Midwest Geological Sequestration Consortium

# Carbon Dioxide Capture and Transportation Options in the Illinois Basin

## Topical Report

## Assessment of Geological Carbon Sequestration Options in the Illinois Basin

## Task 2: Assess Carbon Capture Options for Illinois Basin Carbon Dioxide Sources

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# Disclaimer

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# Foreword

This Topical Report represents a key Year 1 deliverable of An Assessment of Geological Carbon Sequestration Options in the Illinois Basin by the Midwest Geological Sequestration Consortium (MGSC). It includes the results of Task 2, Assess Carbon Capture Options for Illinois Basin $CO_2$ Sources, and Task 3, Assess $CO_2$ Transportation Options in the Illinois Basin. The work of the MGSC is focused on defining the most promising capture-transportation-storage options in the Illinois Basin with an exclusive focus on geological options for storage. The capture options assessment included development of a comprehensive catalog of fixed emission sources with emphasis on the fuel type and plant characteristics for major power plants. We found for example, that the majority of power plant emissions came from a surprisingly small number of plants. Retrofit carbon capture technologies were compared for Powder River Basin coal and Illinois Basin coal, and conventional pulverized coal plants were compared to integrated gasification combined cycle plants. Complete analyses of costs and benefits were undertaken.

Transportation of large volumes of $CO_2$ captured at fixed emissions sources will no doubt require pipeline transportation to any large-capacity geological storage location. Development of such infrastructure will be costly and will require careful planning optimized to the region under consideration. Task 3 results are focused on pipeline transportation attributes that would impact such infrastructure development in the Illinois Basin. Short-term transportation will also be needed to carry out tests for research purposes and to carry out operational proof-of-concept testing, such as the adaptation of enhanced oil recovery practices to Illinois Basin conditions through the development of commercial pilot tests of $CO_2$ reservoir flooding. The short-term transportation options will consist of some combination of truck and rail transport involving quantities from a few thousand tons to a few tens of thousands of tons.

A major effort in the second year of MGSC work will be completion of the assessment of coal beds, mature oil reservoirs, and deep, saline reservoirs for geological storage (Tasks 4, 5, and 6 of the project work plan). Once these tasks are completed, a subsequent task (Task 7) will integrate geological storage options with the capture-transportation assessments described herein. The result will be a preliminary sequestration scenario combining favorable technical and economic pathways applicable within the Illinois Basin as a basis for future assessment and planning.

# Abstract

This report describes carbon dioxide ($CO_2$) capture options from large stationary emission sources in the Illinois Basin, primarily focusing on coal-fired utility power plants.

The $CO_2$ emissions data were collected for utility power plants and industrial facilities over most of Illinois, southwestern Indiana, and western Kentucky. Coal-fired power plants are by far the largest $CO_2$ emission sources in the Illinois Basin. The data revealed that sources within the Illinois Basin emit about 276 million tonnes of $CO_2$ annually from 122 utility power plants and industrial facilities. Industrial facilities include 48 emission sources and contribute about 10% of total emissions.

A process analysis study was conducted to review the suitability of various $CO_2$ capture technologies for large stationary sources. The advantages and disadvantages of each class of technology were investigated. Based on these analyses, a suitable $CO_2$ capture technology was assigned to each type of emission source in the Illinois Basin.

Techno-economic studies were then conducted to evaluate the energy and economic performances of three coal-based power generation plants with $CO_2$ capture facilities. The three plants considered were (1) pulverized coal (PC) + post combustion chemical absorption (monoethanolamine, or MEA), (2) integrated gasification combined cycle (IGCC) + pre-combustion physical absorption (Selexol), and (3) oxygen-enriched coal combustion plants. A conventional PC power plant without $CO_2$ capture was also investigated as a baseline plant for comparison. Gross capacities of 266, 533, and 1,054 MW were investigated at each power plant. The economic study considered the burning of both Illinois No. 6 coal and Powder River Basin (PRB) coal. The cost estimation included the cost for compressing the $CO_2$ stream to pipeline pressure.

A process simulation software, CHEMCAD, was employed to perform steady-state simulations of power generation systems and $CO_2$ capture processes. Financial models were developed to estimate the capital cost, operations and maintenance cost, cost of electricity, and $CO_2$ avoidance cost. Results showed that, depending on the plant size and the type of coal burned, $CO_2$ avoidance cost is between \$47/t to \$67/t for a PC +MEA plant, between \$22.03/t to \$32.05/t for an oxygen combustion plant, and between \$13.58/t to \$26.78/t for an IGCC + Selexol plant. A sensitivity analysis was conducted to evaluate the impact on the $CO_2$ avoidance cost of the heat of absorption of solvent in an MEA plant and energy consumption of the ASU in an oxy-coal combustion plant.

An economic analysis of $CO_2$ capture from an ethanol plant was also conducted. The cost of $CO_2$ capture from an ethanol plant with a production capacity of 100 million gallons/year was estimated to be about \$13.92/t.

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# Part 1 Capture of Carbon Dioxide from Utility and Industrial Stationary Sources in the Illinois Basin

### M. Rostam-Abadi, S.S. Chen, and Y. Lu

## 1. Introduction

Research by the U.S. Department of Energy (U.S. DOE) and International Energy Agency has suggested that carbon sequestration can play an important role in reducing carbon dioxide ($CO_2$) in the atmosphere in the first part of the twenty-first century (Reichle et al., 1999; Gielen, 2003). Although carbon sequestration processes are for the most part technically feasible, their cost-effectiveness must be significantly improved before they become practical.

The $CO_2$ sequestration processes include $CO_2$ capture and separation, transportation, and storage. The capture and separation of $CO_2$ from its emission sources is the first, most expensive step, accounting for more than 60% of the total cost (Herzog et al., 1997; Beecy, 2002). Therefore, reducing the cost of the capture step significantly improves the economics of the whole sequestration process. Table 1.1 lists some of the characteristics of power plants with and without $CO_2$ capture (Gielen, 2003). With $CO_2$ capture, the energy efficiency penalty is between 6% for a future (2020) coal-based integrated gasification combined cycle (IGCC) and 12% for a near-term (2010) pulverized coal (PC) boiler. The capital investment and cost of electricity also increases significantly. The energy efficiency penalty is even higher for current PC power plants. For example, one analysis (Dave et al., 2000) shows that the energy efficiency penalty could be as high as 15% (reduced from 37% to 22%).

**Table 1.1. Characteristics of power plants with and without $CO_2$ capture**

| | | Capital investment ($/kW) | | Efficiency (%) | | | $CO_2$ capture efficiency (%) | Electricity costs (mils/ kWh) | | |
|---|---|---|---|---|---|---|---|---|---|---|
| Technologies | Start year | Without $CO_2$ capture | With $CO_2$ capture | Without $CO_2$ capture | With $CO_2$ capture | Loss (%) | | Without $CO_2$ capture | With $CO_2$ capture | Capture costs ($/t $CO_2$) |
| Coal steam cycle (CA)[1] | 2010 | 1,075 | 1,850 | 43 | 31 | –12 | 85 | 29.1 | 51.0 | 24 |
| Coal steam cycle (Membrane + CA) | 2020 | 1,025 | 1,720 | 44 | 36 | –8 | 85 | 29.2 | 46.3 | 21 |
| Coal IGCC (Selexol) | 2010 | 1,455 | 2,100 | 46 | 38 | –8 | 85 | 37.4 | 52.3 | 20 |
| Coal IGCC (Selexsol) | 2020 | 1,260 | 1,635 | 46 | 40 | –6 | 85 | 33.0 | 41.0 | 11 |
| Natural Gas CC (CA) | 2005 | 400 | 800 | 56 | 47 | –9 | 85 | 26.1 | 36.8 | 29 |
| Natural Gas CC (Selexol) | 2015 | 400 | 800 | 59 | 51 | –8 | 85 | 25.2 | 34.8 | 25 |

[1] CA, chemical absorption; CC, combined cycle; IGCC, integrated gasification combined cycle. Comparison based on 10% discount rate, 30-year plant lifespan. Coal price, 1.5 US$/GJ; gas price, 3 US$/GJ. $CO_2$ product is in a supercritical state at 100 bars. $CO_2$ transportation and storage are not included. $CO_2$ capture costs are expressed relative to the same power plant without capture.

The cost data in Table 1.1 are the $CO_2$ capture costs. $CO_2$ avoidance cost is commonly used to compare various capture technologies. The $CO_2$ avoidance cost is the ratio of the increase in electricity generating costs to the specific $CO_2$ emissions (relating to the electrical energy output). The installation of $CO_2$ capture equipment incurs both the capital cost and an energy efficiency penalty. Due to the low energy efficiency of current power generation technologies, the same degree of energy efficiency penalty translates to a much higher $CO_2$ avoidance cost. Obviously, it is important to identify cost-effective $CO_2$ capture and separation technology for specific power production processes.

## 2. $CO_2$ Emission Sources in the Illinois Basin

Table 2.1 lists data for $CO_2$ emissions from the stationary sources for the United States and the Illinois Basin. In the United States, roughly about one third of the $CO_2$ emissions is from electric power generation, one third from transportation, and one third from other sources including manufacturing industry, commercial, and residential (Energy Information Administration, 2002). The emissions from the manufacturing industry sector, which includes oil refineries, the steel industry, the cement industry, and other industries, represent about 20%.

**Table 2.1. $CO_2$ Emissions in the United States and Illinois Basin**

| Sources | U.S. total tonnes | Illinois Basin tonnes | Basin to U.S. (%) | Industry (% of Basin) |
|---|---|---|---|---|
| **Power generation** | 2,239,700,000[1] | 254,260,000[2] | 11.4 | 92.1 |
| Coal<br>Natural gas<br>Oil | 1,868,400,000[1]<br>299,100,000[1]<br>72,200,000[1] | 249,216,000[2]<br>4,996,000[2]<br>48,000[2] | 13.3<br>1.7<br>0.1 | 90.3<br>1.8<br>0.02 |
| **Industries** | | | | |
| Refinery | 184,918,000[3] | 9,703,000[4] | 5.2 | 3.5 |
| Iron and steel | 54,411,000[5] | 3,857,000[6] | 7.1 | 1.4 |
| Cement | 42,898,000[5] | 3,245,000[6] | 7.6 | 1.2 |
| Ammonia | 17,652,000[5] | 214,000[6] | 1.2 | 0.1 |
| Aluminum | 4,223,000[5] | 820,000[6] | 19.4 | 0.3 |
| Lime | 12,304,000[5] | 273,000[6] | 2.2 | 0.1 |
| Ethanol | 8,383,000[5] | 3,734,000[7] | 44.5 | 1.4 |
| Total | 2,564,489,000 | 276,106,000 | | 100 |

[1] U.S. Environmental Protection Agency (U.S. EPA) greenhouse gas inventory sector analysis.
[2] U.S. EPA acid rain and EGRID data (classified by primary fuel type).
[3] Estimate from U.S. DOE, 2002 BPD totals.
[4] Projected estimates from representative facilities.
[5] U.S. EPA greenhouse gas inventory industrial process analysis.
[6] Source data from U.S. Geological Survey, *Minerals Yearbook 2002*.
[7] Source data from Iowa Department of Agriculture and Land Stewardship: www.distillergrains.com.

Electric power generation plants emit about 2.24 Gt of $CO_2$ annually in the United States. The Illinois Basin contributes about 11.4% of the total U.S. $CO_2$ emissions from electric power generation plants. Coal is the dominant fossil fuel for electric power plants and contributes 83% of the total U.S. emissions and 98% of the Illinois Basin $CO_2$ emissions, respectively. $CO_2$ emissions from manufacturing industries in the Illinois Basin vary from industry to industry. For example, there is only one ammonia manufacturing plant, and its $CO_2$ emissions account for only 1.2% of the U.S. total in the industry. Ethanol production facilities in the Basin, however, contribute about 44.5% of total U.S. $CO_2$ emissions from that industry.

