7 Clean Coal Technologies for Advanced Power Generation

One approach to reducing emissions of air pollutants and greenhouse gases, specifically carbon dioxide (CO$_2$), is to increase the energy efficiency of existing thermal cycles and to develop advanced, higher-efficiency cycles. An increase in energy efficiency reduces the amount of fuel consumed and consequently the quantity of pollutants and greenhouse gases emitted. Increasing the thermal efficiency of a power plant is one of three primary techniques to reduce pollutants and CO$_2$ emissions. A second method is to change to less carbon-intensive sources such as natural gas, nuclear power, or renewable sources. The third method is the capture and storage of CO$_2$ from fossil fuel–fired (i.e., coal in particular) power plants. There are three technological pathways for CO$_2$ capture: postcombustion capture, oxy-fuel combustion, and precombustion capture—that is, integrated gasification combined cycle (IGCC).

In this chapter, technologies to increase thermal efficiency and environmental performance are discussed. These include supercritical pulverized coal combustion; fluidized-bed combustion, both atmospheric and pressurized; and IGCC, which is a technology that both improves cycle efficiency and is a CO$_2$ capture technique. Similarly, oxy-fuel combustion, which is an advanced power-generation technology, is also presented in this chapter for both pulverized coal and fluidized-bed combustion systems. Postcombustion CO$_2$ capture is presented in Chapter 10. This chapter begins with an overview of thermodynamic cycles because of the major efforts underway worldwide to increase their efficiency in power-generation facilities.

7.1 Power Cycles

Thermal efficiency is a measure of the performance of a power plant. The two main thermodynamic cycles used widely in power generation are the Rankine and Brayton cycles. These cycles are also referred to as power cycles because thermal energy is converted into mechanical energy, which is then converted to electrical energy. When these two cycles are combined, as in an IGCC system, it is referred to as a Combined cycle.

7.1.1 Rankine Cycle

The Rankine cycle is the basis of all large steam power plants, as briefly discussed in Chapter 6 (refer to Figure 6.16). In coal-fired power plants, high-temperature, high-pressure steam is produced by converting the chemical energy stored in the
coal into thermal energy and transferring the energy to the working fluid (e.g., water), which passes through the boiler to produce the steam. The steam from the boiler is expanded in a series of high- and low-pressure steam turbines, which convert the energy into mechanical shaft work to drive an electric generator and to produce electricity. After the last turbine stage, the steam is routed to a condenser, and the condensate is then pumped back into the boiler, repeating the cycle.

There are variations of the Rankine cycle, but for our discussion purposes the basic Rankine cycle will be presented in this section for simplicity. The Rankine cycle is shown in Figure 7.1 [1]. The working fluid is compressed from low to high pressure (Step 1-2). At this stage, the pump requires little input energy because the fluid is a liquid. The high-pressure liquid enters a boiler, where it is heated at constant pressure to the compressed liquid (Step 2-3) and two-phase (i.e., water and steam) mixture (Step 3-4). The steam is then passed through the superheaters, where it superheated (Step 4-5). The steam is then expanded through a turbine to generate power (Step 5-6). The expanded vapor, which is decreased in temperature and pressure, enters a condenser, where it is condensed at a constant pressure and temperature to become a saturated liquid and the unavailable heat is rejected to the atmospheric sink (Step 6-1).

The thermodynamic efficiency depends on the temperatures at which heat addition and rejection occur:

$$\eta = \frac{T_1 - T_2}{T_1}$$

(7.1)

Figure 7.1 Temperature-entropy diagram of the ideal Rankine cycle.
Source: Modified from Kitto and Stultz (2005).
where $\eta$ is thermal efficiency of conversion from heat into work; $T_1$ is the absolute temperature of the heat source (°R or °K); and $T_2$ is the absolute temperature of the heat sink (°R or °K). From Equation (7.1), there are three ways to increase ideal cycle efficiency: decrease $T_2$, increase $T_1$, or both. Little can be done to reduce $T_2$ in the Rankine cycle because of the limitations imposed by the temperatures of available rejected heat sinks in the environment [1]. Consequently, much of the effort in improving cycle efficiency is directed toward increasing $T_1$, which is discussed in more detail later in this chapter.

### 7.1.2 The Brayton Cycle

The Rankine cycle efficiency limit is dictated by the ratio of the maximum and minimum cycle temperatures. The maximum temperature of the steam Rankine cycle is about 1,150°F (≈620°C), which is set primarily by material constraints at the elevated pressures of the steam cycles [1, 2]. Gas turbines, however, can operate at much higher temperatures than steam turbines, with gas turbine inlet temperatures as high as 2,350°F (1,288°C), thereby making them more thermodynamically efficient.

The conventional Brayton cycle is a gas turbine cycle in which air is compressed and burned with fuel in a combustor. The hot gases are then expanded in a turbine coupled to an electric generator. A simple gas turbine system is shown in Figure 7.2 [1]. In this system, a portion of the work produced by the turbine is used to drive the compressor, and the remainder is used to produce power. The Brayton cycle in Figure 7.2 consists of compressing the combustion air (Step 1-2), adding heat through the combustion process (Step 2-3), expanding the gases through the turbine and performing work (Step 3-4), and rejecting heat to the atmosphere (Step 4-1).

### 7.1.3 Combined Cycle

The gas turbine (Brayton cycle) efficiently uses high-temperature gases from a combustion process but discharges its exhaust gas at a relatively high temperature, which is wasted heat/energy. The steam turbine (Rankine cycle) is not able to achieve the high temperatures of the gas turbine due to materials of construction. Combining the two cycles (high-input temperatures and low-output temperatures) allows one to improve the overall efficiency of the power plant. The combined cycle power plant utilizes a gas turbine (Brayton cycle) to generate electricity, while the waste heat is used in a heat recovery boiler to produce steam (Rankine cycle) to generate additional electricity using a steam turbine. The combined cycle efficiency is a sum of both cycle efficiencies, since they are both powered by the same fuel source. When using coal as the fuel source, one approach to coal firing is the indirect-fired gas turbine cycle. In this cycle, coal combustion gases are passed over tubes that are cooled with clean, high-pressure air; the tubes heat the air to temperatures above 2,000°F (1,093°C), and the hot air is used to drive the gas turbine. Another approach is to gasify the coal and burn the gases that are produced in the gas turbine. This is the concept behind IGCC, which is discussed in more detail later.
Pulverized coal technology is mature, well understood, and discussed in detail in Chapters 5 and 6. The United States has more than 1,100 pulverized coal units, and about 5,000 more units are in operation worldwide [3]. This technology is the most widely used in power generation and is based on decades of experience. The main developments in pulverized coal combustion have involved increasing plant thermal efficiencies by raising the steam pressure and temperature used at the boiler outlet/steam turbine inlet, while ensuring that the units can operate reliably and follow load requirements satisfactorily [4].

Figure 7.2 Simple gas turbine system (a) and temperature-entropy diagram (b) for a Brayton cycle.

*Source:* Modified from Kitto and Stultz (2005).

### 7.2 Pulverized Coal-Fired Power Plants

Pulverized coal technology is mature, well understood, and discussed in detail in Chapters 5 and 6. The United States has more than 1,100 pulverized coal units, and about 5,000 more units are in operation worldwide [3]. This technology is the most widely used in power generation and is based on decades of experience. The main developments in pulverized coal combustion have involved increasing plant thermal efficiencies by raising the steam pressure and temperature used at the boiler outlet/steam turbine inlet, while ensuring that the units can operate reliably and follow load requirements satisfactorily [4].
7.2.1 Advanced Pulverized Coal-Fired Plants

The thermodynamic efficiency of a Rankine steam cycle increases with the increasing temperature and pressure of the superheated steam entering the turbine. It is possible to increase the mean temperature of heat addition by taking partially expanded (and reduced temperature) steam from the turbine and sending it back to the boiler, reheating it, and reintroducing it to the turbine. In the usual designation of steam parameters, the first temperature is the main steam temperature, and the second and third temperatures refer to single and double reheat—that is, 4,484 psia/1,100°F/1,100°F/1,100°F (30.9 MPa/594°C/594°C/594°C). Boilers are referred to as subcritical, supercritical, and ultra-supercritical, depending on the steam’s state. The term subcritical refers to conditions below the critical point of water, which is 3,208 psia (22.1 MPa) and 706°F (374°C). Water’s critical pressure is the maximum pressure in which liquid and vapor can coexist in equilibrium.

At this critical point, the density of steam and water are equal and there is no distinction between the two states. In subcritical boilers, which have two-phase flow, a drum-type boiler is used to separate the steam from the water (as discussed in Chapter 6) before it is superheated and sent into the turbine. As steam pressure and superheat temperature are increased above 3,208 psia (22.1 MPa) and 706°F (374°C), the steam becomes supercritical—that is, it undergoes a gradual transition from water to vapor with corresponding changes in physical properties. In coal-fired boilers, above an operating pressure of 3,208 psia (22.1 MPa) in the evaporator part of the boiler, the cycle is termed supercritical. The cycle medium is a single-phase fluid with homogenous properties, and it is not necessary to separate steam from water in a drum. Therefore, once-through boilers are used because they do not contain drums.

Figure 7.1 shows the critical point on a temperature-entropy Rankine cycle diagram. Ultra-supercritical steam generally refers to supercritical steam at more than 1,100°F (593°C), although the definitions of supercritical and ultra-supercritical boiler pressure and temperature profiles differ from one country to another. Table 7.1

<table>
<thead>
<tr>
<th>Type of Unit</th>
<th>Main Steam Pressure (psia/MPa)</th>
<th>Main Steam Temperature (°F/°C)</th>
<th>Reheat Steam Temperature (°F/°C)</th>
<th>Efficiency (%), based on HHV\textsuperscript{a} and bituminous coal</th>
</tr>
</thead>
<tbody>
<tr>
<td>Subcritical</td>
<td>&lt;3,201/&lt;22.1</td>
<td>Up to 1,050/565</td>
<td>Up to 1,050/565</td>
<td>33–39</td>
</tr>
<tr>
<td>Supercritical</td>
<td>3,207–3,628/22.1–25</td>
<td>1,000–1,075/540–580</td>
<td>1,000–1,075/540–580</td>
<td>38–42</td>
</tr>
<tr>
<td>Ultra-supercritical</td>
<td>&gt;3,628/&gt;25</td>
<td>&gt;1,075/&gt;580</td>
<td>&gt;1,075/&gt;580</td>
<td>&gt;42</td>
</tr>
</tbody>
</table>

\textsuperscript{a}Based on the coal’s higher heating value

Source: From Nalbandian (2008).
shows one definition of these plants, and it lists approximate pressure, temperature, and efficiency ranges for the subcritical, supercritical, and ultra-supercritical pulverized coal-fired power plants [4].

Note that efficiencies are expressed on the basis of either lower heating value (LHV) or higher heating value (HHV) of the fuel. Europe bases its efficiencies on the LHV, which excludes the latent heat of vaporization of the water/moisture in the combustion process (net heat of combustion). In the United States, efficiencies are expressed in terms of the HHV, which includes the latent heat of vaporization of the water/moisture formed in the combustion process (gross heat of combustion). Since the LHV does not take into account the energy used to vaporize the water, efficiencies expressed in LHV are usually higher than those based on HHV. The result is that reported U.S. efficiencies are generally 2 to 4 percent lower than European efficiencies. This chapter discusses efficiencies based on both heating values and notes which basis is being used.

More than 80 percent of the existing coal-fired power plants around the world are subcritical units, and the majority of them are based on a conventional single reheat thermal cycle with subcritical steam pressure in the range of 2,320 to 2,610 psia (16–18 MPa) and the main/reheat steam temperatures both in the range of 1,000 to 1,050°F (535–565°C) [5]. Under these conditions, the efficiencies (LHV basis) range from 33 to 39 percent but can be as low as 20 to 25 percent (LHV basis) in developing nations. Figure 7.3 illustrates the progression of the efficiency of new coal-fired power plants since the 1950s with efficiency targets in 2015 and beyond [5]. The average efficiency of the existing U.S. coal-fired generating fleet is about 32 percent (HHV basis).

