (May 31, 2010) “Real-Time Monitoring of Natural Gas Fuel Cleanliness”, Power *https://www.powermag.com/real-time-monitoring-of-natural-gas-fuel-cleanliness/*.

23 SGS (October 2013) “Real Time Analysis In the Coal Industry”, *https://www.sgs.com/-/media/global/documents/flyers-and-leaflets/sgs-min-wa021-real-time-analysis-in-the-coal-industry-en-11.pdf#*:~:text=REAL%20TIME%20ANALYSIS%20IN%20THE%20COAL%20INDUSTRY%20Real,an%20online%20analyzer%2C%20an%20integrated%20mechanical%20sampling%20system.

24 Lyncis (n.d.) “Online Coal Analyzer”, *https://www.lyncis.lt/online-coal-analizer/*.

25 Monitoring Solutions (2020) Gallery, *https://monsol.com/products/compliance-monitoring/extraction-cems#gallery*.

26 EPA, 40 CFR Part 60.254 - Standards for coal processing and conveying equipment, coal storage systems, transfer and loading systems, and open storage piles. October 8, 2009.

27 Power (2008) “Plant Design: Trends in Coal Pile Design”, *https://www.powermag.com/plant-design-trends-in-coal-pile-design/*.

28 American National Standards Institute (2020) NFPA 652 Standard on The Fundamentals of Combustible Dust, 2019 Edition. *https://webstore.ansi.org/standards/nfpa-fire/nfpa6522019?gclid=Cj0KCQjwo6D4BRDgARIsAA6uN18G-aa5fVN$_2$kw-YrLJwfbaLeWJn2ph5UiN6Z_PFvKNYyoPTZVLwdxsaAiRNEALw_wcB*.

29 Fike (2020) “Explosion Suppression Systems”, *https://www.fike.com/solutions/explosion-protection/explosion-suppression/*.

30 S. Rutherford (2007) “Coal-Fired SCR Applications in the US – Challenges and Strategies for Successful Operation and Emission Compliance”, VGB Workshop “Flue Gas Cleaning 2007”, Vienna, Austria May 22–23, 2007. *https://www.cormetech.com/wp-content/uploads/2018/05/2007_VGB_Conference_Paper.pdf*.

31 K. Neidig (2020) “Storage and Handling Procedure of SCR Catalyst”, Mitsubishi Hitachi Power Systems, LTD.

32 D. Naqvi, E. Wey, J. Morgan, M. Miller, T. Nguyen (2014) Transportation Considerations in Module Design, Structures Congress 2014, ASCE 2014.

33 J. Shefer (2015) “Rail Freight Transportation: How to Save Thousands of Dollars & The Environment” Archived 2015-12-08 at the Wayback Machine.

34 Anthony M. Inswasty (February 26, 2013) Freight train in Jacksonville, FL *https://commons.wikimedia.org/wiki/File*:Freight_train_in_Jacksonville,_FL.jpg.

35 ReportBuyer (March 22, 2017) “Barge Transportation Market in the US 2017-2021.” *https://www.prnewswire.com/news-releases/barge-transportation-market-in-the-us-2017-2021-300428019.html*.

36 “*Service Details*”. *cms.molpower.com*. Accessed July 29, 2015.

37 Freightos (2020) “Air Freight & Air Cargo Shipping Explained: Air Freight Charges, Rates, and Costs.” *https://www.freightos.com/freight-resources/air-freight-rates-cost-prices/*.

38 Babcock & Wilcox (2020) “Water-tube Package and Industrial Boilers”, *https://www.babcock.com/en/products/package-boilers*.

39 R. Shulz (2013) Pipe Rack Modules. Fluor.

40 R. Shulz (2013) Potential Cost Savings with Modularization. Fluor.

41 CCPS (2008) Guidelines for Hazard Evaluation Procedures. Center for Chemical Process Safety of American Institute of Chemical Engineers. New York, NY.

42 OSHA, 29 CFR 1910 Part 119 - Process safety management of highly hazardous chemicals. February 8, 2013.

43 OSHA (2000) “Process Safety Management, OSHA 3132.” *https://www.osha.gov/Publications/osha3132.pdf*.

APPENDIX A: GASIFICATION OVERVIEW

For the design of a potential gasification-based power or chemical/fuel production system, it is imperative to have a basic understanding of the gasification process. Gasification is a process that (1) consists of a series of chemical reactions that are overall endothermic and (2) can transform a hydrocarbon material into synthesis gas (syngas). Syngas can be used as both a fuel for power production (e.g., in a boiler or gas turbine combustor) or as a feedstock to produce chemicals or higher-grade fuels. It is equally important to distinguish between gasification and combustion, as shown in Table A-1.