## Table 2.2. $CO_2$ Emission Sources in the Illinois Basin

| **Sources** | **Illinois** | | **Indiana** | | **Kentucky** | | **Total no. in basin** |
|---|---|---|---|---|---|---|---|
| | **$CO_2$ (total tonnes)** | **(no.)** | **$CO_2$ (total tonnes)** | **(no.)** | **$CO_2$ (total tonnes)** | **(no.)** | |
| **Power generation** | | | | | | | |
| Coal | 89,555,000 | 75 | 87,983,000 | 26 | 71,677,000 | 21 | 122 |
| Natural gas | 4,476,000 | | 321,000 | | 200,000 | | |
| Oil | 48,000 | | 0 | | 0 | | |
| **Industries** | | | | | | | |
| Refinery | 9,455,000 | 4 | 248,000 | 1 | 0 | 0 | 5 |
| Iron and steel | 3,685,000 | 17 | 142,000 | 5 | 30,000 | 1 | 23 |
| Cement | 1,301,000 | 4 | 1,353,000 | 3 | 591,000 | 1 | 8 |
| Ammonia | 214,000 | 1 | 0 | 0 | 0 | 0 | 1 |
| Aluminum | 0 | 0 | 464,000 | 1 | 356,000 | 1 | 2 |
| Lime | 273,000 | 1 | 0 | 0 | 0 | 0 | 1 |
| Ethanol | 3,665,000 | 6 | 0 | 0 | 69,000 | 2 | 8 |
| Total | 112,672,000 | 108 | 90,511,000 | 36 | 72,923,000 | 26 | 170 |

## 2.1. Power Generation Sector

Table 2.2 lists the total $CO_2$ emissions and the number of emission sources in Illinois, Indiana, and Kentucky that are in the geological boundary of the Basin. Coal-fired electric power plants are by far the largest point sources of emissions. About 254 million tonnes of $CO_2$ are emitted annually in the Illinois Basin from more than 122 coal-fired power plants. Figure 2.1 illustrates the relationship between $CO_2$ emissions and the number of power plants in the Illinois Basin. The figure is plotted according to the volume of $CO_2$ emissions from individual plants. The four largest power plants emit about 22% of the total $CO_2$ emissions, the 13 largest power plants emit more than 50% of total $CO_2$ emissions, and the 30 largest power plants emit over 80% of total $CO_2$ emissions. The largest coal-fired power plant in the Basin is the Rockport power plant in Spencer County, Indiana, which emits about 16 million tonnes of $CO_2$ annually. Considering the economy of scale, the larger power plants are the most suitable sources for any $CO_2$ capture and sequestration retrofits to existing power plants in the Basin. The geographical distribution of these power plants is shown in Figure 2.2.

Most of the power plants in the Illinois Basin are equipped with pulverized coal boilers and use a simple steam cycle. The flue gas from these power plants contains about 14% $CO_2$. Other contaminants in the flue gas, such as nitrogen ($NO_x$) and sulfur ($SO_x$), may have to be removed before the gas enters a $CO_2$ capturing system. Most of the power plants that burn high-sulfur bituminous coals are equipped with wet flue gas desulfurization (FGD) processes. These plants may have an advantage over the power plants without the FGD process (mostly burning western Powder River Basin (PRB) coal).

The power plants that burn natural gas tend to be small and are mostly peak power plants. Total $CO_2$ emissions from these power plants are about 5 million tons annually, which is about 2% of total emissions in the Illinois Basin.

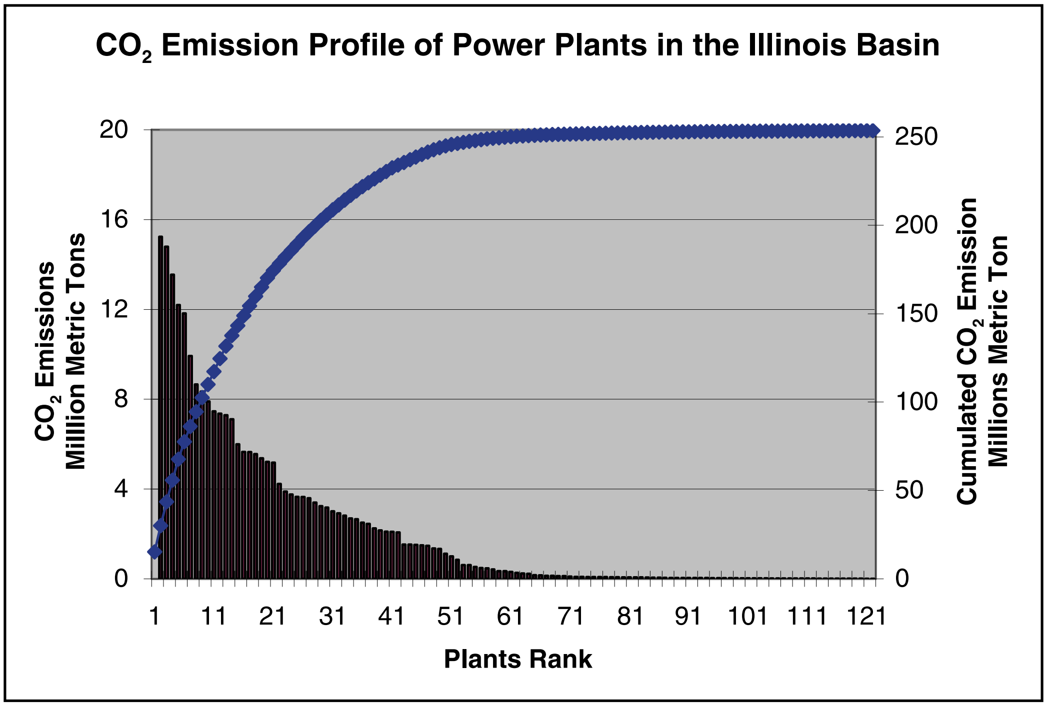


Figure 2.1. $CO_2$ emission profile of the power plants in Illinois Basin

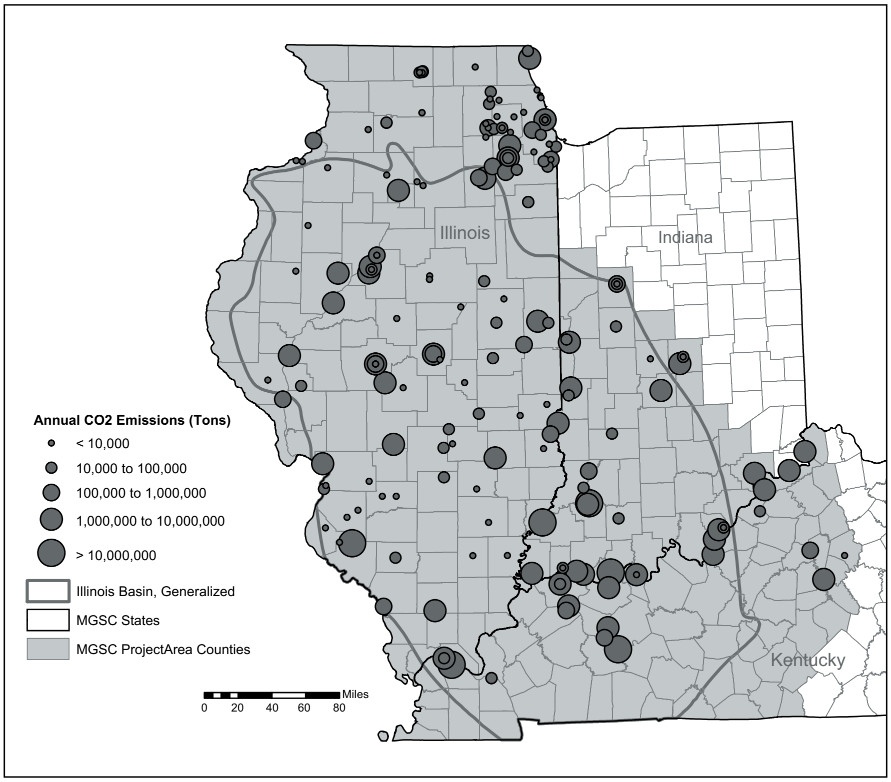


Figure 2.2 Geographical distributions of power plants in Illinois Basin

### 2.2. Industrial Sector

Combustion of fossil fuels is the main source of energy in energy-intensive industries such as petroleum refining, iron and steel manufacturing, and cement and lime production. The opportunity for carbon capture from these processes varies from industry to industry and even from plant to plant. The geographic distribution of these industrial sources in Illinois Basin is shown in Figure 2.3. A brief review of these industries is provided.

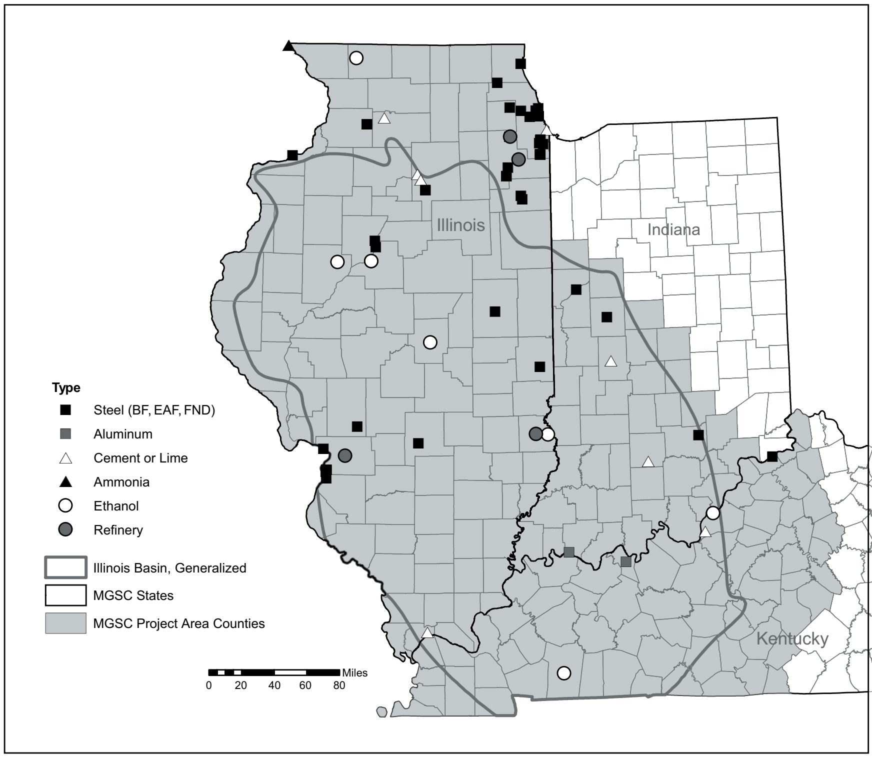


**Figure 2.3 Geographical distributions of industrial emission sources in the Illinois Basin**

#### *2.2.1. Petroleum Refineries*

Of the five petroleum refineries in the Illinois Basin, four are in Illinois, and one is in Indiana. Petroleum refining is an energy-intensive process, consuming about 5% of the total energy contained in crude oil (Phillips, 2002; Clarke, 2002). The refineries in the Illinois Basin emit about 9.7 million tons of $CO_2$ annually, the majority of which is from the combustion of the refinery off-gas and heavy components of petroleum oil, which provide process heat, and from the regeneration of the catalyst used in the fluid catalytic cracking process. About one quarter of $CO_2$ emissions from a refinery is attributed to fluid catalytic cracking regeneration (Clarke, 2002). The $CO_2$ concentrations in these flue gases are comparable to those of flue gases generated from burning coal in a PC-fired boiler. Another source, which is especially attractive for $CO_2$ sequestration, is refinery hydrogen plants. Some hydrogen plants produce a relatively pure (>95%) $CO_2$ stream and may provide the least expensive $CO_2$ source for sequestration, even though the quantity is relatively small.

A refinery is a complex of many processes, and flue gas ($CO_2$ emissions) stacks connected to various process heaters and boilers are scattered within the entire area of the plant. It might not be economical or practical to bring all of these scattered $CO_2$ emission sources to a centralized treatment site.