Supercritical units first came into operation in the early 1960s and experienced material failures [4]. Problems that were encountered in the early/older supercritical units included [4]:

![Figure 7.3 Progression of coal-fired power plant efficiencies.](image-url)  
*Source: Modified from Baruya (2008).*
The erosion of start-up valves due to high differential pressure because of operating at a constant pressure

- Longer start-up times due to a complicated start-up system and operation
- Low ramp rates due to turbine thermal stresses caused by temperature change in the high-pressure turbine during load changing
- High minimum stable operation load due to by-pass operation and pressure ramp-up operation
- Slagging due to an undersized furnace and inadequate soot blower systems
- Waterwall tube cracking due to metal temperature rise because of inner scale deposits and fireside wastage
- Frequent acid cleaning required due to inappropriate water chemistry
- Lower efficiency than expected due to high air leakage because of the pressurized furnace and complications from reheat temperature control
- Low availability due to the preceding problems

Over the last few decades, supercritical technology has improved with better materials of construction developed and countermeasures to the problems just listed implemented such that supercritical pulverized coal combustion is considered a well-established technology and is expected to play a significant role in reducing emissions from coal-fired power plants, especially with the application of CO₂ capture and storage [6, 7]. Efficiencies of 43 to 45 percent (LHV basis) in supercritical pulverized coal-fired power plants built in the last decade are commonly being achieved [7]. State-of-the-art supercritical plants currently have steam temperatures of 1,112°F (600°C) and reheat temperatures of 1,121 to 1,148°F (605–620°C), which, by some definitions, is considered ultra-supercritical [8]. These plants achieve efficiencies of 43 to 44 percent (LHV basis) for lignite-fired plants and 45 to 46 percent (LHV basis) for bituminous coal-fired plants. These plants have a long history of excellent availability.

In an effort to increase efficiencies further, major development programs are underway to achieve higher ultra-supercritical steam conditions. These programs are being funded by the U.S. Department of Energy’s National Energy Technology Laboratory, the Electric Power Research Institute, the Ohio Coal Development Office, the European Commission, and others. Two notable programs are the European Commission’s Thermie Project and the Ultra-Supercritical Materials Consortium in the United States.

The power-generating industry is moving toward the use of advanced, ultra-supercritical steam cycles, with steam temperatures approaching 1,400°F (760°C) and operating pressures of 5,000 psia (35 MPa) [9, 10]. Increasing the steam temperature is key, since the plant efficiency increases by about one percentage point for every 20°F rise in superheat and reheat temperature [11]. Figure 7.4 shows a generalized relationship between steam temperature and pressure and power plant thermal efficiency [12]. Figure 7.5 shows the extent of improvement possible, as well as the relative effects of increasing steam temperature and pressure [11]. Figure 7.6 shows efficiency improvements for the various power plant steam cycles along with future efficiency targets for ultra-supercritical plants [13].
Supercritical pulverized coal-fired power plants are currently routinely chosen in many countries with unit sizes up to 1,000 MW<sub>e</sub> [4]. There are approximately 500 units operating, under construction, and planned with supercritical and, to a lesser extent, ultra-supercritical, steam parameters [4]. The mature nature of this technology is such that further development of the combustion process itself will consist

**Figure 7.4** The effect of steam temperature and pressure on the thermal efficiency of steam cycle power plants.

**Figure 7.5** Efficiency improvement measures for a pulverized coal-fired plant.
*Source:* From National Coal Council (2007).
of refinements rather than major breakthroughs. Emphasis is mainly on improving efficiencies of these facilities. The main area of activity is the further development of steam conditions beyond the current state-of-the-art supercritical and ultra-supercritical technologies in order to raise thermal efficiency even further [4].

Figure 7.7 shows a new technology development curve (often called “the mountain

Figure 7.6 Efficiency improvements for various steam cycles. $P_S =$ steam pressure, $T_S =$ steam temperature, $T_{RH} =$ reheat steam temperature.


Figure 7.7 New technology development curve for coal.

Source: Modified from National Coal Council (2007).
of death”) for coal depicting the approximate stage of development/commercialization for the various technologies discussed in this chapter [11].

7.2.2 Advanced Ultra-Supercritical Research and Development

In the development of ultra-supercritical steam power plants, currently several areas are undergoing research and development to achieve higher steam pressures and temperatures. This includes materials development, improved fabrication technology, and steam turbine improvements [8, 14]. One of the major limitations in designing advanced ultra-supercritical plants is the creep damage of the materials used in the boiler and in the steam turbine [9]. Also, the boiler materials of conventional power plants do not possess the requisite high-temperature mechanical properties and corrosion/oxidation resistance to meet the requirements of advanced ultra-supercritical plants [10].

Materials of Construction

At present the materials are based on ferritic/martensitic alloys that permit steam temperatures up to around 1,112°F (600°C) in state-of-the-art supercritical plants. Although iron-based alloys could be further developed to achieve even higher conditions, it was recognized during the early 1990s that there would be more opportunities for advancing efficiency improvements by using alloys based on nickel [4]. New nickel-based superalloys for long-term operation at steam temperatures in the range of 1,400°F (760°C) are under development for use in thin-walled outlet headers and steam piping, as well as for castings and forgings of turbine parts [8–10, 14]. In order to reduce the costs that are associated with nickel alloys, further development of existing ferritic-martensitic materials is also necessary. These can be used in areas of lower temperatures (e.g., some furnace walls and turbine stages) to reduce the need for expensive nickel-based superalloys. Corrosion continues to be a problem in high-temperature areas. Hence, development of enhanced coating systems to reduce oxidation, corrosion, and erosion in the high-temperature areas of power plants with higher steam conditions is underway [8, 10].

Improved Fabrication Technology

Improved fabrication technology is another necessary area of research and development for advanced supercritical units. This includes developing fabrication methods for large cast and forged components of nickel-based materials (e.g., turbine rotor and casing). Manufacturing methods for large-diameter pipes composed of nickel-based alloys that transport the steam (under high pressure and temperature) from the last superheater to the turbine inlet need to be developed. Work is underway to develop technologies for welding thick-walled components such as headers and main steam lines. Similarly, work is underway on developing welded joints between materials with different thermal stress behavior—in other words, between superalloys and ferritic steel.
Steam Turbine Improvements

Improvement of the sealing technology for steam turbines is a continuing area of research and development. This becomes more important as the steam pressure is further increased. Also, the development of long-term measurement equipment for static strains in turbine components, as well as sensors for measuring the radial clearance in steam turbines, is necessary.

7.2.3 Oxy-Fuel Firing in Pulverized Coal-Fired Boilers

Oxy-fuel combustion is an emerging approach to postcombustion CO\(_2\) capture. Oxy-fuel combustion is not technically a capture technology but rather a process in which coal combustion occurs in an oxygen-enriched (i.e., nitrogen-depleted) environment, thereby producing a flue gas comprised mainly of CO\(_2\) (up to 89 vol.%) and water. The water is easily separated and the CO\(_2\) is ready for sequestration.

The most common oxy-fuel process involves the combustion of pulverized coal in an atmosphere of nearly pure oxygen (> 95 percent) mixed with recycled flue gas. A schematic of this system is provided in Figure 7.8. Almost pure oxygen for combustion is produced in a cryogenic air separation unit (ASU) and is mixed with recycled flue gas (about 2/3 of the flue gas flow from the boiler) prior to combustion in the boiler to maintain combustion conditions similar to an air-fired configuration. Mixing the recycled flue gas with the pure oxygen is necessary because materials of construction currently available cannot withstand the high temperature resulting from coal combustion in pure oxygen. Research has shown that in order to maintain the oxygen/recirculated flue gas flame so the pulverized coal oxy-combustion flame has heat transfer characteristics similar to that of an air-fired system, an oxygen level of about 30 to 35 vol.% is required in the gas entering the boiler [15].

![Figure 7.8 A pulverized coal oxy-fuel combustion system.](source: Modified from Figueroa et al. (2008) [16] and Kather et al. (2008).)
Power plants that are retrofitted to oxy-combustion would likely operate within the range of oxygen levels that will maintain combustion conditions near those of air-fired systems. New oxy-combustion plants or repowered plants, however, will not necessarily be constrained, and smaller boiler equipment will be able to be used, which will increase efficiency. The critical parameter of oxygen content may be adjusted to increase radiative heat transfer at the expense of convective heat transfer zones further from the flame. This is in contrast to air-fired systems, where approximately half of the heat transfer occurs in the boiler and the balance of the heat transfer is convective in the backpasses. In both retrofit and new systems, the changes inherent in oxy-combustion affect many parameters, including flame behavior, heat and mass transfer, combustion gas chemistry and behavior, char burnout, and slag development, chemistry, and deposition.

Even though oxy-combustion technology is very promising, with a significant amount of research and development focused on it and serious interest in designing and constructing new power plants with it [18], many technical (and economic) issues must be addressed. These can be categorized into combustion technology, steam generator design, CO2 purification, and overall process issues [17].

**Combustion Technology Issues**

The following are some combustion technology issues:

- **Identification of optimum oxygen concentration when firing.** The challenge is to distribute a homogenous mixture of gas and coal to the burner to balance high oxygen levels, and consequently high compressor demands, with low oxygen levels where corrosive atmospheres in the furnace form and incomplete burnout might occur.

- **Combustion behavior of coals in atmospheres comprised of O2, CO2, and H2O.** There are many experimental data published, but more information is needed on oxy-fuel operation of large-scale boilers because not all data are readily scaleable to full-scale systems.

- **Formation mechanisms of pollutants.** The formation of pollutants such as sulfur oxides, NOx, and CO needs to be examined in more detail. A thorough understanding of the formation mechanisms is necessary to develop cost-effective strategies for the control of these pollutants.

- **Reliable oxygen addition to the recycled flue gas.** Oxygen is not introduced into the system where coal drying, transportation, or milling occurs due to the concern for explosions; consequently, it is mixed with recycled flue gas directly to the boiler. In doing so, the oxygen must be reliably mixed to avoid substoichiometric areas near the boiler walls to ensure good coal burnout and minimize corrosion. Handling of high concentrations of oxygen also requires appropriate safety measures, which are not typical of conventional coal-fired boilers.

**Steam Generator Design Issues**

The following are some steam generator design issues:

- **Effect of flue gas composition on heat transfer.** Several heat transfer issues must be addressed, especially since radiation is the predominant heat transfer mechanism in the lower furnace. The atmosphere of the oxy-fuel fired boiler chamber is different from that from an air-fired boiler, which leads to different flame characteristics and radiation heat
transfer behavior. The heat capacity and density of CO$_2$ is higher than N$_2$, which results in a smaller furnace cross section to maintain similar flue gas velocities. CO$_2$ and H$_2$O will increase the radiative heat transfer, but more information is needed to determine how significant this is compared to the radiative contribution from coal and fly ash particles.

- **Optimum temperature for flue gas recirculation.** The temperature of the flue gas dictates the cross section and materials of construction of the flue gas ducting, length of flue gas ducting, and power requirements for recirculating the flue gas. Defining the optimum temperature is dependent on the overall plant configuration and is crucial in defining the overall economics.

- **Fouling and corrosion under oxy-fuel environments.** Fouling and corrosion in oxy-fuel combustion environments is expected to be more severe due to the higher loadings of CO$_2$ and sulfur oxides and, in future designs, higher temperatures. This is a critical issue that needs to be addressed.

### CO$_2$ Purification Issues

The following are some issues related to CO$_2$ purification by liquefaction:

- **Determining phase equilibria of flue gas compositions as basis for liquefaction equipment design.** Knowing the phase equilibria of flue gas mixtures is important in the development of high-efficiency liquefaction equipment, developing simulation tools, and designing CO$_2$ transport and storage systems.

- **Influence of kinetic parameters on the concentration of liquid CO$_2$ and the behavior of contaminants such as sulfur oxides, NO$_x$, and CO during humidification.** A better understanding of gas-liquid solution kinetics can result in the development of improved liquefaction and flue gas cleaning equipment. Similarly, the fate of contaminants during humidification must be better understood because many will be dissolved during humidification, and knowing their effect on effective CO$_2$ capture is important.

- **Pumps for transporting liquid CO$_2$.** The knowledge base for pumping CO$_2$ must be expanded. Large pumps need to be developed for the transportation of liquid CO$_2$.

### Overall Process Issues

The following are some overall process issues:

- **Minimizing auxiliary power by integrating key elements.** The oxy-fuel process has large energy penalties from the ASU and liquefaction equipment. Efficiency improvements must be realized through process integration in order to make the economics more favorable.

- **Long-term stability of materials under flue gas humidification operation.** Corrosion of materials during flue gas humidification is a major concern, and the development or identification of suitable materials is needed.

- **Development of ASU technology.** The development of highly efficient ASU technology for power plants with capacities greater than 400 MW$_e$ is necessary. The largest current ASUs are able to produce approximately half of the oxygen necessary for a 400 MW$_e$ plant.

- **Developing new pollution control process schemes.** When high-purity CO$_2$ is required, it is necessary to remove sulfur and nitrogen oxides upstream of the recycle point. This in turn will influence the size of the equipment, power requirements, efficiency, and investment cost. In addition, if a distillation process must be installed for high-purity CO$_2$, increases in energy demand and capital investment will also increase.
• Developing a new mill design. Current mill designs use air for coal drying and transport, which is not acceptable for oxy-fuel systems. Recycled flue gas can be used, but this may require the flue gas to be dried and stripped of sulfur compounds to reduce corrosion in the mill and to achieve proper coal drying, which in turn results in increased process costs. A preferred option is the development of a mill that can use the recycled flue gas without prior removal of the sulfur compounds and water vapor.