Table A-1. Differences between combustion and gasification

COMBUSTION	GASIFICATION
• Requires the use of large amounts of oxygen. • Exothermic process. • Produces high temperature gases for heating or (indirect) power generation. • Produces large amounts of pollutants.	• Performed with little oxygen. • Endothermic process. • Produces an alternative fuel (syngas) or chemicals. • Products can be cleaned to remove sulfur and/or CO_2.

The table shows that the purpose of gasification is not to burn the feedstock; rather, the purpose of gasification is to produce an alternative fuel (i.e., syngas) or, ultimately, chemicals. These new compounds still retain much of the original feedstock's energy content and can be utilized in much "cleaner" ways than combustion alone.

The overall gasification process is summarized in Figure A-1. In the beginning, heat is added to the feedstock until the moisture within evaporates (called demoisturization). As more heat is added, the volatile matters, or volatiles, are also driven out of the fuel's core structure (called devolatilization). The volatile matters can be in either the liquid or vapor state initially. These two processes are collectively called pyrolysis, from the Greek "pyro" meaning heat and "lysis" meaning to separate, referring to a process in which a solid fuel is broken down into its component parts via heat addition. After pyrolysis, more heat is added to break the long hydrocarbon chains into lighter gases and to gasify the remaining carbon (called char) into syngas. The syngas consists of mostly carbon monoxide (CO) and hydrogen (H_2) as the primary end products, with various amounts of other compounds, such as methane (CH_4), carbonyl sulfide (COS), hydrogen sulfide (H_2S), ethylene (C_2H_4), hydrogen cyanide (HCN), hydrogen chloride (HCl), mercury (Hg), and water (H_2O) and carbon dioxide (CO_2) as the primary non-combustible gases. Raw syngas almost always needs to be cleaned of some of these compounds before it can be used in any commercial application.[1]

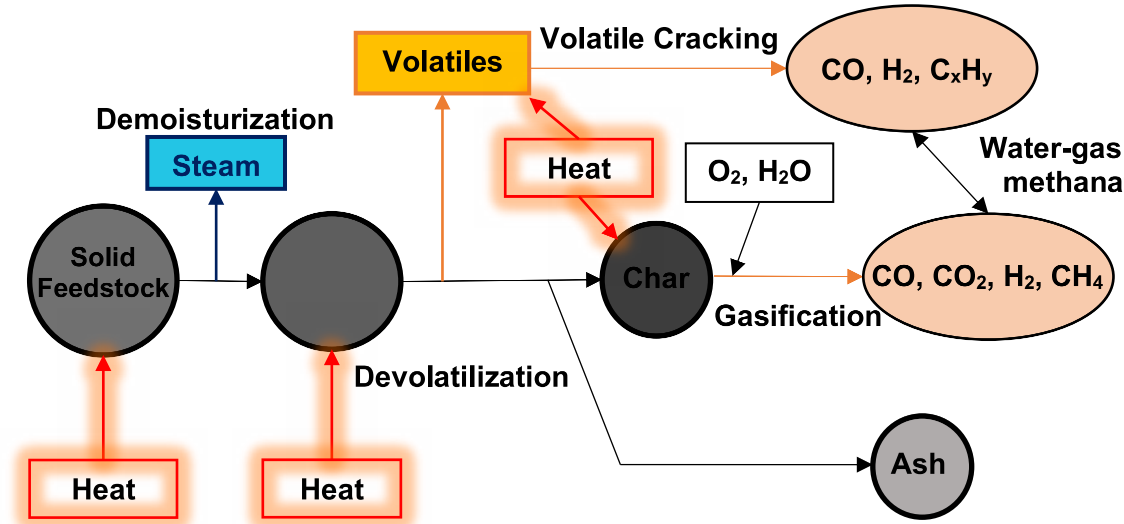


Figure A-1. Gasification process diagram

Nearly any hydrocarbon substance can be gasified, including fossil fuels, biomass, organic wastes, paper, plastic, leather, and petroleum coke. The heat needed for gasification can theoretically come from any source, but the most common method of providing this heat is by partially combusting some of the feedstock, allowing the process to be more directly controllable. This process takes place under very oxidant-lean conditions and generally requires around 25 to 35% stoichiometric oxygen.