### *2.2.2. Iron and Steel Manufacturing*

There are 23 iron and steel plants in the Illinois Basin (17 are located in Illinois), and they emit about 3.86 million tons of $CO_2$ annually. Two major sources of $CO_2$ emissions from iron and steel production facilities are the blast furnace and the coke oven. About 70% of the total $CO_2$ emissions from an iron and steel production facility is from blast furnaces. Typically, blast furnace gas consists of a mixture of $CO_2$, CO, $N_2$ and $H_2$. The concentration of both $CO_2$ and CO in the process gas is about 20 vol%, and the pressure of blast furnace gas is about 1 to 3 bars. The furnace flue gas is usually burned to recover heat. The concentration of the $CO_2$ in the flue gas is about 27% (about double of that in the flue gas from a PC-fired power plant). The blast furnace gas contains about 20 vol% CO and may be shifted to $CO_2$ through a water-gas-shift reaction to increase the $CO_2$ concentration to about 40 vol% (Farla et al., 1995; Gielen, 2003).

### *2.2.3. Ammonia Manufacturing*

There is only one ammonia manufacturing plant in the Illinois Basin. It emits about 214,000 tonnes of $CO_2$ annually. In the ammonia synthesis process, pure hydrogen is produced. Hydrogen is generated through natural gas reforming or coal gasification followed by a water-gas-shift reaction. $CO_2$ can be separated from the shift gas through an absorption process producing a gas stream with >95 vol% $CO_{2.}$ Sometimes, the shifted gas is subject to a pressure swing adsorption (PSA) process to produce high-purity hydrogen gas. After the PSA process, the off-gas stream still contains significant amounts of CO and $H_2$. This off-gas usually is mixed with other fuel gases and combusted. If the gas mixture is burned in air, the $CO_2$ in the flue gas is diluted with nitrogen.

### *2.2.4. Cement Industry*

There are 8 cement plants in the Illinois Basin. They emit about 3.2 million tonnes of $CO_2$ annually. Cement production is an energy-intensive process (Van Oss and Padovani, 2002). There are two main sources of $CO_2$ emissions in a cement production facility. The first source is the decomposition (calcination) of the limestone that occurs in a kiln:

$$CaCO_3 + \text{heat} = CaO + CO_2.$$

CaO is one of the major (about 62%) components of cement. The second source of $CO_2$ is from the combustion of fossil fuel (coal, natural gas, or other fuels) to heat the kiln. The $CO_2$ emissions from calcination and from process heat are about 0.51 ton of $CO_2$ per ton of cement and about 0.43 ton of $CO_2$ per ton of cement, respectively. The two $CO_2$ streams are generally mixed together, which results in a gas stream that has a $CO_2$ concentration ranging from 13 to 33 vol%.

### *2.2.5. Lime Manufacturing*

The one lime manufacturing plant in the Illinois Basin emits about 273,000 tonnes of $CO_2$ annually. As with cement, lime is produced from limestone through a calcination process:

$$CaCO_3 + \text{process heat} = CaO + CO_2.$$

The process heat is generated by direct firing of coal or natural gas in the kiln. The $CO_2$ formed during combustion is mixed with the $CO_2$ released from the decomposition of limestone. The $CO_2$ released from limestone is relatively pure and thus increases the $CO_2$ concentration in the flue gas. The $CO_2$ concentration in a lime plant ranges from 13 to 33% depending on the type of kiln and the fuel used.

### *2.2.6. Ethanol Industry*

About 44.5% (3.7 million tons) of the total U.S. ethanol is produced in the Illinois Basin. There are 8 ethanol plants (6 in Illinois) with total $CO_2$ emissions of more than 50,000 tons annually. Because some of the $CO_2$ is recovered for various uses, the actual volume of emissions that is potentially available is less than reported by this industry.

Ethanol is mostly produced from corn through a fermentation process, and $CO_2$ is a by-product in this process. Each unit (mass) of ethanol produced generates an almost equal amount of $CO_2$. The $CO_2$ stream has a very high purity (>85%). Many ethanol plants recover and market the $CO_2$. The plants without $CO_2$ recovery are excellent sources for pilot $CO_2$ sequestration demonstration studies because the $CO_2$ is relatively easy to capture and inexpensive to dry (if needed to increase the $CO_2$ concentrations to >95 vol%). The cost of compression may be the greatest incremental cost to make these volumes of $CO_2$ available for research.

## 2.3. Concentration of $CO_2$ in Different Flue Gases

The $CO_2$ concentration in the flue gases is an important parameter in selecting cost-effective $CO_2$ separation and capture technologies. A higher $CO_2$ concentration usually translates into lower energy consumption and capture costs. Table 2.3 lists the typical $CO_2$ concentrations in different flue gas streams (Thambinuthu et al., 2002).

**Table 2.3. $CO_2$ concentrations in flue gases of different sources**

| Types of flue gases | Pressure (atm) | $CO_2$ concentration[1] (vol%) |
|---|---|---|
| Power station flue gas | | |
| PC combustion | ~1 | 14 |
| Natural gas combustion (simple steam cycle) | ~1 | 8 |
| NGCC | ~1 | 4 |
| Coal $O_2$/ $CO_2$ combustion | ~1 | >80 |
| Power station, pre-combustion capture of $CO_2$ | | |
| Coal gasification fuel gas | ~30 | 40 |
| Natural gas partial oxidation fuel gas | ~30 | 24 |
| Iron and steel blast furnace gas | | |
| Before combustion | ~2 | 20 |
| After combustion in air | ~1 | 27 |
| Cement kiln off-gas | ~1 | 13–33 |
| Lime manufacturing | ~1 | 13–33 |
| Oil refinery | | |
| Process heaters | ~1 | 8–15 |
| FCC regenerator | ~1 | 15 |
| Ethanol plants | ~1 | >85 |

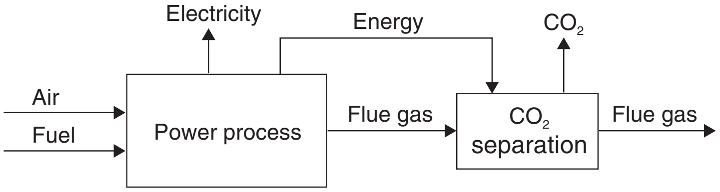
[1] The concentration of $CO_2$ varies from 4 vol% in the post-combustion flue gas of a natural gas combined cycle (NGCC) power plant to 95 vol% in the gas stream from a an ethanol production plant (after dehydration). The pressure of the most combustion flue gases is atmospheric pressure. Pre-combustion gas streams in an integrated gasification combined cycle (IGCC) and NGCC plants are at pressures higher than 25 atmospheres (atm). Under these conditions, the partial pressure of $CO_2$ is as high as 10 atm. FCC, fluid catalytic cracking.

## 3. $CO_2$ Removal Configurations

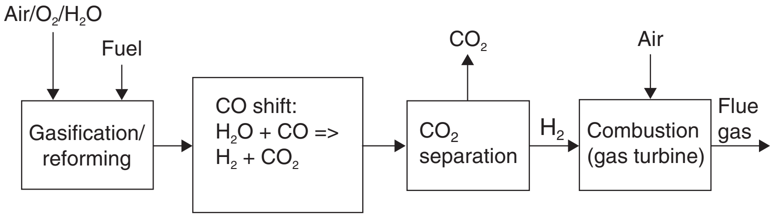
$CO_2$ separation and capture from flue gases of various stationary sources can be described by either post-combustion or pre-combustion configurations (Lyngfelt and Leckner, 1999; $CO_2$ Capture Project workshop, 2004). Figure 3.1 is a schematic of these configurations. Oxygen-enriched coal combustion can also be considered as a special case of the pre-combustion concept and is discussed later in this section. For the power generation sector, the selection of a capture concept mainly depends on the power generation process used. For most industrial emission sources, post-combustion capture is potentially the most suitable option.

### 3.1. Post-Combustion Capture

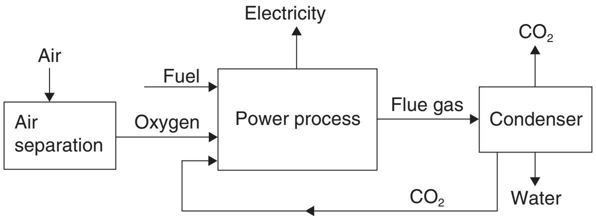
In the post-combustion configuration, $CO_2$ is captured from the flue gas after the fuel is combusted. When air is used as an oxidant, the combustion flue gas is diluted with the nitrogen in air. Thus, the $CO_2$ concentration in post-combustion flue gas is usually low and ranges from 4 vol% for a NGCC process to 14 vol% for a PC-fired power plant. The operating pressure of PC boilers is usually near atmospheric pressure. Due to the low $CO_2$ concentration of the flue gas and the low operating pressure, post-combustion capture tends to require large equipment sizes and, hence, higher capital costs. In addition, the low $CO_2$ partial pressure (concentration) in flue gas requires a stronger separation agent (solvent or adsorbent) to capture the $CO_2$, which, in turn, results in higher energy consumption. For most of the existing conventional coal-fired power plants and most industrial combustion processes, the post-combustion configuration is probably the only choice because it requires the minimum retrofit of the existing power plant.



a. Post-combustion capture



b. Pre-combustion capture



c. Oxygen combustion

**Figure 3.1. Different configurations for $CO_2$ separation processes**

Figure 3.1a is a simplified diagram of a power plant with post-combustion capture of $CO_2$. The configuration may also be used for IGCC and NGCC plants. However, because gasification processes are usually performed at high pressure (30 bars) and possess higher $CO_2$ partial pressure, a pre-combustion capture configuration operating at a high pressure is more attractive.

### 3.2. Pre-combustion Separation

If the $CO_2$ concentration and flue gas pressure could be increased, the size of $CO_2$ capture equipment would be significantly smaller, and weaker solvents, such as physical solvents, could be used. This would result in lowering the overall energy consumption of the separation process. Such a rationale is behind the pre-combustion configuration (Figure 3.1b) in which the original carbon-containing fuel is transformed to a non-carbon-containing fuel (usually hydrogen). Carbon in the fuel is converted to $CO_2$ prior to combustion and separated. Hydrogen is then used to produce power in a gas turbine or a fuel cell process.

Typical fuel reactions under a pre-combustion capture concept are fuel gasification to produce a synthesis gas (known as syngas), followed by water-gas-shift reaction,

$$\text{Coal} + H_2O + O_2 \rightarrow CO + H_2 \rightarrow CO_2 + H_2, \qquad (1)$$

or steam reforming of natural gas followed by a water-gas-shift reaction,

$$CH_4 + H_2O \rightarrow CO + H_2 \rightarrow CO_2 + H_2. \qquad (2)$$

$CO_2$ in the final product gas is separated from $H_2$ by different separation processes such as absorption processes, membrane processes, or adsorption processes. Pre-combustion separation processes operate at high pressure, which leads to higher $CO_2$ concentrations and smaller equipment size than are required in post-combustion $CO_2$ separation processes operating at near-atmospheric pressures.

The overall gasification process is, in principle, the same for coal, oil, or natural gas. But, when coal or oil is used as a fuel, the synthesis gas (syngas) has to be purified. A gas purification process is required to remove ash particles, sulfur compounds, and other minor impurities.

The pre-combustion separation process just described, without $CO_2$ capture, is already used in several commercial-scale IGCC plants fueled by coal, residual oil, and petroleum coke (SFA Pacific, Inc., 1999). One advantage of an IGCC power plant equipped with a $CO_2$ separation process is that it produces a pure hydrogen stream after the water-gas-shift reaction. The combination of the gasification process and the water-gas-shift reaction is the basis for the FutureGen Process (U.S. DOE, 2004).

### 3.3. Oxygen/Recycled $CO_2$ Combustion (Oxyfuel Combustion)

Conventional PC power plants use air as an oxidant. Air contains 80 vol% of $N_2$ and dilutes the $CO_2$ in the flue gas. If concentrated oxygen instead of air is used for combustion, either in a boiler or gas turbine, the concentration of $CO_2$ in flue gas can be increased greatly. High-purity oxygen can be produced by a large-scale, commercially available cryogenic process.