• Control of air in-leakage. The most important source of impurities in the flue gas stream is from uncontrolled air in-leakage, which is typical for boilers operating at slightly sub-atmospheric pressure. The largest concentration of impurities is nitrogen, which in turn is converted to NOₓ. In conventional power plants, air in-leakage is in the range of 4 to 10 percent of the flue gas volume, which is problematic when pure CO₂ streams are the target. It has been reported that a 75 percent reduction in air-leakage should be achievable for an oxy-fuel retrofit and an air in-leakage rate of 2 percent of the flue gas should be feasible for new boilers being designed for oxy-fuel operation [17].

7.3 Fluidized-Bed Combustion

Fluidized-bed combustion (FBC) is a leading technology for the combustion of a range of fuels, fossil and others, because of the many inherent advantages it has over conventional combustion systems, including fuel flexibility, low NOₓ emissions, in situ control of SO₂ emissions, excellent heat transfer, high combustion efficiency, and good system availability. In addition, it is recognized as an early CO₂-mitigation technology by cofiring biomass with coal. General FBC operating principles are presented in Chapter 5, specifically atmospheric bubbling fluidized-bed combustion (BFBC) and atmospheric circulating fluidized-bed combustion (CFBC) systems. This section discusses fluidized-bed combustion units as advanced combustion systems, specifically CFB boilers, pressurized FBC technology, and oxy-fuel firing in CFB boilers. As part of this discussion, some fundamental information on FBC technology is presented as well.

7.3.1 Introduction

Circulating fluidized-bed combustion technology has a number of advantages over BFBC technology including [19]:

• Improved combustion efficiency and sulfur retention due to the use of finer particles, turbulent gas-particle mixing, and a high recycle rate
• Smaller bed area due to the use of high fluidizing velocities
• Reduced number of fuel feed points due to the smaller combustor size and turbulent mixing
• Reduced erosion and corrosion of heat transfer tubes because tubes immersed in the fluidized-bed cooler are subjected to significantly lower gas and particle velocities than in a BFBC, and the cooler is subjected to oxidizing conditions, whereas reducing conditions occur near the fuel feed points in the BFBC
• Increased convective heat transfer coefficients
While CFBC technology has several advantages over BFBC technology, there are also some areas of concern and disadvantages:

- Increased height of the CFBC boiler compared to the BFBC boiler
- Additional height and size of the fluidized-bed ash cooler
- Greater pressure drop across the CFBC boiler resulting in increased fan power requirements
- Need for high cyclone efficiencies for bed material recovery for solids recycle
- Higher erosion in the combustor, cyclone, and associated ducting due to the high gas velocities

CFB boilers have evolved into the utility boiler size range, with a number of units as large as 250 to 300 MW<sub>e</sub> in operation, and they are poised to enter into the realm of larger once-through, supercritical units [20]. However, several opportunities remain to support further development of supercritical circulating fluidized-bed boilers and the pressurized version. Further development in efficiency improvement, fuel flexibility, effective scale-up, and reducing capital cost are underway [21].

As in atmospheric designs, two types of pressurized fluid beds have been applied to power generation: bubbling and circulating. Bubbling bed designs (PBFBC) have been developed, and a few are already in operation, while circulating designs (PCFBC) are mainly in the pilot and conceptual design stages. An example of a PBFBC boiler is shown in Figure 7.9, which is the Tidd PFBC demonstration, the United States’ first PFBC combined-cycle demonstration [22].

Combustion occurs in a large pressure vessel at pressures of 10 to 15 atmospheres and temperatures similar to atmospheric FBC boilers. Because the

![Figure 7.9](image)

Figure 7.9 The Tidd PFBC combined-cycle process.

system is operated under pressure, provisions must be made to feed the fuel and sorbent to the combustor and remove the ash from the combustor across pressure boundaries. Also, primary particulate removal is performed under pressure using high-efficiency cyclones or high-temperature filters. The advantages of PFBC include improved cycle efficiency, reduced emissions, reduced boiler size, improved combustion, and reduced tube erosion [23].

Elevated pressures, and in some cases temperatures, produce a high-pressure gas stream that can drive a gas turbine, and steam generated from the heat in the fluidized bed is sent to a steam turbine, creating a highly efficient combined cycle steam. About 80 percent of the electricity is generated in a conventional steam turbine-generator set, while the balance of the electricity is generated in a gas turbine [21]. Cycle efficiencies of 40 percent are achieved in PBFBC boilers, with efficiencies greater than 50 percent targeted in second-generation PCFBC boilers [24, 25].

The increased pressure of PFBC systems and corresponding air/gas density allows much lower superficial fluidizing air velocities. For a PBFBC boiler, this is approximately 3 ft/sec compared to 10 ft/sec in a BFBC boiler. This reduces erosion of submerged boiler tubes and also permits the use of much deeper beds. The combined effect of lower velocity and deeper bed results in an increased in-bed gas residence time and smoother fluidization, resulting in better gas-solids contact and ultimately better SO₂ capture and improved combustion efficiency. Another advantage of PFBC is reduced boiler size. The high gas density results in a smaller required bed plan area. The lower velocity reduces the total height required for the bed and freeboard [25].

7.3.2 Heat Transfer

In fluidized beds, good mixing is usually achieved, which gives good heat distribution and a uniform temperature distribution. This, in turn, results in effective gas-solids contact, giving a high rate of heat transfer from the burning fuel to the waterwalls or immersed tubes. In conventional furnaces (stoker-fired or pulverized coal-fired), the solids loading in the gas stream is low (i.e., approximately 10 lb per 1,000 lb of gas), and heat transfer from the gas to the waterwalls is primarily by radiation, with a lesser contribution from convective heat transfer [23]. In contrast, in a CFB boiler, the gas leaving the furnace contains a high solids concentration, which can be greater than 5,000 lb per 1,000 lb of gas, and thus convective heat transfer dominates over radiative heat transfer. For equal temperatures, the heat transfer coefficients in an FBC boiler are considerably higher than those in a conventional furnace [23]. However, because the temperatures are lower in an FBC boiler, the overall heat fluxes between the two systems are similar.

What differs between the two systems are the types of heat transfer surfaces employed between the combustion systems. In a conventional system, heat is transferred to the waterwalls (i.e., containment structure) and tubes in the convective pass. In an FBC boiler, three zones of heat transfer must be considered: in-bed, splash zone (interface between the bed and freeboard), and above-bed. In addition, heat transfer areas outside of the furnace must also be considered. In a BFB boiler, heat transfer
surfaces include tube banks in the dense bed, waterwalls in the dense bed, and tubes in the convective pass. CFB boilers do not incorporate a tube bank surface in the bed and rely on heat absorption of the containment walls, internal partitions such as division walls and wingwalls, external ash coolers, and tubes in the convective pass.

7.3.3 Combustion Efficiency

Combustion efficiency, defined as the ratio of heat released by the fuel to the heat input by the fuel, is generally high in FBC systems. The combustion efficiency is typically higher than stoker-fired systems and is comparable to pulverized coal-fired systems. It is generally higher in a CFB boiler than in a BFB boiler because of the use of finer particles, more turbulent environment, and a high solids recycle rate [23]. Similarly, PCFBC boilers achieve higher efficiencies due to smaller and more frequent bubbles, which results in better gas-solid contact.

Combustion efficiency is affected by fuel type, bed temperature, gas velocity, and excess air levels. Combustion efficiency increases with fuel volatile matter content and bed temperature. Combustion efficiency decreases with increasing superficial gas velocity. Combustion efficiency initially increases with increasing excess air level and then decreases. This is believed to be due to an increase in CO and hydrocarbon emissions as the excess air level increases to higher levels [23].

7.3.4 Fuel Flexibility

The flexibility in fuel utilization—mainly the ability to use low-quality fuels—makes FBC boilers an attractive technology. These fuels can be fired solely, in combination with other low-grade fuels, or cofired with coal. The lower combustion temperatures permit burning high-fouling and high-slagging fuels at temperatures below their ash fusion temperature. This greatly reduces operating problems associated with these fuels; however, care is still required, since these fuels contain significant concentrations of alkali and alkaline earth metals. In addition, fuels with low heating values, due to high moisture and/or ash contents, or low volatile matter contents can be successfully burned using an FBC boiler because of the large mass of hot bed material and long residence time that the fuel spends in the bed. Examples of these include fuels such as brown coal, peat, and sludge with moisture contents up to 60 percent; waste coals with ash contents up to 76 percent and higher heating values as low as 2,600 Btu/lb; and petroleum coke with volatile matter content less than 10 percent. The fact that many of these low-grade fuels are difficult to reduce to fine size, due to high ash contents or fibrous structures (as in the case of biomass), makes them candidates for FBC technology, since the fuel does not need to be pulverized but can be processed to sizes 0.25 inch × 0.

Although a main advantage of FBC boilers is that they can be designed to burn a wide variety of low-grade fuels, once an FBC boiler has been designed, there are limitations in deviating from the design values so as not to exceed design limits [23]. As of 2005, biomass is the main fuel type used in BFB boilers, with more than 2,000 MW of installed capacity [21]. This is followed by peat, various ranks of coal
(bituminous coal, subbituminous coal, and lignite), coal wastes, and other wastes, each contributing less than 500 MW_e in installed capacity. In contrast, the main fuel type used in CFBC boilers is bituminous coal (≈10,000 MW_e) followed by lignite (about 4,000 MW_e), with smaller contributions from coal wastes, petroleum coke, biomass, other ranks of coal (anthracite, subbituminous coal, brown coal), peat, wastes, and RDF pellets, where each category contributes 2,000 MW_e or less [21].

Figure 7.10 illustrates the wide range of fuels that have been used/tested in FBCs, but the range of fuels is endless. Candidate fuels include, not inclusive, the various ranks of coal (anthracite, bituminous coal, subbituminous coal, lignite, brown coal), waste coal from coal cleaning operations, petroleum coke, oil shale, refinery bottoms, peat, woody biomass, herbaceous biomass, manure and litter, animal-tissue biomass, tires, paper mill sludge, sewage sludge, refuse-derived fuel, pellet-derived fuel, plastics, industrial wastes, and more.

The fuel characteristics have a minor impact on a CFB utility boiler design compared to that of a pulverized coal-fired boiler. Parameters that must be considered when arriving at a final pulverized coal-fired design include the heat release rate, fuel properties (e.g., ash fusion temperatures, volatile matter content, ash content), percentage of excess air, production of emissions, boiler efficiency, and steam temperature [27], with the most important item to consider being the fuel burned. Furnaces for burning coal are more liberally sized than those for gas or fuel oil firing. This is necessary to complete combustion in the furnace and to prevent formation of fouling or slagging deposits. This is depicted in Figure 7.11, where the impact of coal type on furnace size is shown [28]. By contrast, Figure 7.12 shows the impact of coal type on a CFBC design, which is small [28].

![Figure 7.10 Applicable fuels for FBC technology. Source: Modified from Foster Wheeler (2006) [26].](image-url)
7.3.5 Pollutant Formation and Control

Fluidized-bed coal combustors have been called the “commercial success story of the last decade in the power-generation business” and are perhaps the most significant advance in coal-fired boiler technology in half a century. Originally, development of the technology was focused on manufacturing a compact, package boiler.

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Figure 7.11 Impact of coal type on pulverized coal-fired furnace design. 

Figure 7.12 Impact of coal type on circulating fluidized-bed furnace design. 

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7.3.5 Pollutant Formation and Control

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that could be preassembled at the factory and shipped to a plant site, thereby providing a lower-cost alternative to onsite assembly of conventional boilers. By the mid-1960s, however, it became apparent that the fluidized-bed boiler not only represented a potentially lower cost and more efficient way to burn coal, but it also generated less emissions than conventional boilers. Because the technology can control sulfur dioxide and nitrogen oxides at lower cost than conventional boilers, FBC technology has been developed from the “package boiler” concept to the utility boiler concept. This section discusses pollutant formation and control, with an emphasis on sulfur and nitrogen oxides as it pertains to FBC systems. In Chapter 9, additional pollutant formation and control information are presented.

**Sulfur Dioxide**

Sulfur in the fuel is oxidized to SO$_2$ during the combustion process. In an FBC boiler, the SO$_2$ is captured in situ by adding a sorbent material, which most commonly is limestone but sometimes can be dolomite. Sulfur retention can be greater than 95 percent in an FBC boiler, but sorbent utilization levels are relatively low (e.g., typically only about 40 percent of the calcium is utilized in the capture of sulfur). The effect of operating parameters and sorbent characteristics on sulfur capture are discussed in the following sections.