A partial list of the chemical reactions utilized in gasification is as follows:

GASIFICATION REACTIONS

HETEROGENEOUS REACTIONS

$2C + O_2 \rightarrow 2C$ (RA.1 – Incomplete Combustion)

$C + H_2O \rightarrow CO + H_2$ (RA.2 – Steam Gasification)

$C + CO_2 \rightarrow 2C$ (RA.3 – CO_2 Gasification/the Boudouard Reaction)

$C + 2H_2 \rightarrow CH_4$ (RA.4 – Direct Methanation)

HOMOGENEOUS REACTIONS

$2CO + O_2 \rightarrow 2CO_2$ (RA.5 – Complete Combustion)

$CO + H_2O \leftrightarrow CO_2 + H_2$ (RA.6 – The Water-Gas Shift Reaction)

$CO + 3H_2 \rightarrow CH_4 + H_2$ (RA.7 – Homogeneous Methanation 1)

$2CO + 2H_2O \rightarrow CO_2 + CH_4$ (RA.8 – Homogeneous Methanation 2)

Volatile Cracking (RA.9)

Reactions A.1 to A.4 are Heterogeneous Phase Reactions, as they occur between a solid (carbon [C]) and a gas. Reactions A.5 to A.9, are Homogeneous Phase Reactions, as they only occur between gases. Reactions A.2 and A.3 are the actual "gasification" reactions, which convert the char into H_2 and CO. Reactions A.4, A.7, and A.8 are methanation reactions. Methane is not a typical byproduct of the gasification process unless the process is specifically designed to produce it (like for SNG applications). Reactions A.1 and A.5 are combustion reactions that are performed to provide the heat necessary for all the other endothermic reactions. Gasifiers can utilize either air for this purpose or high-purity oxygen from, for example, an air separation unit (ASU). This facet of the design process is discussed in Chapters 3 and 5.

Finally, the two most important considerations from a design perspective are the volatile cracking (RA.9) and water-gas shift (WGS; RA.6) reactions. Depending on the application, the WGS reaction can be controlled to alter the H_2/CO ratio of the syngas, which is important for applications that require CO_2 capture and/or chemical/fuel production applications (like the Fischer-Tropsch [F-T] process, which is optimally performed at an H_2/CO ratio of around 2.0). The volatile cracking process is a complex series of reactions that are not universally understood to an adequate level of detail and it is difficult to develop a robust computerized model as a result. This is because the exact chemical configuration and composition of the volatile matters is generally unknown and there is little available literature on this subject. Consequently, simulating gasification must be done alongside scalable experimental testing if the end result is to be used to construct a real plant. Volatile cracking and the WGS reaction are discussed in Chapter 3.

APPENDIX B: ELECTRIC GRID CONNECTION GUIDE

This appendix is provided as a supplement to the detailed content in Chapter 8: Guidelines for Design of an Electric Grid Connection Module. Project designers should review the following six best practices when considering an electric grid connection module:

1. UTILIZE ANY EXISTING ELECTRIC GRID INFRASTRUCTURE TO THE MAXIMUM EXTENT POSSIBLE.

Chapter 8 identifies industrial sites as having sufficient land and (likely) ready access to a subtransmission line (i.e., 34.5 or 66 kilovolts [kV]). It is suggested that the generation power plant export power at 4 kV and the transformer then step-up the generation voltage to 34.5 kV or 66 kV (consistent with Institute of Electrical and Electronics Engineers [IEEE] 1547). An existing substation or connection can reduce cost, but increases complexity.

2. DETERMINE THE BEST APPROACH AND EQUIPMENT TO ACHIEVE ELECTRIC GRID CONNECTION (IF NEW ELECTRIC GRID CONNECTIONS ARE NEEDED).

Required equipment includes circuit breakers, reclosers, and fuses (“switchgear”) on both sides of the interconnection. The equipment on the utility side adds to the interconnection cost (subject to terms of the negotiation). An existing substation can reduce the cost for the structure, but newer switchgear may be needed. The design team should decide if the existing substation remains outside or inside the plant, but the principle of stepping up to subtransmission remains.