If fuel is combusted in very pure oxygen, the flame temperature will be excessively high, so a $CO_2$-rich flue gas is recycled to the boiler to reduce the flame temperature. The major advantage of the oxygen-blown coal combustion process is that the flue gas has a $CO_2$ concentration of 80 vol% or higher compared with 4 to 14 vol% for an air-blown coal combustion process. This greater $CO_2$ concentration results in a simpler post-combustion $CO_2$ purification process. The oxygen-blown combustion process has the further benefit of suppressing $NO_x$ formation. Since $N_2$ gas is not introduced to the boiler, the source for thermal $NO_x$ is eliminated, and only fuel $NO_x$ is present. This fuel NOx could be reduced or eliminated through staged combustion. In addition, for sequestration purposes, it also may be possible to eliminate the FGD process (required if the flue gas is treated by an amine-based solvent to capture $CO_2$) if concentrations of $SO_2$ in the 500 to 3,000 ppm range are proved to have minimal impact on the sequestration step.

The disadvantage of oxygen-blown coal combustion is that a large quantity of oxygen is required, which is expensive both in terms of capital cost and energy consumption. Advances in oxygen production processes, such as new and improved membranes that can operate at high temperatures, could improve the overall plant efficiency and economics of oxygen production facilities.

Oxygen-blown coal combustion could be an attractive option for retrofitting existing steam cycle power stations. The required modifications would be relatively minor and in some places supplies of oxygen could be obtained from existing commercial air separation plants. The technology aimed at power plants has so far only been demonstrated in small-scale test facilities (Varagani et al., 2004). Large-scale demonstration of the technology is needed to fully evaluate the technical and economic advantages of this process as a power generation technology and a source of high-purity sequestration-ready flue gas.

### 3.4. Comparison of Different $CO_2$ Configurations

A summary of the advantages and disadvantages of each of the $CO_2$ removal configurations is presented in Table 3.1.

**Table 3.1. Advantages and disadvantages of different configurations**

| Category | Advantages | Disadvantages | Improvement options |
|---|---|---|---|
| Post-combustion | • Can be used for existing systems<br>• High recovery rates<br>• High product purity<br>• Low pressure operation<br>• Mature technologies | • Low $CO_2$ partial pressure<br>• Energy intensive<br>• Large equipment size (due to low pressure)<br>• Acid gases need to be removed | • Better solvents<br>• Better gas-liquid contacting devise<br>• Further process optimization |
| Pre-combustion | • Possible step toward use of $H_2$ fuel<br>• Small equipment size<br>• Higher $CO_2$ concentration<br>• Less energy intensity<br>• Mature technologies | • High pressure operation<br>• Can not be used for PC plants<br>• High temperature operation | • $H_2$ gas turbine<br>• $H_2$ fuel cell<br>• $H_2$ separation membrane<br>• $H_2$ membrane reactor |
| Oxygen/ $CO_2$ recycle combustion | • Based on existing technologies<br>• High $CO_2$ concentrations (>90%)<br>• Can be used for existing systems | • Production of $O_2$ is expensive<br>• Acid gases needs to be removed<br>• Uncertain for use with NGCC | • $CO_2$ gas turbine<br>• More economic $O_2$ separation processes |

## 4. Technology Options for $CO_2$ Capture and Separation

A typical gas separation process has several components: feed, separating agent, separation device, and product stream (King, 1980) (Figure 4.1).

The degree of a separation process can be expressed in terms of a separation factor, $\alpha_{ij}$:

$$\alpha_{ij}^{S} = \frac{x_{i1}/x_{j1}}{x_{i2}/x_{j2}},$$

where $x$ is the mole composition in the streams; 1 and 2 refer to product and feed streams, respectively; and $i$ and $j$ refers to the components (e.g., $CO_2$ and $N_2$ in a binary gas mixture) of the stream. When $\alpha_{ij} = 1$, no separation is achieved.

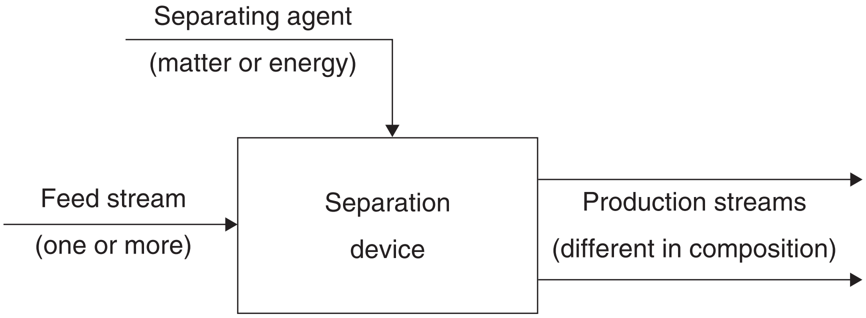


**Figure 4.1. Diagram of a typical separation process (King, 1980)**

Separation processes can be categorized as either equilibration processes or rate-governed processes. Most separation processes are equilibration, also called reversible separation, processes. They operate by equilibration of two immiscible phases that have different compositions at equilibrium. Evaporation, distillation (including cryogenic processes), absorption, and liquid extraction processes are examples of equilibration processes. Rate-governed processes are based on the differences in the transport rate of the gases (or liquids) through some medium. Membrane separation (membrane is the medium) and PSA (adsorbent is the medium) processes are examples of rate-governed processes.

Because the product phases of an equilibration process are immiscible, many equilibration stages can be arranged into one separation device (at the same pressure). For example, distillation processes can have many stages in one column. The advantage of the multi-staging is that the overall separation factor, which is the product of the individual separation factors in all stages, is large even though the separation factor in each stage is small ($\alpha_{ij}$ is close to 1). In the rate-governed processes, product phases are fully miscible with each other, and, as such, several rate-governed stages cannot be installed in one separation device. Due to this limitation, rate-governed processes often have one or only a few stages, and most membrane processes have only one or two stages. For effective separation, the stage separation factor of a rate-governed process has to be large because a small separation factor leads either to low product purity or low product recovery. Although a multi-staged, rate-governed process is possible, multiple separation devices (equipment) would be required, and the capital and operating costs of the separation process would increase, resulting in a costly and generally uneconomical process.

A separating agent (energy or matter) is used in a separation process to achieve a higher degree of separation. Depending on the agent, the process can be categorized as either an energy-separation agent or a mass-separation agent process. Energy is the most common separation agent. Distillation is an energy-separation agent process. Absorption and extraction processes are mass-separation agent processes that use solvents to separate the target components.

An energy-separation agent process usually requires fewer processing steps than does a mass-separation agent process because, in the latter, the separation agent usually has to be recovered through another separation process. For example, in an amine-based absorption process, the absorbed $CO_2$ is stripped out of the solution in a regeneration process to recover the solvent, which is then recycled to the absorption column. The addition of a mass-separation agent also causes corrosion and environmental risks in addition to increased process cost. In adsorption processes, solid adsorbents are used as mass-separation agents. Once the bed of the adsorbent materials reaches its saturation capacity, the bed has to be regenerated. Because a solid adsorbent is involved, adsorption processes often operate in batch mode (unless a fluidized bed is used). The batch-wise operation significantly reduces working capacity of the adsorption column. This disadvantage is critical if the adsorption capacity of the solid adsorbent is small or the concentration of sorbate ($CO_2$) in gas phase is high.

Some important factors to be considered in selecting a separation process are feasibility of the separation process, type of separation process, separation factor and equipment capacity, product value, and maturity. The first and most important criterion for selecting a separation technology is that the process has to be technically feasible. For example, an ion-exchange process is not be feasible for post-combustion $CO_2$ capture because all of the components in the coal combustion flue gas are non-ionic. Second is economic feasibility. A technically feasible separation process may not be economically feasible. Economic feasibility is determined by the cost of the separation process and the value of the products. The separation process (equilibration process vs. rate-governed process, energy-separation agent process vs. mass-separation agent process), separation factor, and equipment capacity of the separation process are all important factors in determining overall cost. The value of the product plays the most important role in the economics of a separation process. Finally, the maturity of the process has to be considered. A well-tested, mature separation process is often preferred to a less-tested, immature separation process, even if the latter process might be economically more favorable. Other specific requirements such as product purity and recovery rate also should be considered. For example, PSA and membrane separation processes tend to have low recovery if high purity is required, or have low purity if high recovery rate is required.

Two factors dominate the selection of a process for separating and capturing $CO_2$ from coal combustion flue gases. First, $CO_2$ is considered to be a relatively low-value product, even when it can be used for enhanced oil recovery (EOR) or coal-bed methane recovery. This factor makes many of the separation processes for this application uneconomical. Second, the scale of the $CO_2$ emissions from power plants is such that many of the commercially available separation technologies are not practical or economical for this application. For example, adsorption-based processes are unrealistic for $CO_2$ capture from flue gas because they require large quantities of adsorbents. Absorption, cryogenic, and membrane processes are potentially favorable for $CO_2$ capture from coal-power plant flue gases. Of these, absorption-based processes have many advantages over the other processes.

### 4.1. Absorption Process

Absorption is a mass-separation agent equilibration process. A chemical or physical solvent is used as the agent to capture the $CO_2$ in flue gas. Figure 4.2 is a schematic diagram of the absorption process using MEA as a solvent. A brief description of the process follows (Thambimuthu et al., 2002). In an absorption process, the flue gas enters into an absorption tower where it comes in contact with the chemical solvent. The solvent absorbs most of the $CO_2$ through a chemical reaction. The $CO_2$-rich solvent exits the bottom of the absorber and is passed into a stripper (desorption) column where it is heated with steam to release a concentrated $CO_2$. The $CO_2$ released in the stripper is recovered, and the $CO_2$-lean solvent is recycled to the absorption tower. Usually, $CO_2$ recovery rate (>98%) and $CO_2$ product purity (in excess of 99%) are high when a chemical solvent such as MEA is used.

Contaminants such as $SO_x$ and $NO_x$ that are present in flue gases from conventional coal or heavy oil-burning power plants have stronger affinities to react with chemical solvents, such as MEA, than $CO_2$ does. The presence of these contaminants will gradually reduce the absorption capacity of the chemical solvent. In order to avoid such detrimental effects, pretreatment of the flue gas may be needed.

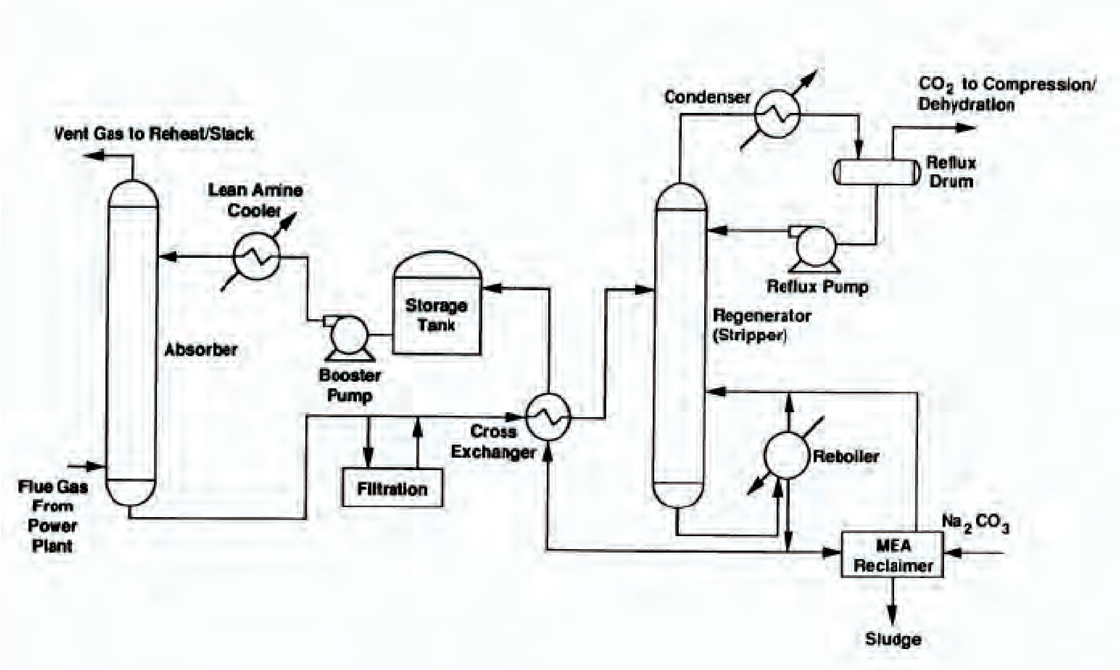


**Figure 4.2. Schematic diagram of the amine absorption process (Herzog, 1997)**

$CO_2$ absorption capacity (the solubility of $CO_2$ in the solvent) in a chemical or physical solvent depends on the properties of the solvent, the partial pressure of $CO_2$ in flue gas, and the operating temperature of the process. For chemical solvents, absorption is achieved through chemical interaction between $CO_2$ and solvent molecules. Chemical interaction is often selective, and chemical solvents tend to have limited absorption capacities. That is, absorption capacity levels off as the $CO_2$ partial pressure increases. For physical solvents, absorption is achieved through physical interaction (van der Waals forces) of $CO_2$ with the solvent molecules, and absorption capacity is approximately proportional to the partial pressure of $CO_2$ in the gas phase (Henry's Law).