**Transformation of Sorbents in the FBC Process**

In an FBC system, limestone and dolomite will undergo thermal decomposition, a process commonly known as calcination. The decomposition of limestone proceeds according to the following equation:

$$\text{CaCO}_3 \rightarrow \text{CaO} + \text{CO}_2$$  \hfill (7.2)

Calcination of limestone is an endothermic reaction, which occurs when limestone is heated above 1,400°F (760°C). Calcination is necessary before the limestone can absorb and react with gaseous sulfur dioxide. Calcined limestone is porous due to the loss of carbon dioxide.

Capture of the gaseous sulfur dioxide is accomplished via the following equation to produce a solid product—calcium sulfate:

$$\text{CaO} + \text{SO}_2 + \frac{1}{2}\text{O}_2 \rightarrow \text{CaSO}_4$$  \hfill (7.3)

The reaction of porous calcium oxide with sulfur dioxide involves a continuous variation in the physical structure of the reacting solid with conversion. One of the major factors responsible for these changes is the formation of calcium sulfate, which has a higher molar volume than calcium oxide (52 cm$^3$/mole for CaSO$_4$ compared to 17 cm$^3$/mole for CaO). Because of this expansion in the solid volume from the reactant to the product, the pore network within the reactant will be progressively blocked as conversion increases. As a result, the reaction rate will decay rapidly as soon as a shell of the product layer is formed on the outside of the reacting
solid. For pure CaO prepared by the calcination of reagent grade CaCO$_3$, the theoretical maximum conversion of CaCO$_3$ to CaSO$_4$ has been calculated to be 57 percent.

At this conversion, the void spaces in the solid are completely filled by the CaSO$_4$ formed. In practice, the actual conversion obtained using natural limestones is much lower. Calcium utilization as low as 15 to 20 percent has been reported in some cases [29] and is typically 30 to 40 percent. Low utilization efficiency can be explained by early pore blockage making it more difficult for SO$_2$ to reach the reactive CaO in the particle interior. Sulfur dioxide can usually penetrate only a small distance (50–100 $\mu$m) into a particle until pore plugging inhibits further sulfation [30]. Therefore, fine particles can be sulfated more completely than coarse particles, but they also tend to be elutriated from the system before they have had time to become fully sulfated.

Physically, the limestone undergoes structural changes such as porosity, pore volume, and surface area development; changes in pore size distribution; and sintering. The particles also undergo degradation, which is commonly referred to as attrition, with the resultant small particles removed from the system with the offgas. The effect of each of these physical properties on sulfur capture is discussed later in this chapter.

The decomposition of dolomite, MgCa(CO$_3$)$_2$, which is a double carbonate of magnesium and calcium, is similar in many respects to the decomposition of limestone. The reaction proceeds either in a single step or in stages, depending on the operating conditions and the chemical composition of the sample. For a single-step decomposition, the overall process can be described by the equation

$$MgCa(CO_3)_2 \rightarrow MgO + CaO + CO_2$$  \hspace{1cm} (7.4)

When decomposition occurs in stages, MgCO$_3$ will decompose first. The following reactions will take place in sequence:

$$MgCa(CO_3)_2 \rightarrow MgO + CaCO_3 + CO_2$$  \hspace{1cm} (7.5)

$$CaCO_3 \rightarrow CaO + CO_2$$  \hspace{1cm} (7.6)

MgO will not react with sulfur dioxide in the reaction gas at temperatures above 1,400°F (760°C); therefore, the sulfation reaction of dolomite is basically the reaction of sulfur dioxide with calcium oxide (Eq. 7.3).

In PFBC systems, however, the partial pressure of CO$_2$ (1,400°F; 760°C) is so high that calcination of limestone does not proceed because of thermodynamic restrictions. For example, at 1,560°F (850°C), calcium carbonate does not calcine if the CO$_2$ partial pressure exceeds 0.5 atmospheres. In these conditions, the sulfation reaction for limestone is

$$CaCO_3 + SO_2 + \frac{1}{2} O_2 \rightarrow CaSO_4 + CO_2$$  \hspace{1cm} (7.7)
Increasing the pressure from 1 to 5 atmospheres significantly increases the sulfation rate and calcium utilization [31, 32]. Under pressure, dolomite may partially calcine, and sulfation is represented by the following reaction:

$$CaCO_3 \cdot MgCO_3 + SO_2 + 1/2O_2 \rightarrow CaSO_4 \cdot MgO + 2CO_2$$ (7.8)

**Bed Temperature**

The effect of bed temperature on sulfur capture is well known. The peak sulfur retention in an atmospheric fluidized-bed combustion boiler is between 1,450°F and 1,650°F (790–900°C), which coincides with the increasing extent of calcinations. Plants with insufficient heat transfer area have exhibited high bed temperature and poor limestone utilization. In addition to the level of the bed temperature, the uniformity of the bed temperature is important because hot and cold locations in the bed must be avoided.

An optimum temperature exists that is sensitive to sorbent type, particle size, calcination conditions, and pressure. This temperature factor is due to an irreversible effect—the generation and loss of pores by sintering [33, 34]—and decomposition of CaSO₄ formed during oxidizing conditions [35].

For PFBC boilers, however, there is no pronounced maximum for sulfur retention as a function of temperature. Within the normal operation range, sulfur capture is found to increase slightly with temperature [19]. When dolomite is used as a sorbent, the increase in sulfur retention with an increase in bed temperature can continue to as high as 1,750°F (950°C) [19].

**Particle Residence Time**

The residence time of the sorbent particles in an FBC boiler is a function of the particle size (and the fluidizing gas velocity). The particle size distribution is a continuum, but a variety of particle sizes are found in the system. Sorbent particle size (and thus residence time) is important because the rate of sulfation is proportional to the particle surface area, which is inversely proportional to the particle diameter. In other words, the smaller the particle, the faster the rate of sulfation. Unfortunately, the smaller particles, which may have a fast sulfation rate, have the shortest residence time in the combustor (~2 seconds, similar to that of the gas residence time). The larger particles may have slower rates of sulfur capture, but they are contained in the system for a significantly longer period of time.

**Bed Quality**

Bed quality, which includes limestone distribution, mixing, and fluidization, is important in capturing sulfur in the bed. Obviously, the better the contact between the limestone, sulfur dioxide, and oxygen, the greater the quantity of sulfur that can be captured. Also, increasing the fluidizing gas velocity decreases the residence time of the gas in the bed. This, in turn, reduces the effective contact between SO₂ and the sorbent and, consequently, reduces sulfur capture.
Gaseous Environment

Another important parameter is the gaseous atmosphere in the combustor. Oxygen is necessary for the sulfation reaction. FBC boilers normally operate with about 4 percent vol O₂ in the bed, which is far in excess of the SO₂ concentration. Under this condition it is commonly assumed that the sulfation reaction has zero dependency on O₂ concentration. Oxygen can indirectly reduce the rate of CaO sintering, however. Sintering is a complex process that results in a reduction in the total surface area and thus a change in the pore size. Carbon dioxide enhances sintering, which can be detrimental to sulfur capture by causing CaO particle shrinkage and densification (which occurs during the later stages of sintering). Excess O₂ will reduce the CO₂ partial pressure and thus impede sintering.

The initial rate of sulfation is approximately proportional to the SO₂ concentration, but the reaction is terminated by pore blockage occurring more quickly as the SO₂ concentration is raised. In addition, it appears that the ultimate extent of sulfation attainable is slightly dependent on the amount of SO₂ present in the bulk gas.

Combustor Pressure

Combustor pressure also affects sulfur capture in an FBC boiler. No peak in the sulfur retention versus temperature curve is observed at high pressure (≈12 atmospheres). Limestone performance is reduced, while dolomite performance improves. Dolomite is the preferred sorbent in pressurized fluidized-bed combustors.

Chemical Composition

Sorbent performance in FBC boilers has been well documented in literature. It is generally accepted that sorbent performance has no relationship to chemical composition [19, 29, 30, 36, 37]. A high-purity limestone (>95 percent CaCO₃) will not necessarily perform better than a low-purity limestone (<70 percent CaCO₃).

Porosity

The total porosity, a dimensionless quantity, is defined as the ratio of the sum of the pore volume to the total volume. It is an important physical property because it provides insight into whether the solid has sufficient interior vacancy in which the sulfation reaction can occur. Porosities of raw stones generally vary between 0.3 and 12.0 percent and that of calcines from 20 to 50 percent, which is less than the theoretical porosity of 54 percent for CaO derived from carbonates [38].

The porosity of a stone increases upon heating because of the release of CO₂. The rate of porosity increase depends on the calcination conditions. Since this porosity increase is so condition-specific, it is impossible to predict a priori. However, porosity increases for different stones between a factor of 5 and 180 for the same calcination conditions have been reported [39]. This increase in porosity can be quickly negated by four processes: sulfation, thermal sintering (more than 1,470°F; less than 950°C), moisture-activated sintering, which is believed to be most prevalent at temperatures around 1,110°F (600°C), and CO₂-activated sintering, which is most intense at approximately 1,650°F (900°C) [40].
The importance of using an initially porous stone can be seen from the following equation [41]:

\[ x_s = \frac{\varepsilon_o}{(z - 1)(1 - \varepsilon_o)} \]  

(7.9)

where \( x_s \) = the maximum conversion at the surface of CaO to CaSO\(_4\) before pore closure occurs; \( \varepsilon_o \) = the initial porosity of the calcine; and \( z \) = molar volume of CaSO\(_4\)/molar volume of CaO.

Using 0.53 for the initial porosity of CaO, \( x_s \) is maximized at 56 percent utilization. However, stone utilizations typically range between 30 and 40 percent. The 16 to 26 percent difference is partly because the porosity of the stone is quickly reduced by the formation of CaSO\(_4\), and only a 30 to 40 mol\% fraction of the outermost region of the stone reacts with the gaseous SO\(_2\). This phenomena is often referred to as pore mouth plugging and is the dominant cause for only partial sulfation of sorbent particles in the size range between 18 and 140 mesh (1,000 and 105 \( \mu \)m) [40]. There are four main ways to increase stone utilization via optimizing calcine structure:

- Calcining with high CO\(_2\) partial pressures to produce a wide-mouth pore size distribution
- Using slow calcination rates
- Calcining at temperatures between 1,380°F and 1,470°F (750–950°C)
- Using impure stones

Particle porosity is believed to be the controlling parameter for larger particles. Porosity accounts for larger product molar volumes, which in turn can result in larger sorbent utilizations at long residence times (i.e. when pore plugging becomes rate limiting) [42]. The pore size distribution plays only a minor role in controlling sulfation, whereas increasing the sorbent porosity appears to be the single most promising method to enhance sulfation, especially if the stone is initially 15 percent porous [43]. However, it has been shown that large-mouth pores greatly enhance stone utilizations and identified pore volume distribution and grain size as the determining factors for stone utilization [44].

Surface Area

Surface area is related to porosity. As with porosity, the surface area of the resulting calcine increases substantially upon heating. The surface area development is controlled by the particle size and calcining conditions: temperature, heating rate, and environment. Surface areas as large as 90 m\(^2\)/g have been obtained by using low temperatures (1,110–1,650°F; 600–900°C), high-heating rates, small particles, and CO\(_2\)-free environments.

As with porosity, internal surface area is the limiting parameter for smaller particles, smaller product molar volumes, lower conversions, and short residence times (i.e., when kinetics may be rate determining) [42].

Like chemical composition and porosity, there is no experimental correlation between surface area and stone performance. However, for an irreversible first-order gas-solid reaction under chemical control, the time required to reach a given conversion is inversely proportional to the active surface area and inversely proportional to the square of the surface under product layer diffusion control. For fluidized-bed applications, the latter relation is of most interest [45].
Particle Size
Conversion of CaO to CaSO₄ increases as particle size decreases over the entire particle size range of 1 to 1,000 µm. From a fundamental viewpoint, reducing the particle size is beneficial because less CaO will be inaccessible (on a weight basis). The ideal particle size distribution for an FBC power plant is site specific. That is, it depends on the manufacturer and specifications of the combustor and cyclone, the economics of producing and feeding fines, the ash content of the fuel, and the propensity for the stone to attrit. These are only a few of the many considerations that must be included in the evaluation process. Laboratory results indicate that reducing the particle size will be beneficial in capturing more sulfur per unit mass of stone. However, in the field, operating conditions and economics will dictate the specifications of the particle size distribution.