3. MODULARIZE ELECTRIC GRID CONNECTION.

The electric grid connection can be modularized by allowing the repetition of standard circuits into a “package” that informs the acquisition of similar gasifier-fueled power plants on industrial sites, brownfields, or other land of sufficient size and access. The standard circuits can be controlled by a standards aware framework, such as Open Field Message Bus (OpenFMB), that support International Electrotechnical Commission (IEC) 61850 and Distributed Network Protocol (DNP) 3 standard communication protocols.

4. CONSIDER HOW TO BEST ADDRESS PERIODIC POWER GENERATION OUTAGES, AS MODULAR UNITS WILL NOT LIKELY BE ABLE TO AFFORD BACKUP REDUNDANCY.

Outages are categorized as electric grid based (i.e., external) or power plant based (i.e., internal). The internal outages are due to planned maintenance of the gasifiers or generators, while unplanned outages are based on failure of either the gasifiers, generators, or faults on the circuits. Battery electrical storage (BESS) and gas storage are options to address these concerns. Gas can also be diverted to the external natural gas system. In any case, this storage allows for the system to be down for several hours and the storage promotes an easier restart. This solution is not as costly as backup redundancy, but it does not guarantee continuous uptime.

5. DETERMINE PROCESS CONTROL REQUIREMENTS.

A gas mixing station outside the electrical generation circuit is recommended for process control. This satisfies the needs of the generators and customers for syngas/gas meeting their requirements and provides for safety from explosion. It is anticipated that a chemical engineer and an electrical engineer are always available. The balance of plant (i.e., supporting components and auxiliary systems) includes station service at 240 volts (V), which powers lights, computers and security systems. Exhaust systems, water pumps, fire suppression equipment, parking lots, a fence, and other elements should be included as part of the design.

6. DETERMINE ANY PROCESS AND OPERATOR SAFETY REQUIREMENTS.

Safety is addressed via Underwriters Laboratory (UL) 1741 and best practices and involves equipment (e.g., fire extinguishers), signage, secure locking of facilities, and education.

INTRODUCTION TO INTERCONNECTION

The interconnection of a distributed power resource to the utility grid is one aspect of project development that can be new to small-scale project developers, and it can be easy to underestimate the time, cost, and complexity of this process. Although new rules for small generator interconnections in Oregon, USA, were completed in 2009—helping to make interconnection more transparent and understandable for project developers—there are many distinct steps in the process of interconnection, and it can be challenging to understand the entire process and practical considerations. There are numerous documents listing the rules and procedures of interconnection, but they lack explanation and practical tips. This guidebook is intended to help navigate through the complexities by highlighting the rules, regulations, and procedures associated with the interconnection of small, renewable energy generation projects producing less than 20 megawatts (MW) of power to the utility grid. Please note that this guide is not a final authority. Rules and regulations are subject to change: project developers should always check their information according to the geographic location and jurisdiction of the project.

WHAT IS INTERCONNECTION?

While small-scale energy systems are capable of powering houses and businesses without any connection to the electric grid, also known as running "off-grid," a grid-connection can offer the system owner significant benefits.

- Any excess electricity produced beyond what is needed is directly fed back into the electric grid. Likewise, if the system is not producing enough to meet energy needs, electricity from the electric grid is available, eliminating the need for electricity storage devices like batteries.
- It is possible to generate revenue from power sales if power is produced beyond the facility's need.

In either case, connecting a system safely to the electric grid in a coordinated manner with the utility is critical. The process of interconnection is designed to ensure that systems are safely interconnected according to established rules and standards and should be incorporated into the project development cycle. It can take up to 18 to 24 months (or longer) from application for interconnection to having a fully interconnected system. Shorter timelines are common with smaller projects, particularly net-metered projects. Besides having the physical work of installing the necessary connection equipment completed, time may be spent studying how the characteristics of the proposed system will impact the electric grid at its point of interconnection (POI) and surrounding area. The electric grid was designed to provide a one-way delivery of electricity from central power plant to home or business. Interconnection of a distributed resource means that the system must act in a different way. There are technical standards and industry rules nested within the interconnection process that must be followed to ensure the continuation of a safe, reliable electric grid system. An overview of the process and common questions are included below. Project developers are advised to start planning and preparing as early as possible.