Figure 4.3 illustrates the relationship between $CO_2$ capacity in a solvent and the $CO_2$ partial pressure in the gas phase (partial pressure of solute). At low $CO_2$ partial pressures, chemical solvents have higher absorption capacities than do physical solvents because they have stronger interactions with $CO_2$. At higher partial pressures, physical solvents have higher $CO_2$ absorption capacities.

Chemical absorption processes are recommended for post-combustion capture configuration because the $CO_2$ concentration in the flue gases of conventional power plants is about 14%, and the corresponding partial pressure is about 0.13 bar. The partial pressure is even lower in the flue gases of NGCC power plants. At these low partial pressures, physical solvents have little capacity for $CO_2$ capture due to their weaker intermolecular forces

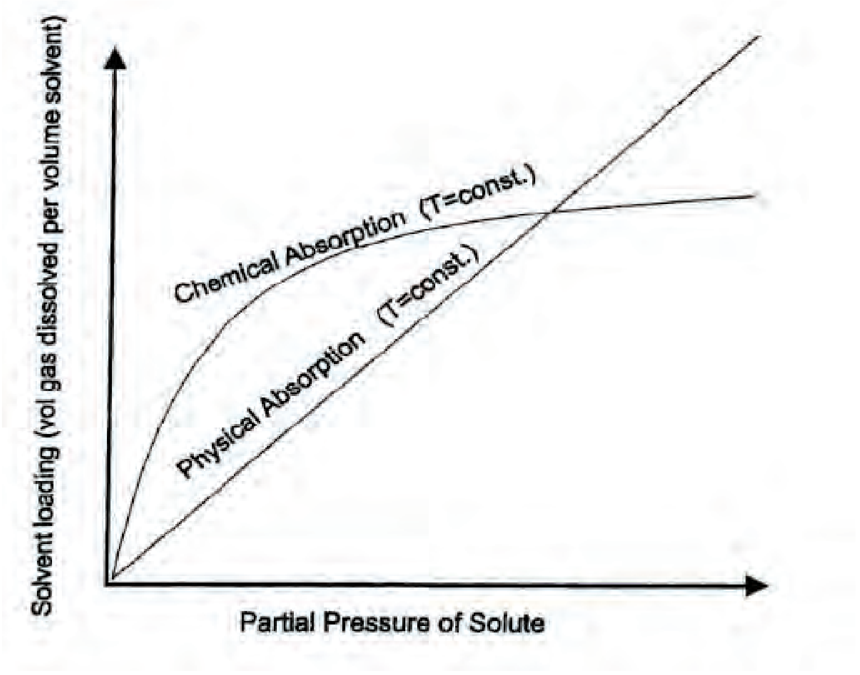


**Figure 4.3. Relationship between $CO_2$ capacity and its partial pressure (Gottlicher, 2004)**

Chemical absorption processes are well-understood unit operations in chemical engineering. Amine-based absorption processes to remove $CO_2$ from gas streams have been in practice for over 60 years in the chemical and oil industries, and absorption processes are expected to have few obstacles in scaling up from current commercial application processes to future $CO_2$ removal processes at power plants.

The main concerns with MEA and other amine solvents are corrosion in the presence of $O_2$ (DuPart, 1993a, 1993b) and other impurities, high solvent degradation rates from reaction with $SO_2$ and $NO_x$ (Strazisar et al., 2003; Stewart, 1994a, 1994b), and the large amounts of energy required for the regeneration step. These factors generally contribute to high energy consumption and large solvent losses (Chapel et al., 2001). However, if solvents with higher $CO_2$ absorption capacities, faster $CO_2$ absorption rates, lower degradation rates, lower corrosiveness, and lower energy requirements for regeneration than those of the currently available solvents could be developed, then the economics of chemical absorption processes could be greatly improved. Nevertheless, chemical absorption processes are the best technologies currently available for $CO_2$ separation from dilute, atmospheric pressure, post-combustion flue gas.

Physical absorption processes are more suitable for high pressure systems. For example, in an IGCC power plant, the $CO_2$ concentration in the syngas after the water-gas-shift step is about 40%. In addition, because the process operates at high pressures (around 30 bars), the $CO_2$ partial pressure is more than 10 bars, which is about 70 times higher than the $CO_2$ concentration in the flue gas from a conventional power plant. At this level of $CO_2$ partial pressure, the physical absorption process is preferable to chemical absorption. The weaker bonding between the $CO_2$ and the physical solvent allows the $CO_2$ to be separated from the solvent in a stripper by reducing the total pressure. Such a mechanism results in low energy consumption. The most common physical solvents that are commercially used for $CO_2$ capture include cold methanol (Rectisol process), dimethylether of polyethylene glycol (Selexol process), propylene carbonate (Fluor process), and sulfolane (Gupta et al., 2003; Kohl and Nielsen, 1997).

Physical absorption processes are also well established in ammonia and hydrogen production plants. The developmental improvements for the physical absorption processes are similar to those for the chemical absorption processes, in particular, the need for higher efficiency gas-liquid contactors and novel solvents with lower energy requirements for regeneration.

### 4.2. Adsorption Process

Adsorption is a mass-separation agent process. A porous solid adsorbent, such as zeolites or activated carbon, is used as a mass-separation agent. Gases or vapors can be captured by contact with an adsorbent through chemical or physical interaction. Adsorption can be either a rate-governed process or an equilibration process. Adsorption is a surface phenomenon. The adsorption capacity depends on the properties of the adsorbent and adsorbate (such as $CO_2$ or $H_2$) and the partial pressure of the adsorbate. Adsorption capacity is often limited by surface area, pore size, pore size distribution, and porosity of the adsorbent. Adsorption capacity tends to be low compared with a bulk process such as absorption process. In addition, because a solid phase is involved, the continuous operation of an adsorption process is more difficult. In fact, most adsorption processes operate on a repeated cycle between adsorption and desorption (regeneration) steps. Due to these drawbacks, adsorption processes are often used to separate low-concentration components from a mixture.

In an adsorption-based process for $CO_2$ capture, flue gas is fed into an adsorption column where $CO_2$ is selectively adsorbed on the surface of the adsorbent. The bed of $CO_2$-rich adsorbent is then subjected to a regeneration step to recover $CO_2$. Depending on the regeneration step used, the adsorption process can be PSA, temperature swing adsorption (TSA), or electric swing adsorption (ESA). Figures 4.4 and 4.5 are schematic diagrams of the PSA and TSA processes.

TSA processes are not generally suitable for capturing $CO_2$ from flue gases of coal-fired power plants, mainly because of the huge amount of $CO_2$ to be recovered and the relative low capacity of adsorbents (Riemer, 1993; Gottlicher, 2004).

PSA processes, in contrast, use rapid cycles of pressure change, which can partially compensate for the low capacity of the adsorption process. However, PSA processes tend to have either low product recovery rate or low product purity. It is very difficult to achieve both high recovery and high purity at the same time. PSA is not an attractive post-combustion capture process for $CO_2$ removal from flue gases; however, it may offer applications for co-production of power and $H_2$ in an IGCC plant where $H_2$ separation from syngas requires only high purity $H_2$, but not a high recovery rate.

The equipment capacity of an adsorption process is comparable to or less than that of an absorption-based process. However, because an adsorption process is performed in repeated cycles and the equipment capacity is not fully used all the time, the working (actual) capacity is much lower than that of absorption processes. As a result, the size of adsorption equipment is larger than the size of absorption equipment for the same gas separation application.

The process equipment for the adsorption process is similar to that for the absorption process. The unit equipment cost of adsorption process is also similar to that for the absorption processes. The total equipment cost for adsorption-based $CO_2$ capture, however, will be higher than that for absorption-based $CO_2$ capture processes.

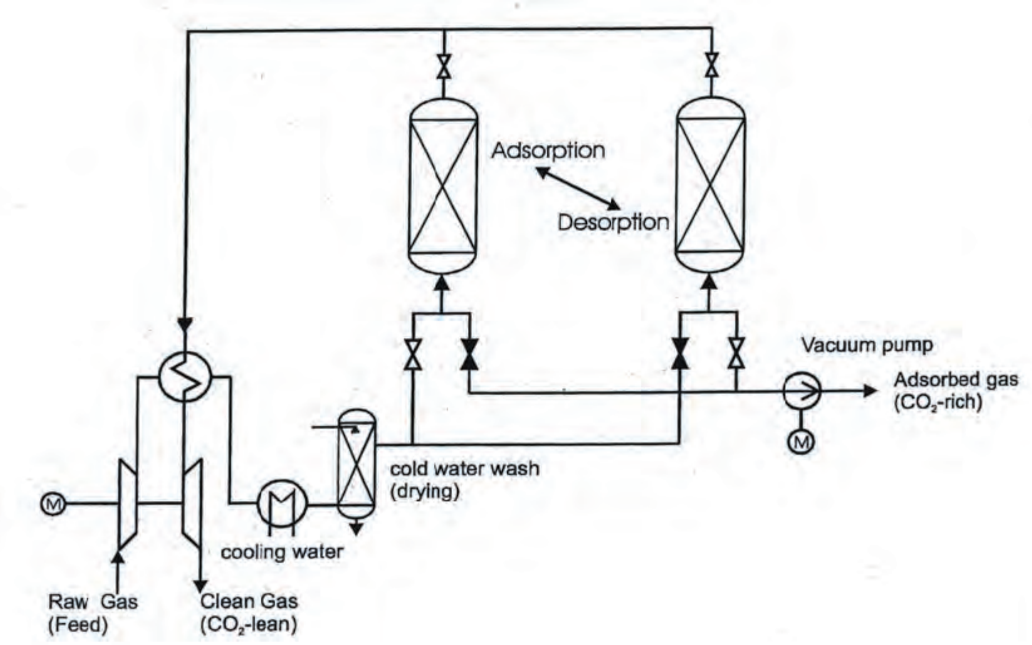


Figure 4.4. Schematic of a pressure swing adsorption process (Gottlicher, 2004)

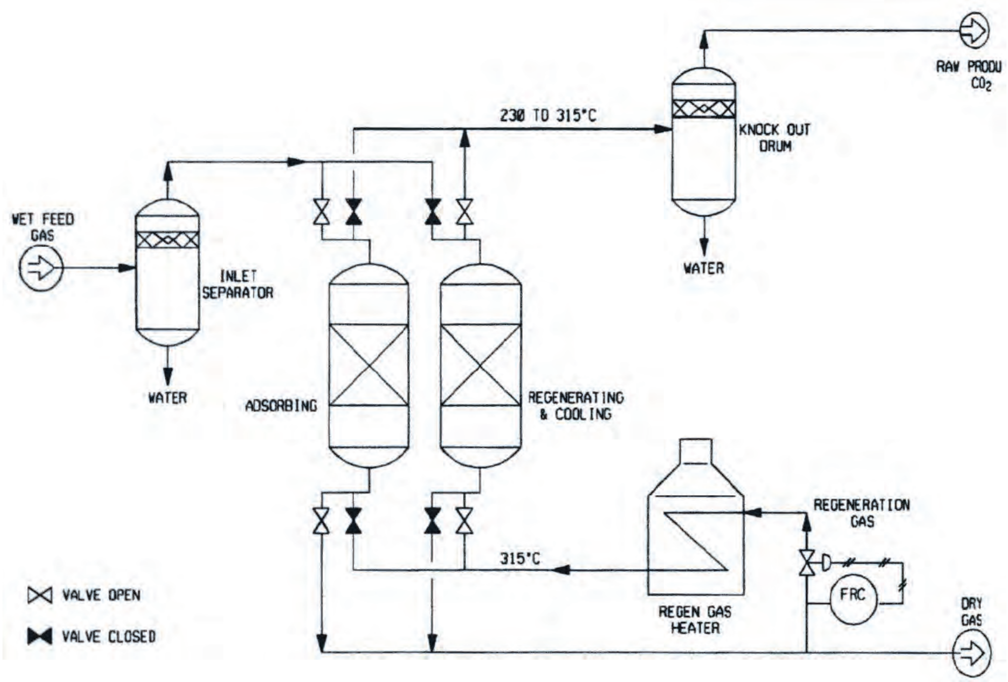


**Figure 4.5. Schematic of a temperature swing adsorption process (Riemer, 1993)**

### 4.3. Membrane Processes

Membrane separation is an energy-separation agent and rate-governed process. This process is based on the differences in physical or chemical interactions between gases and a membrane material, thereby causing one component to pass through the membrane faster than others. Various types of membranes are currently available, including porous inorganic membranes, palladium membranes, polymeric membranes, and zeolite membranes. Figure 4.6 is a schematic diagram of one-stage (left) and two-stage (right) membrane processes.