Nitrogen Oxides
The nitrogen oxides (mainly NO, NO₂, and N₂O, but referred to as NOₓ) that are formed in an FBC derive from two sources: oxidation of fuel nitrogen, which is referred to as fuel NOₓ, and reactions of oxygen and nitrogen in the air, which is referred to as thermal NOₓ. Since thermal NOₓ is a product of a high-temperature process, mainly above 2,700°F (1,480°C), it makes only a minor contribution to the overall NOₓ emissions, and fuel NOₓ is the primary contributor. Typically, more than 90 percent of the NOₓ is in the form of NO, approximately 20 to 300 ppm as N₂O, with the balance NO₂ [19]. A discussion of NOₓ formation mechanisms, the effect of fuel characteristics and operating parameters on NOₓ formation, and NOₓ reduction techniques is presented in the following sections.

NOₓ Formation
The formation of NOₓ compounds is complex. Upon heating the fuel, the organically bound nitrogen volatilizes, forming volatile-nitrogen and char-nitrogen. The volatile-nitrogen compounds are primarily released as HCN and NH₃. In the char, the nitrogen is bound in aromatic structures [19].

HCN and NH₃ undergo the following reactions during combustion [19]:

\[
HCN + \frac{5}{4}O_2 \rightarrow NO + CO + \frac{1}{2}H_2O \tag{7.10}
\]

\[
HCN + \frac{3}{2}O_2 + NO \rightarrow N_2O + CO + \frac{1}{2}H_2O \tag{7.11}
\]

\[
HCN + \frac{3}{4}O_2 \rightarrow \frac{1}{2}N_2 + CO_2 + \frac{1}{2}H_2O \tag{7.12}
\]

\[
NH_3 + \frac{5}{4}O_2 \rightarrow NO + \frac{3}{2}H_2O \tag{7.13}
\]

\[
NH_3 + \frac{3}{4}O_2 \rightarrow \frac{1}{2}N_2 + \frac{3}{2}H_2O \tag{7.14}
\]
During char oxidation, the nitrogen is mainly oxidized to NO and \( \text{N}_2\text{O} \). These are, in turn, partially reduced to \( \text{N}_2 \). NO is reduced through reactions with NH\(_3\), carbon in the char, and CO. \( \text{N}_2\text{O} \) can be reduced by temperature effects or reactions with char and CO. NO and \( \text{N}_2\text{O} \) reduction reactions are [19]

\[
\text{NO} + \text{NH}_3 + \frac{1}{4}O_2 \rightarrow \text{N}_2 + \frac{3}{2}H_2O \quad (7.15)
\]

\[
\text{NO} + \frac{2}{3}\text{NH}_3 \rightarrow \frac{5}{6}\text{N}_2 + H_2O \quad (7.16)
\]

\[
\text{NO} + C(\text{char}) \rightarrow \frac{1}{2}\text{N}_2 + \text{CO} \quad (7.17)
\]

\[
\text{N}_2O \rightarrow \text{N}_2 + \frac{1}{2}O_2 \quad (7.18)
\]

\[
\text{N}_2O + C(\text{char}) \rightarrow \text{N}_2 + \text{CO} \quad (7.19)
\]

\[
\text{N}_2O + \text{CO} \rightarrow \text{N}_2 + \text{CO}_2 \quad (7.20)
\]

**Fuel Nitrogen and Volatile Matter Content, and Fuel Rank**

In general, NO and \( \text{N}_2\text{O} \) emissions increase with increasing nitrogen content in the fuel, whereas with increasing fuel volatile matter content, NO emissions usually increase, but \( \text{N}_2\text{O} \) emissions decrease. \( \text{N}_2\text{O} \) emissions are primarily dependent on the type of fuel (as well as combustion temperature, as discussed later) with low-rank fuels generating the lower amounts of \( \text{N}_2\text{O} \) than higher-rank fuels—that is, \( \text{N}_2\text{O} \) emissions from lowest to highest—biomass, peat, oil shale, brown coal/lignite, bituminous coal [46, 47]. This is attributed to the amount of NH\(_3\) released from the fuel as lower-rank fuels (i.e., higher-volatile-matter fuels) release more NH\(_3\), and their NH\(_3\) to HCN ratio is greater than for higher-rank fuels. Oxidation of NH\(_3\) results in NO formation, while HCN produces both NO and \( \text{N}_2\text{O} \).

**Combustion Temperature**

Combustion temperature has a significant effect on both NO and \( \text{N}_2\text{O} \) emissions [46, 47]. NO emissions increase, but \( \text{N}_2\text{O} \) emissions decrease. Higher temperatures promote the oxidation of nitrogen radicals to NO and reduce the char and CO concentrations, which in turn decreases the reduction of NO to \( \text{N}_2 \) on the char surface. The reduction reactions of \( \text{N}_2\text{O} \) with hydrogen radicals (H, OH) are significantly enhanced via the following reactions [19]:

\[
\text{N}_2\text{O} + H \rightarrow \text{N}_2 + \text{OH} \quad (7.21)
\]

\[
\text{N}_2\text{O} + \text{OH} \rightarrow \text{N}_2 + \text{HO}_2 \quad (7.22)
\]
Excess Air
An increase in excess air leads to increases in NO and N\textsubscript{2}O emissions. As excess air is increased, the combustion rate increases. However, excess air has a minor effect on N\textsubscript{2}O emissions compared to fuel type and combustion temperature [47]. This is especially true in pressurized units [19].

Gas Velocity/Residence Time
The superficial gas velocity—that is, the residence time in the bed—has an effect, although minor, on N\textsubscript{2}O emissions but not on NO emissions. With increasing gas velocity, the contact time between the gas and particles decreases, thereby resulting in higher N\textsubscript{2}O emissions because there is less time for the carbon-N\textsubscript{2}O reduction reaction to occur [47].

Limestone Effects
The effect of limestone addition on NO\textsubscript{x} emissions is complex. In a BFBC boiler, the limestone is primarily in the dense phase of the bed where CO concentration is high and O\textsubscript{2} concentration is low, and the catalyzed reduction of NO by CO may dominate over oxidation of NH\textsubscript{3}, thereby leading to lower NO emissions [19]. In a CFBC boiler, however, the limestone is distributed through the entire combustor, and oxidation of volatiles to NO may occur in the upper zone where CO concentrations are lower, thereby increasing NO emissions. The addition of limestone is usually found to decrease N\textsubscript{2}O emissions, although the effect is minor [19, 47]. The decrease is attributed to the limestone catalyzing the decomposition of N\textsubscript{2}O.

NO\textsubscript{x} Reduction Techniques
The FBC process inherently produces lower NO\textsubscript{x} emissions due to its lower operating temperature. However, where necessary, additional combustion modifications or flue gas treatment for NO\textsubscript{x} control can also be employed. Techniques currently used for FBC systems include reducing the peak temperature by flue gas recirculation, natural gas reburning, overfire air/air staging, fuel reburning, low excess air, and reduced air preheat. Postcombustion control is also used, including selective noncatalytic reduction (SNCR) and selective catalytic reduction (SCR), which achieve 35 to 90 percent NO\textsubscript{x} reductions. Air staging, SNCR, and SCR will be briefly discussed here but are presented in more detail in Chapter 9.

Air staging can reduce both NO and N\textsubscript{2}O emissions from FBC boilers. This is accomplished by introducing less than the theoretical amount of combustion air through the distributor plate and adding the remainder of the combustion air above the dense bed or, in the case of CFBC boilers, injecting some of the secondary air into the cyclone. As a result, some of the fuel nitrogen compounds decompose into molecular nitrogen rather than forming NO\textsubscript{x} due to the reducing atmosphere in the bed. Char and CO concentrations in the bed increase, thereby enhancing the rates of NO and N\textsubscript{2}O reduction on the char.

SNCR is a technology that involves injecting nitrogen-containing chemicals into the FBC boiler within a specific temperature window without the expensive use of
catalysts. The chemicals, with the two most common being ammonia and urea, selectively react with NO in the presence of oxygen to form molecular nitrogen and water. The main reactions when using ammonia or urea are, respectively,

\[
4NO + 4NH_3 + O_2 \rightarrow 4N_2 + 6H_2O
\quad (7.23)
\]

\[
4NO + 2CO(NH_2)_2 + O_2 \rightarrow 4N_2 + 2CO_2 + 4H_2O
\quad (7.24)
\]

The optimum temperature window for ammonia is 1,560 to 1,920°F (850–1,050°C) and for urea is 1,830 to 2,100°F (1,000–1,150°C). While both have been used, ammonia is favored due to its optimum temperature window and that fact that ammonia produces less \( N_2O \) than urea [47]. SNCR operation is very effective for reducing NO, but it does increase \( N_2O \) emissions. Fortunately, the overall concentration of \( N_2O \) is much less than NO. Where it is necessary to reduce \( N_2O \) emissions even further, a combination SNCR/SCR can be employed for overall \( NO_x \) control [47].

**Particulate Matter**

Particulate matter is generated from two sources: the fuel and the sorbent. The mineral matter in the fuel is released during combustion, and this residue, along with calcined sorbent and reacted sorbent, is referred to as ash. In addition to the ash, there can be unburned carbon in the residue, although this amount should be minimal provided the system is properly operating. Some of the ash remains in the fluidized bed and is discharged by the bed material drain system. This ash is normally larger than 105 \( \mu m \) and is relatively easy to handle and transport [23]. The remaining ash is fine and leaves the boiler in the flue gas. This material is typically less than 44 \( \mu m \) and requires a high-efficiency collection device. Normally this is a fabric filter but can also be an electrostatic precipitator (see Chapter 9 for a discussion of particle removal devices). Ash from an FBC system contains a significant amount of CaSO₄, CaO, and CaCO₃ and is more alkaline that that from conventional combustion systems.

**Carbon Monoxide/Hydrocarbons**

Carbon monoxide is the product of incomplete combustion of carbon. It is formed when the oxygen supplied is less than the amount required for stoichiometric combustion of carbon to CO₂, when there is inadequate fuel/air mixing or insufficient residence time for combustion. Fuel reactivity and bed temperature also influence CO emissions. Hydrocarbon emissions are produced under similar conditions. Typical flue gas concentrations are less than 200 ppm for CO and 20 ppm for hydrocarbons in a CFB boiler burning coal [23]. Overfire air—airstaging—is a common technique used for controlling CO and hydrocarbon emissions.

**Trace Elements**

All solid fuels contain small concentrations of trace elements, usually measured in parts per million. Trace elements enter the atmosphere through natural processes,
and sources of trace elements include soil, seawater, and volcanic eruptions. Human activities such as power generation and combustion of fuels in the industrial and commercial sectors also lead to emissions of some elements.

Title III of the U.S. Clean Air Act Amendments of 1990 designates 188 hazardous air pollutants. Eleven trace elements are in the list: antimony (Sb), arsenic (As), beryllium (Be), cadmium (Cd), chromium (Cr), cobalt (Co), lead (Pb), manganese (Mn), mercury (Hg), nickel (Ni), and selenium (Se). In addition, barium (Ba) is regulated by the Resources Conservation and Recovery Act, and boron (B) and molybdenum (Mo) are regulated by Irrigation Water Standards. Vanadium (V) is regulated based on its oxidation state, and vanadium pentoxide (V$_2$O$_5$) is a highly toxic regulated compound. Other elements, such as fluorine (F) and chlorine (Cl), which produced acid gases (i.e., HF and HCl) on combustion, and radionuclides, such as radon (Rn), thorium (Th), and uranium (U), are also of interest.

The distribution of trace elements in the bottom ash, ash collected in the particulate control devices, and fly ash and gaseous constituents emitted into the atmosphere depend on many factors, including the volatility of the elements, temperature profiles across the system, pollution control devices, and operating conditions [48, 49]. Numerous studies have shown that trace elements can be classified into three broad categories based on their partitioning during coal combustion. A summary of these studies is presented by Clarke and Sloss [49], and Figure 7.13 illustrates the classification scheme for selected elements [50].

Class I elements are the least volatile and are concentrated in the coarse residues (i.e., bottom ash) or are equally divided between coarse residues and finer particles (i.e., fly ash). Class II elements will volatilize in the boiler but condense

![Figure 7.13](image-url)  
Figure 7.13 Classification scheme for selected trace elements relative to their volatility and partitioning in power plants.  
*Source:* Adapted from Clarke and Sloss (1992) and Miller et al. (1996).
downstream and are concentrated in the finer-sized particles. Class III elements are the most volatile and exist entirely in the vapor phase. Overlap between the classifications exists and is a function of fuel, combustion system design, and operating conditions, especially temperature [49].

The operating temperature in an FBC boiler is lower than that from a conventional system (i.e., pulverized coal or stoker-fired), which may lead to reduced volatilization of some elements. However, this reduction may be offset to some degree by the longer residence times at a relatively high temperature in the FBC boiler, allowing more volatilization to occur [49].