PROJECT DEFINITION

- What is the project size (MW) and variability around that amount?
- What type of generator will be connected (induction, synchronous, model/size)?
- Where will the connection to the electric grid be made (determine point of common coupling [PCC]/POI)?
- To which utility will the interconnection be made (location dependent)?
- What rules apply to the project (usually Federal Energy Regulatory Commission [FERC]; could be AR 521 or similar for projects less than 10 MW)?
- Will some or all the generated power be exported to the electric grid?
- What type of interconnection (i.e., tier, level, or other category)?

SCOPING (< 1 MONTH)

- Can the technical details of the project be communicated to the utility?
- What are the options for interconnection (other than POIs)?
- Does the design include basic connection equipment (transformers and protection equipment)?

STANDARDS FOR DISTRIBUTED GENERATION: UL 1741 AND IEEE 1547

Procedures reference technical standards that are established industry requirements for system design and operation. UL 1741 and the Institute of Electrical and Electronics Engineers (IEEE) 1547 are the two standards relevant for distributed generation equipment. The reasoning behind generator design and power plant layout may stem from these standards. If a well-proven generator is chosen and a competent engineer assists in the design, the project developer may not need to learn these standards in detail. The use of new or experimental generators may require extensive testing to verify adherence to these standards.

UL 1741 requires that interconnection equipment meet construction constraints, protect against the risks of injury to persons, prescribe rating and marking, and set specific distributed resource tests for various technologies. IEEE 1547 is the standard for interconnecting distributed resources to the electrical grid, required by the Oregon Public Utility Commission (for the interconnection of all distributed resources less than or equal to 10 milli-volt amps [MVA]). IEEE 1547 describes the design technical and testing requirements that guarantee operation of the electric system is not degraded in normal operating conditions. Newer or experimental generators may require extensive testing for potential power quality issues and anti-islanding. Islanding is when the distributed generation is running, but the POI is isolated in an abnormal way from all or part of the local electric grid.

PROJECT DEVELOPMENT

The interconnection process is one the many tasks necessary when developing generation projects. Besides the main formal Interconnection Process, there are several points along the way that deserve to be highlighted.

SITE SELECTION

Starting with site selection, interconnection potential should be considered by the project developer (e.g., what is the access to transmission at this location? what is the capacity available?) There may be multiple distribution lines, transmission lines, or substations nearby. Distribution lines and transmission lines have a relatively fixed capacity and it can be difficult for a project developer to determine how much accessible capacity exists. The project developer is responsible for the costs of building new distribution or transmission lines (i.e., from the project site to the POI on the existing system) and any upgrades necessary to increase transmission capacity or improve system protection; having to design and build long distribution lines, transmission lines, new substations, or other upgrades may be financially prohibitive.

PERMITS

It is the responsibility of the project developer to obtain all permits necessary to build new lines (i.e., it is not the utility's responsibility). When working through permitting for project construction, permits for transmission work may also be needed.

POWER PURCHASE

The power purchase agreement (PPA) often determines the primary revenue stream for the project and is negotiated between the project and utility. A key element to the agreement, in addition to purchase rate, is the start date. If the system is not fully interconnected to the electric grid on the agreed-on start date, the project developer will likely face penalties for not delivering. This is a common issue for projects and project developers must be aware of the dates and obligations in all contracts and how they relate before construction begins. Although there is communication with representatives at the interconnecting and purchasing utility, these individuals are typically in different departments and may even be at different companies (i.e., if the choice is made to wheel the authority to sell electricity to consumers to another utility besides the interconnecting utility). Project developers should not assume that other parties are communicating about the project.

APPENDIX C: ELECTRIC GRID CONNECTION MODULE: ADDITIONAL INFORMATION

This appendix is provided as a supplement to the detailed content in Chapter 8: Guidelines for Design of an Electric Grid Connection Module.