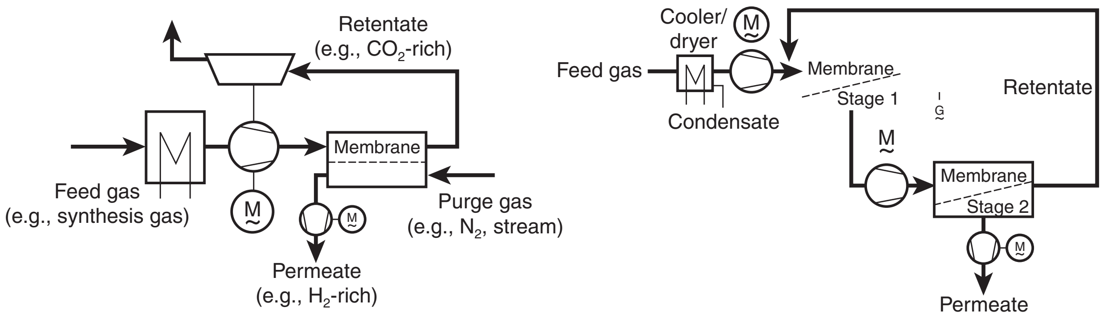


**Figure 4.6. Schematic diagram of one-stage (left) and two-stage (right) membrane processes (Gottlicher, 2004)**

Selectivity and permeability of the membrane are two of the important factors in determining the economics of a given membrane process. High selectivity is required to produce a high-purity product gas stream. To achieve high purity at an acceptable recovery rate, multiple stages and/or recycle of one of the streams may be necessary. This option, however, significantly increases the capital cost and energy consumption. Permeability determines the capacity of a membrane system. Low permeability translates to a larger membrane area and higher equipment cost. One study (Van Der Sluijs et al., 1995) concluded that for a post-combustion membrane separation process to compete economically with an amine-based absorption process, the $CO_2/N_2$ selectivity and $CO_2$ permeability of the polymeric membrane should be >200 and 100 Barrer, respectively (see Figure 4.7). The selectivity and permeability of currently available commercial membranes (the data points in Figure 4.7) are far smaller than the required target values (shaded area). Membrane processes are not an attractive option for removal of $CO_2$ from post-combustion flue gas at the present time. The future application of this technology depends mainly on the development of improved membranes.

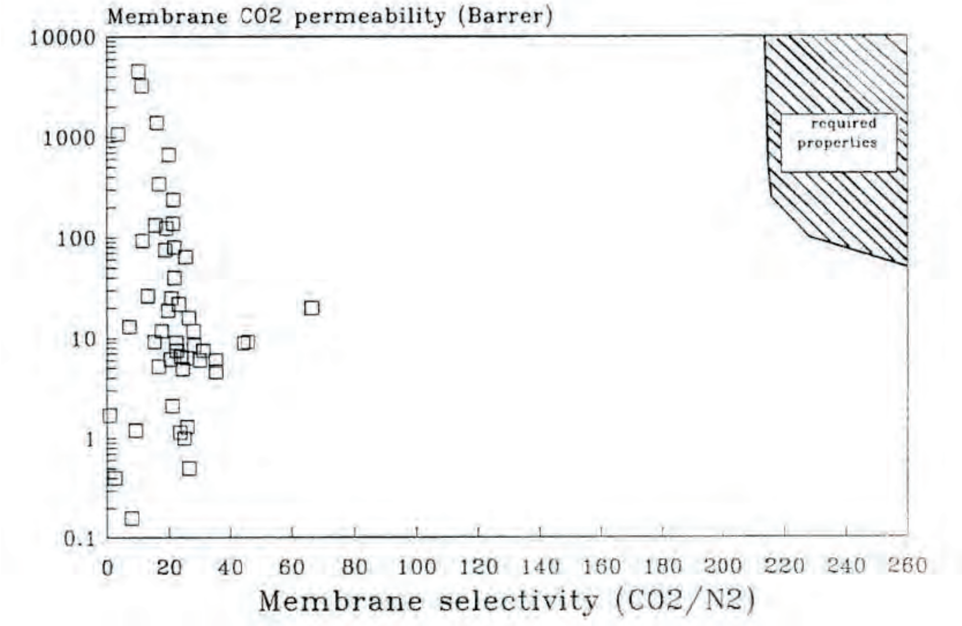


**Figure 4.7. Commercial membrane and required membrane properties (Van Der Sluijs et al., 1995)**

Membrane processes could be used to separate $CO_2$ from the shift syngas in an IGCC power plant (membrane shift reactor). In this case, gas cooling and heating would not be required, and the equipment cost for $CO_2$ separation could be significantly reduced. Indeed, conceptual level techno-economic studies have shown that the IGCC + membrane reactor is a very attractive pre-combustion $CO_2$ removal process (Middleton, 2004). The unit equipment cost for a membrane separation system generally is much higher than that of absorption and adsorption processes.

## 4.4. Cryogenics

A cryogenic process, also called low-temperature distillation, is an energy-separation agent and an equilibration separation process (Riemer, 1993) (Figure 4.8). This process relies on component volatility conditions in the mixture at cryogenic conditions. $CO_2$ can be separated from flue gases by cooling and condensation.

To analyze the suitability for $CO_2$ separation by cryogenic processes, it is important to study the $CO_2$ phase diagram (Figure 4.9). Because distillation processes can operate only in the vapor-liquid region of the phase diagram, the flue gas temperature must be cooled to between –56.6°C to 31°C. The vapor pressures of $CO_2$ at these two temperatures are 5 and 74 atm, respectively, which is much higher than the partial pressure of $CO_2$ in most flue gases. To employ a distillation process, the flue gas must be pressurized. For example, in conventional PC power plants with a $CO_2$ concentration of 14% (or 16% dry basis) in the flue gas, the flue gas has to be compressed to a total pressure of at least 30 atm to recover any $CO_2$ at –56°C. If 90% recovery is required, total pressure needs to be 300 atm. Obviously, to compress and cool the flue gas to such high pressure and low temperature is extremely energy intensive. Clearly, cryogenic processes are not an attractive option for low-concentration $CO_2$ streams.

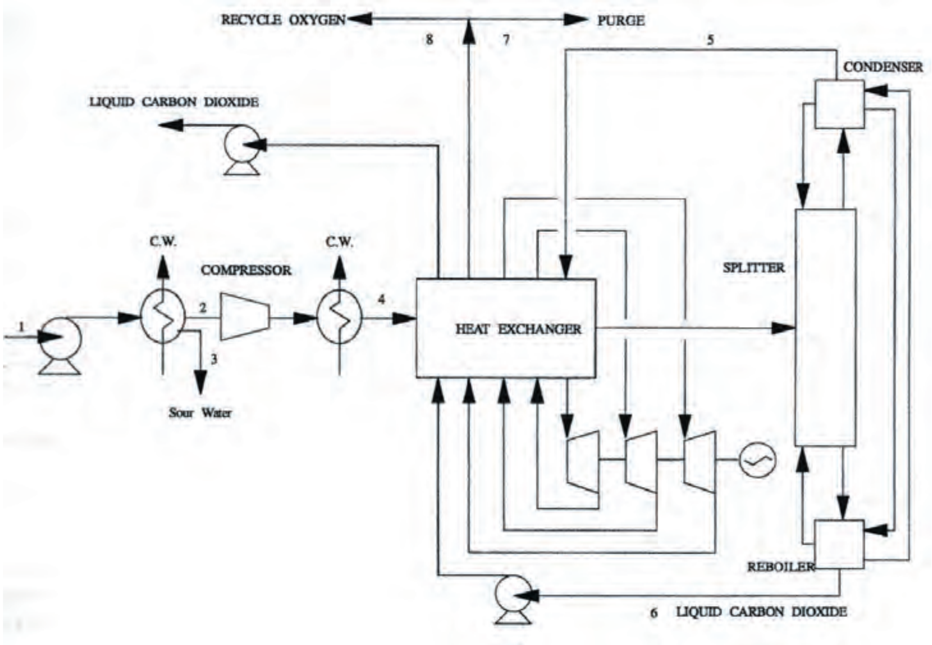


**Figure 4.8. Schematic diagram of a cryogenic process (Riemer, 1993)**

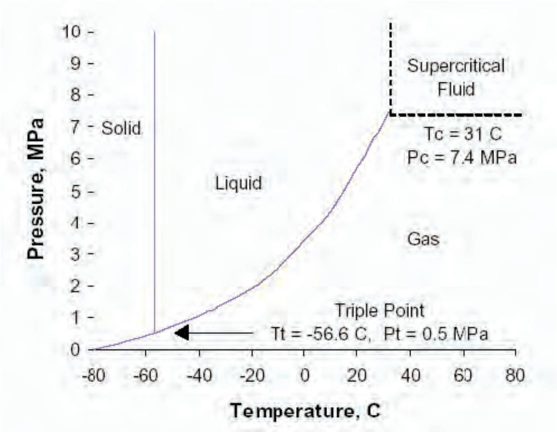


**Figure 4.9. Phase diagram of $CO_2$ (Tc and Pc, critical temperature and pressure; Tt and Pt, triple point temperature and pressure.**

A cryogenic process is more attractive for IGCC plants because $CO_2$ concentration in the shifted syngas is about 40% and the total partial pressure of $CO_2$ is about 12 bars. A system pressure of about 120 bars is required for 90% recovery of $CO_2$ from the shifted syngas stream. Cryogenic processes are even more favorable for an oxyfuel combustion process in which the flue gas stream contains >90% of $CO_2$ (dry basis). A 90% recovery of $CO_2$ requires an operating pressure of 50 bars.

Cryogenic processes are commercially employed for the purification of $CO_2$ from streams that contain high $CO_2$ concentrations (typically >90%). A major disadvantage of cryogenic separation of $CO_2$ is the amount of energy required to provide the necessary refrigeration, particularly for dilute gas streams. Another disadvantage is that some of the components in the gas, such as water, have to be removed before the gas stream is cooled to avoid freezing and blockage in the heat exchangers. The most promising applications of the cryogenic process are separation of $CO_2$ from high-pressure gases, such as in pre-combustion capture processes or in oxyfuel combustion in which the input gas contains a high concentration of $CO_2$.

The equipment capacity of a cryogenic process is comparable to that of a distillation process and is similar to that of absorption processes. The unit equipment cost is similar to that of absorption processes. Thus, the capital cost of this process would be similar to the absorption process and lower than that of the membrane and adsorption processes.

## 4.5. Emerging Power Generation and $CO_2$ Capture Concepts

Zero emissions coal technology and chemical looping process $CO_2$ have been proposed as new concepts for power generation, and the hydrate separation process is considered as a potential $CO_2$ capture method for IGCC power plants. Each of the emerging power generation concepts has incorporated a $CO_2$ capture step in the overall process scheme. Each of these concepts is described briefly.