Mercury, which has become an element of much interest due to its toxicity and recent regulation for coal-fired boilers [51], has been shown to be captured by the limestone that is used for sulfur control in CFBC boilers. The mercury is captured by the fine limestone particles and removed by the particulate control device. According to Hughes and Littlejohn [52], most trace elements associated with the particulates and emissions of the trace elements from the system depend primarily upon the efficiency of the particulate control device. Conventional baghouse filters have an overall removal efficiency of greater than 99 percent. Once a dust cake is formed, the efficiency for removal of even the smaller particles (down to 0.1 μm) can approach 100 percent [49, 53]. The following discussion is evidence of the low mercury emissions rate from CFBC boilers, where the mercury is associated with the limestone particles captured by bagfilters.

In 1999, the U.S. Environmental Protection Agency (EPA) approved an Information Collection Request (ICR) to study mercury emissions from power plants to provide some framework for comparison as a basis for implementing mercury control legislation. Part III of the ICR was to determine speciated mercury emissions from stationary sources. This included power plants of various sizes, systems, and firing configurations throughout the United States. The details of the ICR can be found at the U.S. EPA website [54]. Three CFBC units were included in the ICR: Stockton CoGen Plant, Stockton, California; PG&E Scrubgrass Generating Station, Unit 1, Kennerdell, Pennsylvania; and Tractebel Power Inc., Kline Township Cogeneration Plant Unit 1, New York.

The Stockton plant is a CFB boiler that fired bituminous coal and fluid coke, with mercury concentrations of 26 to 29 ppb and 12 to 45 ppb, respectively, at 55,000 lb/h during the test. Ammonia and limestone were also fed into the system for emissions control. The emissions tests showed no detection of mercury at the baghouse outlet. Detection limits were equivalent to 0.1 mg/dscm (milligram per dry, standard cubic meter), which translates to 0.00005 to 0.00008 lb/h. Mercury was detected in the particulate samples at the baghouse inlet; however, no oxidized or elemental mercury (gas phase) were detected.

The Scrubgrass plant fired waste bituminous coal in a CFB boiler with limestone injection. The fuel was fed at a rate of 75,000 lb/h (3.95 lb Hg/h based on 527 ppb (parts per billion) Hg in the waste coal). Emissions tests showed that 99.8 percent of the mercury was captured with the ash in the baghouse. This corresponds to emissions of 1.39 lb Hg/year. Capture efficiency across the baghouse resulted in a reduction of 99.99, 85.88, and 19.00 percent, respectively, in particle bound mercury, oxidized mercury, and elemental mercury.
Similarly, Tractebel’s plant fired anthracite cleaning wastes (i.e., culm) in a CFB boiler. Test results showed that 99.80 percent of the mercury was captured with the ash in the baghouse. This amounts to approximately 0.53 lb Hg release per year. The average mercury in the fuel during the test was 330 ppb; however, the plant reported that the mercury concentration is normally about 170 ppb, which correlates to emissions of 0.25 to 0.30 lb per year. The baghouse removal efficiency of the particulate bound mercury was 99.995. Oxidized and elemental mercury removal through the baghouse was 43.69 and 91.16 percent, respectively.

Overall, certain trace elements are captured by the bed material, which is dependent on the type of element, bed temperature, and fuel characteristics. The remaining elements will exit the combustor by the flue gases and, as the temperature decreases, condense onto cooler surfaces and fly ash particles, which will be collected by the particulate collection devices. The highly volatile elements will remain in the flue gas or condense on the submicron particles that escape the particulate removal device.

Ash Chemistry and Agglomeration Issues

An important issue that must be assessed in an FBC system is the behavior of the inorganic elements toward bed agglomeration. As previously discussed, the lower bed temperatures are advantageous in that they can be kept below the ash-softening temperatures, especially when firing coals. However, because an FBC boiler can fire a wide range of fuels with varying ash chemistries, the interactions of the ash with the bed material can lead to agglomeration, which has a detrimental effect on fluidization. This is especially true with biomass materials, where care must be taken in the design of the system (e.g., incorporation of kaolin clay injection systems for agglomeration control) or selection of the feedstocks.

It has long been recognized that the mode of occurrence of inorganic elements in fossil fuels has a direct bearing on their behavior during combustion [55–58]. The occurrence of inorganic elements in bio-based fuels is also important, especially as cofiring coal and biomass is becoming increasingly more popular [59–64]. Inorganic species can occur as ion-exchangeable cations, as coordination complexes, and as discrete minerals. In the case of firing a single fuel, such as coal, it is possible to predict ash behavior to avoid system problems. However, it becomes more complex to predict ash behavior in the case of firing multiple fuels in proportions that vary with time—for example, seasonal changes with biomass, and are extremely heterogeneous.

Low-rank coals and biomass materials often contain significant amounts of alkali metals (potassium and sodium) and alkaline earth metals (calcium and magnesium), which are rapidly released into the gas phase and interact with other elements resulting in problems with fouling, slagging and corrosion. In general, potassium and sodium that are associated with the organic structure of the fuel tend to be problematic in that they can contribute to the formation of inorganic phases that have lower melting points. Studies conducted on ash formation during coal combustion show that the incorporation of moderate amounts of alkali and alkaline earth
elements into silicates enhances the coalescence and agglomeration of inorganics due to formation of “sticky” molten phases [56, 57, 65–67].

The presence of low-melting-point phases in a fluidized-bed combustor results in the formation of clinkers that can compromise the bed fluidity. It is also important to recognize that the blending of biomass feedstocks and coal does not necessarily result in simply an additive effect of problematic elements. Changes in the feed blend may or may not have devastating effects on system operation. Predicting these effects is based on understanding the manner in which the inorganics in fuels interact during combustion and their effect on the chemical and physical properties of the ash and gas phases in the system. Their behavior can be predicted from detailed fuel analysis, such as chemical fractionation, to assess the mobility of the inorganic elements and models to predict sintering potential and viscosity behavior of ash produced during combustion.

7.3.6 Supercritical Fluidized-Bed Boilers and Oxy-Coal Firing in Fluidized-Bed Boilers

Power plant efficiency improvements through the use of advanced steam cycles are as big a concern to the FBC boiler manufacturers as to the pulverized coal-fired boiler vendors. Supercritical boiler technology is being implemented on CFB boilers by the major boiler manufacturers to realize efficiency improvements that lead to CO₂ mitigation, savings in fuel costs, reductions in power plant auxiliary energy consumption, and reducing heat losses of the boiler [68].

Similarly, there is growing interest in oxy-fuel combustion in circulating fluidized-bed (CFB) boilers to build upon their inherent advantages over conventional combustion systems discussed in the previous sections [68–70]. Many of the challenges for pulverized coal oxy-fuel boilers are valid for CFB boilers. To date, all oxy-fuel demonstrations (not including pilot-scale test units) use pulverized coal-fired boilers. Projects are underway to demonstrate oxy-fuel firing in CFBs [70].

7.4 Integrated Gasification Combined Cycle

Integrated gasification combined cycle (IGCC) is the most advanced precombustion carbon capture technology available [3, 6–8, 65]. IGCC is the integration of two different technologies: coal gasification from the chemical industry and combined-cycle power generation from the power industry. The characteristics of the IGCC technology make it attractive for CO₂ capture because syngas is generated in the gasifier at elevated pressures and is then chemically converted to enrich the CO₂ concentration, which are two of the driving forces for CO₂ separation. In addition, IGCC technology is able to take advantage of the efficient high-temperature gas turbines that are available. As a result, IGCC can offer high efficiencies, typically in the mid-40s (LHV basis), thereby producing less CO₂ per unit of fuel than lower efficiency systems. In addition to the carbon capture capability, IGCC has the
advantages of using less water, about 33 percent, than a similar-size pulverized coal-fired plant and produce a usable by-product (i.e., a glassy slag used in the manufacture of cement or roofing shingles, or as asphalt filler or aggregate) [71].

7.4.1 Introduction

The basic layout of an IGCC plant is shown in Figure 7.14, which is a simplified flowsheet with precombustion CO$_2$ capture and cold-gas cleanup (modified from [6, 8]). Within this framework, however, there is considerable variation among systems, as is demonstrated by five existing coal-based IGCCs (as of 2009) of the 250 to 350 MW$_e$ class listed in Table 7.2 and discussed later in this chapter. Note that there are 17 IGCC units in operation worldwide, but only five are coal-based [8]. The purpose of this section is to review the choices available within the basic structure and discuss the main systems in the IGCC plant, which are further summarized following. The IGCC process is often divided into blocks or islands, but the definitions of these blocks vary among manufacturers.

The first variation to be considered is the extent of air-side integration, which can range from 0 percent, as in Polk or Wabash, to 100 percent, as in Buggenum or Puertollano, where the degree of integration is defined as the percentage of air supplied to the Air Separation Unit (ASU) by extraction from the gas turbine. This difference in the first generation (1990s) of IGCCs can largely be attributed to the gas turbines available at the time. The 100 percent integration is clearly disadvantageous in today’s economic environment, since a long start-up period using an

![Figure 7.14 Simplified IGCC process flowsheet with precombustion CO$_2$ capture and coal gas cleanup.](source: Modified from Henderson (2003) and Kather et al. (2008).)
expensive backup fuel is required. On the other hand, zero air extraction from the gas turbine air compressor does not allow optimum use of the machine over a full range of ambient conditions. The optimum degree of integration is dependent on many factors, including the ambient temperature range and the turbine selection. A typical figure today might be around 30 percent.

The quality of the oxygen is another variable. Over the range 85 percent (Puertollano) to 95 percent (most other plants), the optimum curve for energy consumption is fairly flat. Additional energy is required in the ASU to raise the quality further. A purity of 99.5 percent O₂ is typically not attractive for a straight IGCC application, but for chemical applications it is generally the standard. Where production of power with one or more chemicals (often known as polygeneration) is to be considered, one would need to review the oxygen purity specification on an individual case. If the coproduct were to be ammonia for instance, 95 percent O₂ would be acceptable.

### 7.4.2 Gasification Island

Variations in the gasification island are almost entirely dependent on the choice of technology supplier. Important differences are discussed in this section. In feedstock preparation, rod mills are typically used for slurry preparation (e.g., GE Energy or ConocoPhillips) or roller mills and drying are used in dry-feed systems (e.g., Shell or Siemens). In both cases, the particle size is of the order of magnitude of less than 100 μm. In the case of fluid-bed processes, the particle size is much larger: about 6 mm.

More specifically, in a dry-feed system, the coal is ground and dried in a roller mill with a hot gas drying circuit, similar to those used in conventional pulverized
coal units. The pulverized coal is then fed through a lock hopper system into the pressurized feed vessel. The coal is then transported to the burners from the feed vessel by pneumatic conveying in the dense phase. The carrier gas is typically pure nitrogen from the ASU. The lock hopper system, which relies on gravity flow to move the coal from the uppermost, atmospheric bunker through the lock hopper into the feed vessel requires a support structure that can be as tall as the gasifier itself, so there is great interest in the development of a “solids pump” that could reduce the cost of the feed system.

For wet-feed systems, the slurry is made in a rod mill into which precrushed coal and water are fed. The coal is ground in a wet milling process to a size of about 100 μm. The product slurry is sieved to remove oversize material, which can be discarded or recycled according to need. The slurry is pumped to the reactor pressure typically with a membrane piston pump.

For an entrained-flow gasifier, the flow direction can be downflow (e.g., GE Energy or Siemens) or upflow (e.g., ConocoPhillips or MHI). The temperature containment can be with refractory (e.g., GE Energy or ConocoPhillips) or using a membrane wall (Shell or MHI). ConocoPhillips and MHI use a two-stage gasifier. The other technology suppliers use single-stage gasification. The oxidant can be oxygen (e.g., Shell ConocoPhillips, Prenflo, or GE Energy) or air (e.g., MHI).

Syngas cooling is available in a number of variations. Water quench (GEE or Siemens) is not currently used in the coal-based IGCC configuration, primarily because of the associated efficiency penalty, whereas it is used in chemical applications, particularly where CO shift for hydrogen manufacture is involved (Kingsport, Tennessee; Coffeyville, Kansas). It is also used in a number of refinery-based IGCC units. Should CO₂ capture be implemented, then this cooling technique would probably be favored also for coal-based IGCC applications.

Of the different steam-raising configurations, radiant cooling is only offered by GE Energy (e.g., Polk). ConocoPhillips uses a firetube convection cooler after the second stage of its E-Gas gasifier. GE Energy has also used a firetube convection cooler as a second cooling stage in Polk, but it has been deleted in current designs. Shell uses a gas quench and watertube syngas cooling.