APPLICATION CHECKLISTS

GENERAL INFORMATION REQUIRED FOR ALL APPLICATIONS FOR INTERCONNECTION

✓ Customer Contact Information
✓ Documentation of Site Control or Ownership of the Property
✓ Application Level or Tier
✓ Facility Location
✓ Interconnecting Utility
✓ One-Line Diagram
✓ Generator Manufacturer
✓ Generator Model
✓ Generator Nameplate Rating (kW and kVA)
✓ Generator Voltage
✓ Type of Service (Wye or Delta)
✓ Phase
✓ Design Capacity
✓ Prime Mover (Inverter or Turbine—Induction, Synchronous)
✓ Energy Source
✓ Installation Date, In-Service, and Commissioning Dates
✓ Information Generally Required for Smart Grid Interoperability Panel (SGIP)

FAST-TRACK AND STUDY PROCESS

✓ Tiers 2, 3, and 4 Processes
✓ Documentation of Field-Tested Equipment, Optional for Tier 2
✓ Documentation of Lab-Certified Equipment
✓ Transformer (Phase, Voltages, Configuration, Impedance)
✓ Operation Mode (Qualifying Facility or Not, which is 80 MW or Less, with Biomass or Waste in Gasification Case)
✓ Energy Production Equipment Type
✓ Total Electric Nameplate Power Rating (kW and kVA)
✓ Rated Voltage
✓ Rated Current
✓ Synchronous Generator Specifications
✓ Induction Generator Specifications

GENERATOR INFORMATION

Generator specifications require communication among vendors and engineers before applying for interconnection with the utility. It is recommended to communicate with the generator(s) manufacturer(s) early on about purchasing a generator and obtaining all necessary information. Interconnection applications require specific generator information.

When requesting generator specifications, it is recommended to provide as much specificity as possible. If purchasing a generator secondhand, the project developer should ensure that the technical information is available for the make and model.

SYNCHRONOUS GENERATOR SPECIFICATIONS

- Manufacturer
- Model Number
- Version Number
- Field Amperes
- Reactances

OTHER INFORMATION: INDUCTION GENERATOR SPECIFICATIONS

- Manufacturer
- Model Number
- Version Number
- Locked Rotor Current
- Resistances
- Reactances
- Phase
- Frame Size
- Design Letter
- Temperature Rise

EPACT 2005 (EXCERPT)

The excerpts from EPACT 2005 provide a guarantee of interconnection.

> *(15) INTERCONNECTION.—Each electric utility shall make available, upon request, interconnection service to any electric consumer that the electric utility serves. For purposes of this paragraph, the term ‘interconnection service’ means service to an electric consumer under which an onsite generating facility on the consumer’s premises shall be connected to the local distribution facilities.*

Applicable standard from EPACT 2005:

> *Interconnection services shall be offered based upon the standards developed by IEEE: IEEE Standard 1547 for Interconnecting Distributed Resources with Electric Power Systems, as they may be amended from time to time. In addition, agreements and procedures shall be established whereby the services are offered shall promote current best practices of interconnection for distributed generation, including but not limited to practices stipulated in model codes adopted by Applicability.*

APPENDIX D: WORKING CAPITAL AND SPARE PARTS INVENTORY

This appendix is provided as a supplement to the detailed content in Chapter 13: Guidelines for Design of Gasification-Based Ancillary Systems.

WORKING CAPITAL

Working capital is the money used to meet short-term expenses, including payroll, inventory, and payments on short-term debt. Working capital is critical, because it is used to keep a business smoothly operating and meet all its financial obligations within the coming year.[2]

Net working capital (NWC), is the difference between a company's current assets and its current liabilities.

$$\textit{Net Working Capital} = \textit{Current Assets} - \textit{Current Liabilities}$$

Working capital is one way to measure a company's liquidity, operational efficiency, and short-term financial health. If a company has positive working capital, it should have the potential to invest and grow. If a company's current liabilities exceed its current assets, then it will have difficulty growing or paying back creditors, possibly leading to bankruptcy.[3]

Working capital ratio is the ratio of current assets to current liabilities.

$$\textit{Working Capital Ratio} = \textit{Current Assets} \div \textit{Current Liabilities}$$

A healthy working capital ratio is considered in the range of 1.2 to 2.0 and varies by industry. A working capital ratio of less than 1.0 indicates that current liabilities exceed current assets.

Net working capital fluctuates over time, because the asset and liability figures are based on a rolling 12-month period. It is important to compare NWC from year to year to analyze company health.[4]

SPARE PARTS INVENTORY

Spare parts are crucial for maintaining an operating facility, although there are costs associated with spare parts that are reflected on a company's financial statements. These costs include insurance, taxes, salaries for employees that manage the spare parts inventory and borrowing costs for capital used to purchase the parts.