### *4.5.1. Zero Emissions Coal Technology*

Zero emissions coal technology (Ziock et al., 2001) has four components: hydrogasifier, CaO reformer, calciner, and fuel cell (Figure 4.10). The reactions involved in each stage are

Hydrogasifier:

$$C + 2H_2 \rightarrow CH_4, H_2O\ (\text{liquid})\ H_2O\ (\text{gas})$$

CaO reformer:

$$CH_4 + 2H_2O \rightarrow CO_2 + 4H_2O, CaO + CO_2 \rightarrow CaCO_3$$

Calciner:

$$CaCO_3 \rightarrow CaO + CO_2$$

Fuel cell:

$$2H_2 + O_2 \rightarrow 2H_2O$$

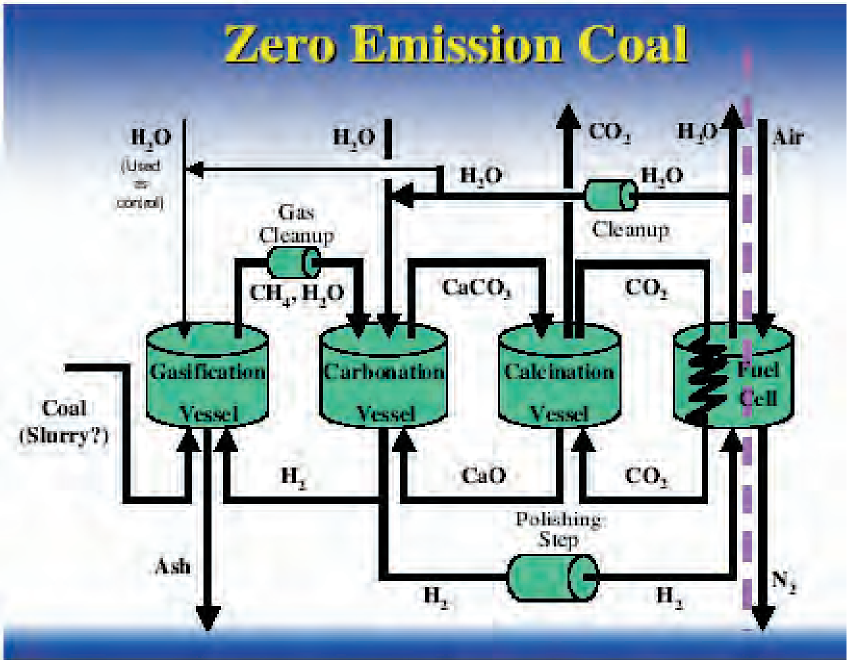


**Figure 4.10. Schematic diagram of zero emission coal technology**

In the hydrogasification stage, both hydrogen and water are used to gasify coal to produce a methane-rich fuel gas, which is reformed to hydrogen and $CO_2$. $CO_2$ is removed by CaO to form $CaCO_3$. High-purity $CO_2$ is recovered during the decomposition of the $CaCO_3$ in a calciner. The $CO_2$ recovery is a pre-combustion stage. The process is attractive when it is coupled with a solid oxide fuel cell process. The waste heat from the exhaust gas of a solid oxide fuel cell process can be used to decompose $CaCO_3$.

### *4.5.2. $CO_2$ Hydrate Separation Process for Synthesis Gas*

The $CO_2$ hydrate separation process is used to separate $CO_2$ and $H_2$ in a shifted syngas (Tam et al., 2001). The hydrate is a mixture of water and a gas. The gas molecules are trapped in the cavity of a crystal water (ice) structure. Many small gas molecules can form gas hydrate at certain temperatures and pressures. $CO_2$ hydrate forms at temperatures near $0^{o}$C and pressures from 10 to 70 atm, depending on the gas composition. The $CO_2$ hydrate separation process consists of two stages (Figure 4.11). In the first stage, nucleated water, also called seeds, are formed by an ammonia cooling process. The nucleated water and shifted syngas enters the second reactor at pressures ranging from 6 to 20 bars. $CO_2$ in the gas phase reacts rapidly with nucleated water to form $CO_2$ hydrates. $CO_2$ hydrates are separated from the gas to produce an $H_2$-rich product gas.

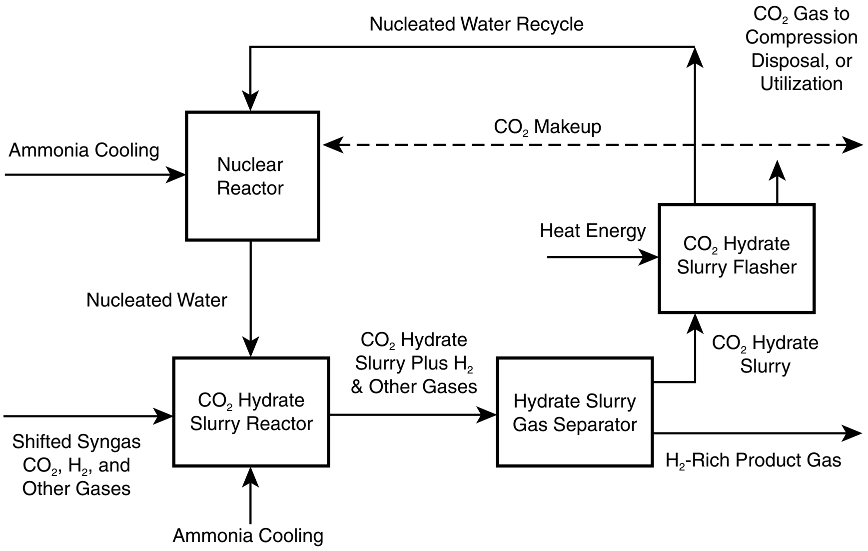


**Figure 4.11. Simple block diagram of a $CO_2$ hydrate process (Tam et al., 2001)**

The potential technical barriers of $CO_2$ hydrate separation include efficient release of $CO_2$ from the hydrate; efficient capture of $CO_2$ (increased recovery rate); formation of stable pre-hydrate; and the interference of trace contaminants with hydrate formation. The hydration process is suitable for IGCC processes where the gas streams are at pressures of 20 atm or above and the $CO_2$ partial pressure is at 8 atmospheres or above. These conditions ideally meet $CO_2$ hydrate-forming requirements.

### *4.5.3. Chemical Looping Process*

The chemical looping combustion processes (Figure 4.12) use a metal oxide (or other oxygen carriers) to transfer oxygen to the fuel in a two-stage process (Mattisson and Lyngfelt, 2001). In the reduction reactor, the metal oxide oxidizes the fuel, and the metal oxide itself is reduced. The metal oxide is transported to an oxidation reactor where the reduced metal oxide is re-oxidized in air. The oxidation and reduction processes are repeated.

The major advantage of the chemical looping process is that no oxygen separation process is required, which eliminates the air separation unit and significantly reduces capital cost. The major disadvantage is that the process cannot be directly used with coal because coal is a solid fuel and is not able to efficiently contact the metal oxide, which is also a solid.

Other new power generation concepts and $CO_2$ capture technologies have also been proposed, but all of them are still at conceptual stages (White et al., 2003; Gupta et.al., 2003).

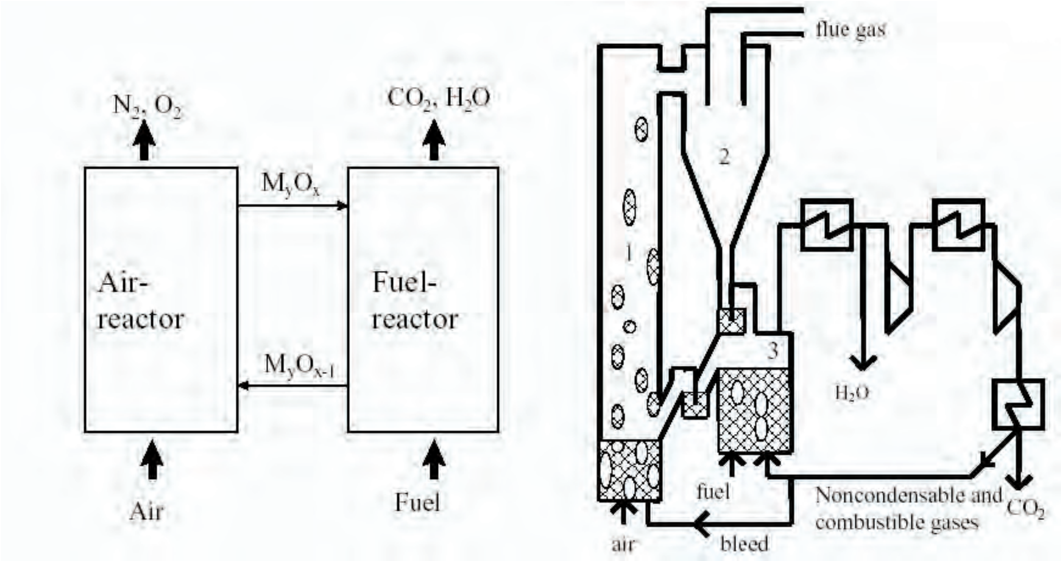


Figure 4.12. Two proposed processes for the chemical looping technology

## 5. Comparison of Different $CO_2$ Capture Technologies

The $CO_2$ capture technologies discussed in the previous section are potential candidates for $CO_2$ capture from large-scale power plants and industrial facilities. Some of the advantages and disadvantages of these technologies have been discussed in previous sections. In this section, a more complete comparison is presented.

When comparing different separation technologies, the ultimate criteria are cost and safety. The most important factors include equipment size, unit equipment cost, energy consumption of the process, operating cost, reliability of the process, product recovery rate, product purity, flexibility of the process, and other parameters. Table 5.1 lists these factors for selected separation processes.

The excergy (useful energy, not heat) efficiencies of these processes are also listed in Table 5.1 (Gottlicher, 2004). Since the minimum (theoretical) energy requirements are the same for all capture technologies listed, excergy efficiency is thus an indication of the actual energy requirement of each process:

Excergy efficiency = theoretical energy requirement/actual energy consumed.

Energy consumption often dominates the operating cost of a separation process. From the data in Table 5.1, it can be seen that, when $CO_2$ partial pressure in the gas mixture is low, chemical absorption process generally has a higher excergy efficiency. When $CO_2$ partial pressure is high (e.g., ~10 bars), the physical absorption process has higher exergy efficiency. Membrane processes have high excergy efficiency when they are used for $H_2$ separation rather than for $CO_2$ separation. The excergy efficiency of a membrane process is strongly correlated to the selectivity of the membrane used and, to a lesser extent, to its permeability (Van Der Sluijs et al., 1992; Hendriks, 1994; Gottlicher, 2004). Due to the high selectivity and permeability of $H_2/CO_2$ membrane, the membrane system can achieve higher exergy efficiency.

### Table 5.1. Comparisons of the potential separation technologies

| | Absorption | PSA/TSA | Membrane | Cryogenic |
|---|---|---|---|---|
| Excergy efficiency (%) | | | | N/A |
| Post-combustion | | | | |
| PC (P = 1.2 bars; $CO_2$ = 14%) | 21 (CA) | 9 | 10~14 | |
| IGCC (P = 1.2 bars; $CO_2$ = 8%) | 14 (CA) | 5 | 6.5~8 | |
| NGCC (P = 1.2 bars; $CO_2$ = 4%) | 13 (CA) | 1.5 | 1.5~2 | |
| Pre-combustion | | | | |
| IGCC (P = 24 bars; $CO_2$ = 40%) | 20~32 (PA) | 8~12 | 25~32 ($H_2$) | |
| NGCC (P = 24 bars; $CO_2$ = 36%) | | 8~12 | 25~32 ($H_2$) | |
| Ease of scaling-up | good | poor | good | good |
| Economy of scale | good | fair | poor | good |
| $CO_2$ recovery rate | good | poor | poor | good |
| $CO_2$ purity | good | fair | fair | good |
| $H_2$ recovery (for IGCC after shift) | good | poor | fair | good |
| $H_2$ purity (for IGCC after shift) | good | good | good | good |
| Equipment size | fair | poor | good | fair |
| Unit equipment cost | good | fair | poor | good |
| Operating cost | fair | fair | good | fair |
| Maintenance | fair | fair | good | fair |
| Maturity | good | fair | fair | good |
| Pre-treatment requirement | fair | fair | good | good |

It is difficult for a membrane process or a PSA process to achieve high $CO_2$ recovery percentage in a one- or two-stage process (unless the selectivity of the membrane or sorbent is extremely high), especially when high product purity is required. For these processes, there is often a trade-off between product purity and recovery rate. When high recovery is not required, PSA and membrane systems may have potential.