A certain amount of syngas pretreatment is generally included in the scope of the gasification technology supplier. This includes removal of particulate matter and a number of trace components in the gas, particularly ammonia and chlorides. Shell and ConocoPhillips remove the particulates and the water-soluble gases in separate stages, using a candle filter (sinter metal for ConocoPhillips and ceramic for Shell) for particulate removal, and a water wash for ammonia and chlorides. GEE combines these steps in a single scrubber. Slag removal from the pressurized gasifier is achieved using a lock hopper arrangement in most processes. ConocoPhillips has a proprietary continuous letdown system.

7.4.3 Gas Treatment and Sulfur Recovery

Although sulfur species (primarily H₂S) are the principal targets of the gas treatment system, it is necessary to consider the full range of potential contaminants, which
include COS (a minor sulfur species) and mercury. Depending on the selection of desulfurization technology, COS will probably need to be hydrolyzed to H$_2$S to achieve the required level of sulfur removal. Typical temperatures for COS hydrolysis are between 320°F and 390°F (160°C and 200°C).

Mercury removal is best performed at ambient temperatures upstream of the acid gas removal, so some of the gas treatment will need to be integrated with the low-temperature gas cooling. Mercury removal from syngas has only been practiced industrially at Kingsport, although it is a regular feature of natural gas pretreatment in LNG plants.

There is an extremely wide variety of acid gas removal (AGR) systems on the market. These can be classified as chemical washes (which include all amines such as methyldiethanolamine (MDEA)) and physical washes such as Selexol or Rectisol. In addition, it is possible to have a mixed characteristic solvent such as Sulfinol. All of these named processes have been used in IGCC or chemical plant gasification operations. Selection is based on requirements for high purity (Rectisol) versus low cost (MDEA), with Selexol and Sulfinol lying in between on both counts. All these processes have a long track record in industrial practice, all with high availability records.

**Chemical Solvent Processes**

Figure 7.15 shows the flow sheet of a typical MDEA wash, although this flow sheet is representative of many other chemical washing processes.

**Amine Processes**

The raw syngas is contacted in a wash column with lean MDEA solution, which absorbs the H$_2$S and some of the CO$_2$. MDEA is to some extent selective in that

Figure 7.15 A typical MDEA flowsheet.
the bonding of the amine with H₂S takes places faster than with CO₂ and advantage can be taken of this in the design. The rich solution is preheated by heat exchange with the lean solution and enters the regenerator. Reboiling breaks the chemical bind and the acid gas components discharged at the top of the regenerator are cooled to condense out the water, which is recycled.

**Physical Solvent Processes**

**Physical Washes**
The following are important characteristics for any successful physical solvent:

- There should be good solubility for CO₂, H₂S, and COS in the operating range, preferably with significantly better absorption for H₂S and COS compared with CO₂ if selectivity is an important issue for the application of interest.
- Low viscosity at the lower end of the operating temperature range. Although lowering the operating temperature increases the solubility, the viscosity governs, in effect, the practical limit to lowering the operating temperature.
- A high boiling point reduces vapor losses when operating at ambient or near ambient temperatures.

**Selexol**
The Selexol process was originally developed by Allied Chemical Corporation and is now owned by UOP. It uses dimethyl ethers of polyethylene glycol (DMPEG). The physical properties of DMPEG are listed in Table 7.3. The typical operating temperature range is 15 to 100°C or 41°F to 212°F. The ability to operate in this temperature range offers substantial cost benefits by eliminating or minimizing refrigeration duty. On the other hand, for a chemical application such as ammonia, the residual sulfur in the treated gas may be 1 ppm H₂S and COS each after the CO₂ wash. This is, however, not an issue in power applications where the sulfur slip is less critical.

The ratio of absorption coefficients for H₂S, COS, and CO₂ is about 1:4:9 in descending order of solubility. A plant designed for 1 ppm COS in the clean gas would

### Table 7.3 Properties of Physical Solvents

<table>
<thead>
<tr>
<th>Process</th>
<th>Selexol</th>
<th>Rectisol</th>
<th>Methanol</th>
</tr>
</thead>
<tbody>
<tr>
<td>Solvent</td>
<td>DMPEG</td>
<td>Methanol</td>
<td>CH₃OH</td>
</tr>
<tr>
<td>Formula</td>
<td>CH₃O(CH₂CH₂O)₆CH₃</td>
<td>CH₃OH</td>
<td>CH₃OH</td>
</tr>
<tr>
<td>Molecular weight</td>
<td>lb/lb mol</td>
<td>178 to 442</td>
<td>32</td>
</tr>
<tr>
<td>Boiling point at 760 Torr</td>
<td>°F</td>
<td>415 to 870</td>
<td>147</td>
</tr>
<tr>
<td>Melting point</td>
<td>°F</td>
<td>−4 to −20</td>
<td>−137</td>
</tr>
<tr>
<td>Viscosity</td>
<td>cP</td>
<td>4.7 at 86°F</td>
<td>0.85 at 5°F</td>
</tr>
<tr>
<td></td>
<td>cP</td>
<td>5.8 at 77°F</td>
<td>1.4 at −22°F</td>
</tr>
<tr>
<td></td>
<td>cP</td>
<td>8.3 at 59°F</td>
<td>2.4 at −58°F</td>
</tr>
<tr>
<td>Specific weight</td>
<td>kg/m³</td>
<td>1.031</td>
<td>790</td>
</tr>
<tr>
<td>Selectivity at working temperature</td>
<td>(H₂S:CO₂)</td>
<td>1:9</td>
<td>1:9.5</td>
</tr>
</tbody>
</table>
require about four times circulation rate of a plant for 1 ppm H₂S, together with all the associated capital and operating costs. In a gasification environment, it is therefore preferable to convert as much COS as possible to H₂S upstream of a Selexol wash. In a plant using raw gas shift for hydrogen or ammonia, this will take place simultaneously on the catalyst with the carbon monoxide shift. Where no CO shift is desired, COS hydrolysis upstream of the Selexol unit provides a cost-effective solution to the COS issue. Other characteristics that are favorable for gasification applications include high solubilities for HCN and NH₃ as well as for nickel and iron carbonyls.

The Selexol flowsheet in Figure 7.16 exhibits the typical characteristics of most physical absorption systems. The intermediate flash allows coabsorbed syngas components (H₂ and CO) to be recovered and recompressed back into the main stream. For other applications such as H₂S concentration in the acid gas or separate CO₂ recovery using staged flashing, techniques not shown here may be applied.

Rectisol
The Rectisol process, which uses cold methanol as a solvent, was originally developed to provide a treatment for gas from the Lurgi moving-bed gasifier, which in addition to H₂S and CO₂ contains hydrocarbons, ammonia, hydrogen cyanide, and other impurities. Figure 7.17 contains a flowsheet of the Rectisol process.

In the typical operating range of −20 to −75°F, the Henry’s law absorption coefficients of methanol are extremely high, and the process can achieve gas purities

![Figure 7.16 A Selexol flowsheet for selective H₂S removal.](image-url)
unmatched by other processes. This has made it a standard solution in chemical applications, such as ammonia, methanol, or methanation, where the synthesis catalysts require sulfur removal to less than 0.1 ppm. This performance has, however, a price in that the refrigeration duty required for operation at these temperatures involves considerable capital and operating expense.

Methanol as a solvent exhibits considerable selectivity, as can be seen earlier in Table 7.3. This allows substantial flexibility in the flowsheeting of the Rectisol process, and both standard (nonselective) and selective variants of the process are regularly applied according to circumstances.

As a physical wash, which uses at least in part flash regeneration, part of the CO₂ can be recovered under an intermediate pressure. Typically, with a raw gas pressure of 710 psig, about 60 to 75 percent of the CO₂ would be recoverable at 40 to 60 psig. Where CO₂ recovery is desired, whether for urea production in an ammonia application or for sequestration, this can provide significant compression savings.

Following the solvent circuit is an intermediate H₂S flash from which coabsorbed hydrogen and carbon monoxide are recovered and recompressed back into the raw gas. The flashed methanol is then reheated before entering the Hot Regenerator. Here the acid gas is driven out of the methanol by reboiling, and a Claus gas with an H₂S content of 25 to 30 percent (depending on the sulfur content of the feedstock) is recovered. Minor adaptations are possible to increase the H₂S content if desired.

Water entering the Rectisol unit with the syngas must be removed, and an additional small water-methanol distillation column is included in the process to cope with this. Typically the refrigerant is supplied at between −20°F and −40°F. Depending on application, different refrigerants can be used. In an ammonia plant,
ammonia is used, and the refrigeration system is integrated with that of the synthesis. In a refinery environment, propane or propylene may be the media of choice.

The Rectisol technology is capable of removing not only conventional acid gas components but also, for example, HCN and hydrocarbons, and metallic mercury.

**COS Hydrolysis**

The chemical washes are generally not capable of absorbing COS, which must be converted to H₂S in a COS hydrolysis step upstream of the wash. Physical washes can absorb COS. In the case of Selexol, this capability is not very strong, and economics usually dictate the use of a COS hydrolysis as well (but not after a CO shift as in Coffeyville). Rectisol does not require any upstream COS hydrolysis.

**Sulfur Recovery**

Sulfur recovery is generally achieved using Claus technology, although Polk is an exception in that it manufactures sulfuric acid rather than elemental sulfur. Differences in the Claus technology itself are generally only of a detailed nature. Considerable variety is shown in the handling of the tailgas from the Claus plant, which in addition to H₂S also contains small quantities of SO₂, COS, CS₂, and elemental sulfur. In all plants, these are hydrogenated back to H₂S over a catalyst. In some plants, this gas is then treated separately in another washing stage and then incinerated and discharged to the atmosphere. In others it is recycled to a point and mixed with the raw gas upstream of the main AGR so that this remaining gas is treated there. The point at which the recycle is fed into the main gas stream varies.

In all plants, syngas dilution is used to reduce the NOₓ emissions from the gas turbine. The dilution medium may be nitrogen only (e.g., Polk initially), steam only (e.g., Wabash), or a combination of the two (e.g., Buggenum). Steam is generally added by saturation using low-level heat to provide the necessary hot water. In some cases it is added by direct injection.

### 7.4.4 Combined Cycle Power Plant

The combined cycle power plant covers the typical scope of a natural gas combined cycle system complete with balance of plant. The principal difference lies in the use of syngas as a fuel. Note that to date little experience is available on the use of selective catalytic NOₓ reduction (SCR) in IGCCs, where the residual sulfur content in the syngas could impact on the availability of the HRSG. Two units in Italy are equipped with SCRs, but the required NOₓ emissions levels are not comparable with the values required of a NGCC plant. In one case, the SCR is only used when the gas turbine operates on the backup distillate fuel. The only significant experience with SCR in IGCC is at the Negishi plant of Nippon Oil. Values of less than 2.6 ppm NOₓ and less than 2.0 ppm SOₓ have been reported [72].

### 7.4.5 IGCC with Carbon Capture

Integration of carbon capture into an IGCC is relatively simple using techniques common in, for example, the fertilizer industry, where CO₂ is captured in the
ammonia plant and compressed to typically 2,200 to 2,900 psig for urea manufacture. The principle changes to an IGCC power plant without carbon capture are the introduction of a CO shift reactor after the particulate removal and the modification to the acid gas removal to draw off the CO$_2$ stream separately from the H$_2$S. The configuration up to the outlet of the AGR is a regular feature of coal-based ammonia plants. The CO shift uses a sulfided chrome-molybdenum catalyst. The selective AGR is generally a physical wash. This has the added advantage that much of the CO$_2$ is available under pressure, which reduces compression costs. One well-documented example is the 1,000 t/d (100 MW$_e$ equivalent) ammonia plant in Coffeyville, Kansas, which feeds petroleum coke into a GEE Quench Gasifier and uses Selexol for acid gas removal. It achieves on-steam times of 98 percent between annual turnarounds [73]. There are a number of basically similar plants in China using the Shell gasifier with Rectisol.

The main area where industrial experience is limited is on the use of hydrogen as fuel in a gas turbine, although there is more experience available than generally appreciated, much of it in industrial applications. One example described by GE as the “H2 Fleet Leader” is a frame 6B unit operating regularly on 85 to 97 percent hydrogen at “an availability of 96.5 percent+ running in uninterrupted operation, 24 hours a day over the year” since 1997 [74].

7.4.6 Benefits and Limits of IGCC

Several benefits and limits of IGCC warrant discussion. The most important ones are associated with system efficiency, environmental impact, pollutant emissions, system availability, and capital requirements.