Normally, a company should strive to only keep in stock the minimum number of spare parts critical to the production process. Partnerships with parts suppliers can be cultivated to provide expedited parts availability without the need to maintain stock in the plant. The need for "just-in-case" inventory must be balanced with the costs associated with maintenance.[5]

The project developer should consider critical spare availability and cost/resource impact of onsite storage and vendor/original equipment manufacturer (OEM) spare availability and delivery lead times when developing spare parts inventory. It may even be advantageous for very critical items to include installed spares in the project layout so that costly shutdowns can be avoided or minimized by quickly switching to such spare equipment should the operating equipment fail unexpectedly. This could include items such as spare gasifiers or spare gasifier feed pumps.

To reduce the overall spare parts inventory, it is important that the project developer/owner try to standardize equipment and equipment unit sizes across the entire project to the greatest extent possible. For modular projects, that also means attempting to standardize parts utilized across what may be multiple subcomponent module fabricators. This is especially true for frequently used parts such as control valves and trims, motors, pumps, heaters, filters, sampling systems, process control system hardware, etc. For those companies seeking to market standard modularized gasification systems, it might be advantageous to stock such common spare part items as a contracted service for their clients, reducing the clients' spare parts inventory requirements and improving overall value to the supplier through a continuing revenue stream. A well-designed and executed preventive maintenance program can also reduce the need for spare equipment inventory.

A circumstance when additional inventory is necessary is during a planned outage for maintenance. During these outages, systematic maintenance is performed throughout the plant. Detailed plans and schedules need to be developed to ensure that all the required maintenance work can be completed in as short a time and as safely possible to minimize plant down time. In addition, all the equipment, parts, supplies, and materials required for the planned outage need to be onsite and readily organized and available to support the aggressive schedule.

One maintenance item typically covered during planned outages is replacing spent catalysts. The catalyst suppliers can supply detailed plans for any passivation of the catalyst in the system, required storage requirements for the new catalyst (which may require special temperature, pressure, and/or humidity conditions to avoid catalyst damage), loading instructions, and activation requirements for the new catalyst after it is loaded into the system. All the catalyst, supplies, and materials required for replacing the spent catalyst need to be onsite prior to initiating the shutdown for the planned outage.

For modular systems, the potential to facilitate these planned maintenance outages can be improved by designing the modular system such that equipment that requires more frequent maintenance are grouped in sub-modules that can be rapidly removed and replaced. Any equipment repair and maintenance of the removed sub-module can occur offline when the plant is back up and operating normally, generating the replacement sub-module for the next planned outage. The concept of sub-modules also lets improvements to any module be easily and effectively implemented as part of any planned outage, allowing continuous improvements for efficiency, performance, and/or cost to be readily made to the plant.

Endnotes

1 J. Rezaiyan and N.P. Cheremisinoff (2005) Gasification Technologies: A Primer for Engineers and Scientists. New York, NY: Taylor and Francis.

2 C. White-Boyte (February 23, 2020) “What Can Working Capital Be Used For?”, *https://www.investopedia.com/ask/answers/102315/what-can-working-capital-be-used.asp*.

3 M. Hart Meredith (2018) “Working Capital: The Definition & Formula”, *https://blog.hubspot.com/sales/working-capital-definition-formula*.

4 A. Bloomenthal Website. “What Does High Working Capital Say About a Company”, (March 30, 2020) *https://www.investopedia.com/ask/answers/102815/what-does-high-working-capital-say-about-companys-financial-prospects.asp*.

5 W. Wilson (n.d.) “What’s the Real Cost of Spare Parts Inventory”, Life Cycle Engineering, *https://www.lce.com/Whats-the-real-cost-of-spare-parts-inventory-1189.html#:~:text=For%20states%20that%20have%20an%20inventory%20tax%2C%20the,up%20to%20as%20much%20as%2020%25%20or%20more*.

NATIONAL
ENERGY
TECHNOLOGY
LABORATORY

1450 Queen Avenue SW
Albany, OR 97321
541-967-5892

3610 Collins Ferry Road
P.O. Box 880
Morgantown, WV 26507
304-285-4764

626 Cochrans Mill Road
P.O. Box 10940
Pittsburgh, PA 15236
412-386-4687

Program staff are also located in
Houston, TX and **Anchorage, AK.**

CUSTOMER SERVICE
1-800-553-7681

WWW.NETL.DOE.GOV