**Efficiency**

As previously mentioned, one of the motivations for development of the IGCC power plant was to harness the high efficiency of combined cycle gas turbines for use with coal. The prototype 100 MW$_e$ Cool Water IGCC operated between 1984 and 1988 with a heat rate of 10,950 HHV Btu/kWh net [75]. Wabash, which is representative of the 250 MW$_e$ class of IGCC built in the mid-1990s and based on the GE 7FA gas turbine, had a heat rate of 8,900 HHV Btu/kWh net [76]. The 630 MW$_e$ class currently being planned is based on two larger GE 7FB or Siemens SGT6-5000F gas turbines with a heat rate of about 8,500 HHV Btu/kWh net on bituminous coals. Dry-feed gasifiers achieve a similar heat rate when operating on high-moisture subbituminous coals such as those from the Powder River Basin. The heat rate of a slurry feed gasifier operating with PRB coals increases to 9,000 HHV Btu/kWh net or more.

IGCC efficiencies are typically between 40 and 50 percent with recent turbine technology [7]. Net efficiencies of the Buggenum and Puertollano plants are 43 percent and 45 percent (LHV basis), respectively. Further increases in efficiency
(with projections of net efficiencies of 56 percent (LHV basis) are possible through the following [7]:

- Optimizing the integration of the gasification and power-generation blocks
- Advances in gas turbine technology, such as higher pressure ratios, higher turbine entry temperatures, and reheat
- Further development of the hot gas cleaning processes for the raw gas leaving the gasifier
- Advanced air separation technologies

**Environmental Impact**

Another motivation for development of the IGCC power plant was its potential for extremely low environmental emission rates compared with other coal-based technologies. The first-generation IGCC plants built during the 1990s have achieved the goals set at the time, which are discussed under the individual pollutants following. The technology to reduce these rates further is available, though its use will depend on a balance between cost and regulatory requirements.

**Sulfur Emissions**

An IGCC can readily reduce sulfur emissions to about 4 ppm SO$_2$ (wet basis, referred to 1 percent O$_2$) in the turbine exhaust. This is equivalent to about 30 ppm total sulfur (H$_2$S + COS) in the dry, undiluted gas leaving the acid gas removal, a value achievable with an MDEA or Selexol system. If a selective catalytic reactor (SCR) is required for Denox, then this would typically need to be reduced to about half this value, which would also be possible with a rather more elaborate version of Selexol. A further two orders of magnitude reduction would be possible using Rectisol instead of the currently used chemical or synthetic fuel applications. This is, however, considered to be unnecessarily expensive.

**NO$_x$ Emissions**

The high hydrogen content of syngas (which would be even higher in the carbon capture scenario) prohibits the use of current dry low NO$_x$ burners as developed for natural gas. Instead, diffusion burners are used. Typically one could achieve about 15 ppm (dry basis referred to 15 percent O$_2$) in the exhaust of a gas turbine without SCR. In fact, some IGCCs such as the plant in Buggenum regularly achieve values under 10 ppmv without SCR.

Some areas may, however, require SCR to achieve lower NO$_x$ emission rates of 3 ppmv (dry basis referred to 15 percent O$_2$) comparable with those for natural gas fired turbines. In such a case, it is necessary to desulfurize the fuel gas further to about 15 ppmv to avoid formation of ammonium bisulfate from ammonia slip in the SCR, which can deposit on heat exchange surface downstream the SCR. The SCR in the oil-fired IGCC at Negishi in Japan is reported as meeting its permit level of 2.6 ppmv (dry basis referred to 15 percent O$_2$) [72].

**Mercury**

Mercury can be removed from the fuel gas with a fixed bed of activated carbon. Approximately 95 percent of the mercury leaving the gasifier in the fuel gas is captured.
Other Emissions
Typical values for other pollutants referred to the gross heat input to the gasifier, based on a heat rate of 8,500 Btu (HHV)/kW (net), are as follows:

- Particulate matter: 0.0145 lb/million Btu (including condensables)
- CO: 10–25 ppm (dry basis referred to 15 percent O₂)
- Unburned hydrocarbons: 7 ppmv (wet basis)
- VOC: 1.4 ppmv (wet basis)

Availability
It is generally acknowledged that while early IGCC plants met their efficiency and environmental goals, the availability results were not good [77]. This was also surprising to those in the industry who were accustomed to high availabilities such as those achieved by, say, Eastman at their Kingsport methanol plant, where 98 percent is regularly reported. An analysis of the causes of outage revealed some unexpected results [78]. Much of the lack of availability was due to fleet issues on early models of the gas turbines involved (in some cases up to 25 percent loss of annual availability), which had no relation to their utilization in an IGCC environment. This is contrasted with three refinery-based IGCC units built in Italy about five years later, which after a two- to three-year ramp-up period are reporting availabilities (and on stream times) of 90 to 95 percent. One of these plants achieved more than 90 percent availability in its second year of operation.

Capital Requirements
A discussion on capital costs of IGCC with numerical values is difficult in a period of rapid inflation in the capital plant industry. Estimates vary significantly, depending on a large number of factors, including gasification technology used, rank of coal, and level of built-in redundancy. It is, however, generally accepted that current IGCCs require between 10 and 20 percent additional investment when compared with conventional technology. Existing IGCCs are all tailor-made units with all the associated high costs of engineering, procurement, and construction. Vendors of IGCC systems have recognized the necessity of reducing costs through modularization and standardization and are all preparing so-called “reference designs” for two-train 630 MW net output plants. One vendor has estimated that current efforts could reduce the cost premium to about 10 percent [78].

Introduction of syngas operation to the next generation of gas turbines (H-class), which have now started operation with natural gas, will also reduce the specific costs per kW installed capacity by extracting a higher-power output from the same gas production facility.

It should also be noted that CO₂ capture from the high-pressure fuel gas is much less costly from flue gas and that in the event of carbon capture becoming a necessity, the overall cost of electricity is expected to be lower with IGCC than with conventional technologies and postcombustion capture. This is illustrated in Table 7.4, which lists CO₂ emissions and costs from an Electric Power Research Institute (EPRI) study for different technologies (modified from [79]).
7.4.7 Commercial Status

As previously mentioned and listed in Table 7.2, five coal-fueled IGCC units are in operation worldwide, with capacities of 250 to 350 MWₑ [6, 71]. Each of these units uses a different gasifier technology. Four of the five units use oxygen for gasification, and one uses air. Two of these are slurry-fed gasifiers. This section presents some of the key information of the units.

**Peurtollano (Peurtollano, Spain)**

The Prenflo gasifier-based IGCC (300 MWₑ) at Peurtollano, Spain, owned by ELCOGAS has been operating since 1996 [6, 8]. Oxygen is used for gasification. The plant has a Siemens V94.3 gas turbine. The raw product gas is cooled to 380°C in an integral syngas cooler above the gasification zone. Ceramic filters operating at about 240°C remove particulates before the scrubber. Quench gas is recycled to cool the raw gas to 800°C before the syngas cooler.

Initially, difficulties were encountered with fuel and slag flow and the gas turbine combustor, which resulted in intermittent operation (i.e., low availability). This was the result, in part, of the feed, which is a mixture of high-ash coal and high-sulfur petroleum coke. The combustor was modified based on experience from Buggenum, and since 2000, improved syngas availability has been achieved by restricting the proportion of coal fed to 38 percent by mass and the use of better offline cleaning cycles for the hot gas cleanup system, which are candle filters. Availability is approaching 75 percent, which is still below the 85 percent average availability of the natural gas combined cycle and the pulverized coal-fired power plants [8].

<table>
<thead>
<tr>
<th>Technology: Carbon Capture and Sequestration</th>
<th>Supercritical Pulverized Coal</th>
<th>Ultra-Supercritical Pulverized Coal</th>
<th>Oxy-Fuel PC</th>
<th>IGCC</th>
</tr>
</thead>
<tbody>
<tr>
<td>without CCS</td>
<td>830</td>
<td>738</td>
<td>104</td>
<td>824</td>
</tr>
<tr>
<td>with CCS</td>
<td>109</td>
<td>94</td>
<td>824</td>
<td>101</td>
</tr>
<tr>
<td>Efficiency, % (HHV)</td>
<td>38.5</td>
<td>43.4</td>
<td>30.6</td>
<td>38.4</td>
</tr>
<tr>
<td>Total capital requirement, $/kW</td>
<td>1,937</td>
<td>1,976</td>
<td>2,990</td>
<td>2,080</td>
</tr>
<tr>
<td>Cost of electricity, ¢/kWh</td>
<td>5.50</td>
<td>5.39</td>
<td>7.93</td>
<td>5.90</td>
</tr>
</tbody>
</table>

Source: Modified from Booras and Holt (2004).

Table 7.4 CO₂ Emissions, Efficiency, and Costs of Advanced Power-Generation Technologies with and without CCS
**Polk Power Station (Tampa, Florida)**

The Global Energy’s E-Gas two-stage, upflow slurry feed gasifier-based IGCC (250 MW_e) at the Polk Power Station, owned by Tampa Electric, has been operating since 1996 [6, 8]. Oxygen is used for gasification. The feed is a mixture of coal and petroleum coke. The gas emerging from the radiant cooler is cooled in two parallel firetube boilers and by heat exchange with the cleaned gas after desulfurization in MDEA scrubbers. The plant has a General Electric STAG 107FA gas turbine combined cycle system.

The IGCC system availability is typically 60 to 70 percent. Carbon conversion has been lower than anticipated, low to mid-90s versus about 98 percent, as expected. The resultant heat rate penalty has been around 200 to 500 kJ/kWh [6].

**Wilem-Alexander Cenrale (Buggenum, Netherlands)**

The Shell entrained-flow gasifier-based IGCC (250 MW_e) at Beggenum, owned by NUON Power, has been operating commercially in the Netherlands since 1998 [6, 8]. Oxygen is used for gasification. The plant has a Siemens V94.2 gas turbine. The membrane-walled gasifier operates at a temperature of 1,500°C and pressure of 2.8 MPa. The raw gas is quenched to 900°C at the exit at the top of the gasifier by addition of a recycle stream of cooled, ash-free gas before being sent to a convective syngas cooler. The gas then passes to a cyclone then a ceramic filtration unit at 250 to 285°C. Final cleaning is by cold scrubbing to remove ammonia, chlorides, and sulfur gases, after which the gas is saturated with water for NO_x control and sent to the gas turbine for combustion [6]. IGCC system availability is approximately 60 percent and has been operating at this level for several years [8].

**Wabash River (W. Terre Haute, Indiana)**

The ConocoPhilips E-Gas (formerly Dow Chemical then Dynergy then Global Energy) gasifier-based IGCC (260 MW_e) at Wabash River, Indiana, owned by SG Solutions/Duke Energy Indiana has been operating since 1999 [6, 8]. Oxygen is used for gasification. The feedstocks are a high-sulfur Illinois Basin bituminous coal and petroleum coke. The plant has a General Electric Frame 7FA gas turbine. Raw gas from the top of the gasifier is cooled to 370°C in a vertical firetube syngas cooler. A candle filter unit removes fly ash and char for recycle to the gasifier. The filtered gas is then cooled and cleaned of sulfur compounds in an MDEA scrubber. The clean flue gas is reheated and moisture added to it for control of NO_x emissions before combustion in the gas turbine. IGCC system availability is approximately 70 percent, which is among the highest levels any of the five coal-fueled systems.

**Clean Coal Power R&D Company (Nakoso, Japan)**

The MHI air-blown, two-stage entrained-flow gasifier-based IGCC (250 MW_e) in Nakoso, owned by a consortium of Japanese utilities, corporations, and the Ministry of Economy, Trade, and Industry has been operating since September 2007 and as
of April 2009 has accumulated more than 2000 hours of operation [80]. The plant uses filtration for particulates removal followed by scrubbing with MDEA for desulfurization [6]. The system is new with limited operational data available; however, target net efficiency is 42 percent (LHV basis) and target emissions are 8 ppm SO2, 5 ppm NOx, and 4 mg particulate/m3.

### 7.5 IGCC Research Needs

Several areas of research and development are required to further the IGCC technology and the integration of CO2 capture and storage [7]. These are generally divided into areas of availability and reliability, modeling and simulation, process and components, and precombustion capture. With respect to availability and reliability, both need to be increased; there must be an increase in robustness by preventing slagging, fouling, and corrosion; and dry-feed systems need optimizing or new ones must be developed. Research needs for modeling and simulation include buildup of databases for substance and process modeling, developing models for optimizing gasification, and modeling and simulation of gasifiers.

In the area of process and components, avenues of research include developing methods to use high-temperature heat without decreasing availability, improving hot gas cleanup technologies, optimizing gas turbines for syngas, increasing gas turbine inlet temperatures, optimizing flowsheets, and reducing capital costs. Research needs related to precombustion capture in IGCC power plants include developing an optimized hydrogen-fueled gas turbine; optimizing the process by integrating the ASU, CO-shift, and CO2 capture technologies; developing dynamic modeling of the entire process; and performing analysis of part-load ability and variable CO2 capture.

### References