The relationship between oxygen cost and different scales of membrane, cryogenic, and adsorption processes is shown in Figure 5.1 to illustrate some conclusions for large scale $CO_2$ separation and capture. Membrane separation processes do not have good economy of scale. It is expected that the scale-up behavior of an absorption process will be similar to that of a cryogenic process because the process equipment of these two processes is similar. Obviously, large-scale processes such as $CO_2$ capture from utility flue gas definitely favor absorption processes.

The study cited in Figure 5.1 is about 10 years old. However, the relative cost-capacity trends shown are still valid because they are mainly influenced by the nature of the separation processes.

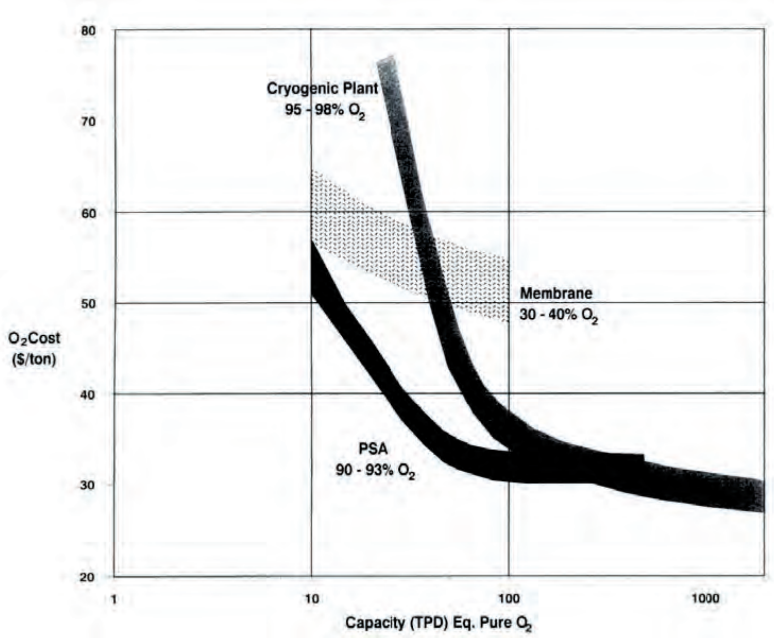


**Figure 5.1. Scale-up behavior of different technologies for $O_2$ separation (Prasad et al., 1994)**

## 6. Selection of $CO_2$ Capture Technologies for Different Emission Sources

The selection of a $CO_2$ capture technology mostly depends on the characteristics of the emission source and the subsequent disposal of the captured $CO_2$. The most important parameters are the $CO_2$ concentration and the total pressure in the flue gas stream. Others include contaminants in the gas stream, transportation, and disposal methods. Table 6.1 lists the characteristics of different $CO_2$ emission sources and recommended capture technologies.

For a post-combustion flue gas, $CO_2$ partial pressure (both concentration and total pressure) is low, and a chemical absorption process is the best choice for the present and the near future. Development of new solvents and gas/liquid contactors is expected to improve the economics of absorption-based processes for $CO_2$ capture in the future.

Another possible alternative for conventional PC power plants is to retrofit them with the oxyfuel combustion process. This technology may offer a low-cost option for $CO_2$ capture if combustion with recycled $CO_2$ is demonstrated to be a viable low-cost technology for the existing PC boilers.

**Table 6.1. Capture technologies for power plants and industrial facilities**

| Emission source | Pressure (bar) | $CO_2$ (%) | Impurities | Capture technology |
|---|---|---|---|---|
| Power plants | | | | |
| PC | 1.2 | 14 | $SO_x$, $NO_x$ | CA |
| IGCC post-combustion | 1.2 | 8 | $NO_x$ | CA |
| IGCC shifted syngas | 30 | 40 | $H_2S$ | PA or $H_2$ membrane |
| NGCC post-combustion | 1.2 | 4 | $NO_x$ | CA |
| NGCC shifted syngas | | | | PA or $H_2$ membrane |
| PC + $O_2/CO_2$ | 1.2 | >90 | $SO_x$, $NO_x$, $H_2O$ | Cryogenic |
| Industrial processes | | | | |
| Iron and steel | 1.2 | 20–27 | | CA or shift + PA |
| Refineries | | 8–15 | $SO_x$, $NO_x$ | CA |
| Cement | 1.2 | 14–33 | | CA |
| Lime | | 14–33 | | CA |
| Ammonia | 30 | >95 | | Pure |
| Ethanol | 1.0 | 95 | VOCs, $H_2O$ | Cryogenic |

For pre-combustion capture, such as in IGCC + shift or NGCC + shift power plants, the best technology would be to shift CO in syngas to $CO_2$ followed by a physical absorption process to separate $CO_2$. Another option is to separate $H_2$ from the shifted syngas instead of separating $CO_2$ from the syngas. $H_2$ can be separated by using a membrane separation process or a membrane reactor, which integrates water-gas-shift reaction and $H_2$ separation in the same unit. Because inorganic membranes can operate at high temperatures, this option may increase the thermal efficiency of the power plant. The membrane reactor may represent the best future opportunities for $CO_2$ capture in IGCC and NGCC processes. The key to the success of the membrane reactor process is the development of high $H_2$ selectivity and permeability separation membranes.

For industrial processes with a low $CO_2$ concentration in the flue gas, chemical absorption processes offer the best option at the present time.

## 7. Selected Scenarios for Illinois Basin

Currently, the predominant power generation technology is simple steam cycle PC power plants. As discussed in the previous sections, the chemical absorption process is the most suitable $CO_2$ capture technology for these power plants. Chemical absorption is also the best option for most of selected industrial processes. The oxyfuel combustion process could be a near term alternative for PC power plants. Because the oxyfuel combustion process does not increase the thermal efficiency of the power plants, such technology will not be attractive if $CO_2$ removal becomes mandatory. This technology will likely be phased out with PC boilers.

IGCC technology is potentially the next coal-based power generation technology to replace PC technology. IGCC offers higher thermal efficiency and, thus, fewer $CO_2$ emissions, and is a bridge to a hydrogen-based economy. For the distant future, higher thermal efficiency power generation technologies, such as solid oxide fuel cell and zero emissions coal technologies, have good potential.

**Table 7.1. Likely combinations of $CO_2$ separation and power generation technologies**

| Current technologies | Near term | Future |
|---|---|---|
| PC + chemical absorption<br>PC + oxyfuel process | IGCC + shift + Selexol<br>IGCC + shift + Membrane reactor<br>NG reforming + shift + Selexol<br>NG reforming + shift + membrane reactor | Coal gasification + shift + Selexol + SOFC[1]<br>NG reforming + shift + Selexol + SOFC<br>ZEC technology<br>Innovative technologies |

[1] SOFC, solid oxide fuel cell; NG, natural gas; and ZEC, zero emissions coal.

Table 7.1 lists the likely combination of $CO_2$ capture technologies and power generation technologies in the near and more distant future. From this review, the following three processes are selected for techno-economic analysis: PC + chemical absorption (MEA), IGCC + shift + absorption (Selexol), and oxyfuel combustion processes.

In addition, a techno-economic analysis is performed for capturing $CO_2$ from ethanol plants. Ethanol plants may be a good $CO_2$ source for phase II of the DOE sequestration program in the Illinois Basin, which involves actual field demonstration of $CO_2$ sequestration at a selected site.

## 8. Techno-economic Analysis of $CO_2$ Capture in the Illinois Basin

Existing coal-fired power plants will be in operation for some time in the future and, therefore, their $CO_2$ emissions are potential targets for sequestration. As stated earlier, the initial results from the analysis of various $CO_2$ capture processes indicate that a post-combustion chemical (MEA) absorption process would be the most suitable option for capturing $CO_2$ from the flue gases of the existing coal-fired power plants. IGCC is expected to be a promising future coal-based power generation technology, especially if hydrogen becomes a more attractive fuel for the transportation sector. A pre-combustion physical absorption process (Selexol) is thought to be the most promising $CO_2$ capture process for this technology at the present time. A promising retrofit and/or new technology for generating high-purity $CO_2$ coal combustion flue gas is oxygen-enriched combustion (OEC) with $CO_2/O_2$ flue gas recycle. A techno-economic study was performed to evaluate the relative performance and cost of these three power generation systems (i.e., PC + MEA, IGCC + Selexol, and OEC). The study determined the cost of electricity generation and the $CO_2$ avoidance costs. A conventional PC power plant without $CO_2$ capture was also investigated to provide a baseline for comparison.

The scale of a power plant, especially when the plant scale is small, can strongly influence its economics. The economic analysis for three plant capacities, around 250, 500, and 1,000 MW, was conducted to examine economies of scale. Plants with capacities less than 250 MW were not considered because they only account for 4% of the total utility emissions in the Illinois Basin (Figure 2.1).

Illinois bituminous coal and western Powder River Basin (PRB) sub-bituminous coal are the two main coals used in utilities in the Illinois Basin. The techno-economics of these coals in the $CO_2$ capture process were therefore evaluated. The largest currently active coal mine in Illinois is the Galatia Mine located in southern Illinois. Because coal compositions can change from mine to mine in the same seam and from seam to seam, an average composition of the coal samples was used in this study. The PRB coal selected was the Antelope Mine coal. The ultimate analyses of the two coals are listed in Table 8.1 (Chen et al., 2002).

**Table 8.1. Typical composition of Illinois coal and PRB coal**

| | **Illinois coal** | | **PRB coal** | |
|---|---|---|---|---|
| **Composition, wt%** | **As-received basis** | **Dry basis** | **As-received basis** | **Dry basis** |
| Moisture | 6.08 | - | 26.7 | - |
| Carbon, C | 70.28 | 74.83 | 51.35 | 70.05 |
| Hydrogen, H | 4.77 | 5.08 | 3.59 | 4.9 |
| Nitrogen, N | 1.44 | 1.53 | 0.78 | 1.06 |
| Chlorine, Cl | 0.21 | 0.22 | 0.01 | 0.01 |
| Sulfur, S | 2.28 | 2.43 | 0.24 | 0.33 |
| Oxygen, O | 6.25 | 6.65 | 12.08 | 16.48 |
| Ash | 8.90 | 9.48 | 5.25 | 7.16 |
| Total | 100.00 | 100.00 | 100.00 | 100.00 |
| HHV[1], BTU/lb | 12,475 | 13,283 | 8,800 | 12,005 |

[1] HHV, higher heating value.

## 8.1. Reference Plant

### *8.1.1. Overall Process Descriptions*

The reference plant is a conventional air-blown power plant (Figure 8.1). The design and configuration of the reference PC plant were based on those of a standard plant developed in a U.S. DOE-funded study (Gilbert/Commonwealth Inc., 1995).

A PC power plant consists of a boiler, a super heater, a reheater, an economizer, and air heater components. Air at a 15 vol% excess is generally used for the combustion process. The temperature of the flue gas exiting the air preheater is about 295°C. The plant uses a single reheat sub-critical steam power cycle. In this study, the performance of the steam power cycle was based on the nominal 2,415 psi throttle steam conditions. The following nominal steam conditions were employed:

Main steam to high-pressure (HP) turbine: 2,415 psi/1,000°F;
Reheat steam to intermediate-pressure (IP) turbine: 545 psi/1,000°F.

The $NO_x$, fly ash, and $SO_2$ in the flue gas are removed by means of a selective catalytic reduction (SCR) unit, an electro-static precipitator (ESP) unit, and a limestone slurry forced oxidation (LSFO) unit if necessary. A 90% $NO_x$ removal efficiency was assumed for the SCR. A 99% fly ash removal efficiency was assumed for the ESP, and a size distribution of fly ash particles prior to the ESP is referred to that provided in the AP-42 guideline (U.S. EPA, 1998). The LSFO is applied only to Illinois coal with a removal efficiency of 95%. No flue gas desulfurization was assumed for the PRB coal in the reference plant.

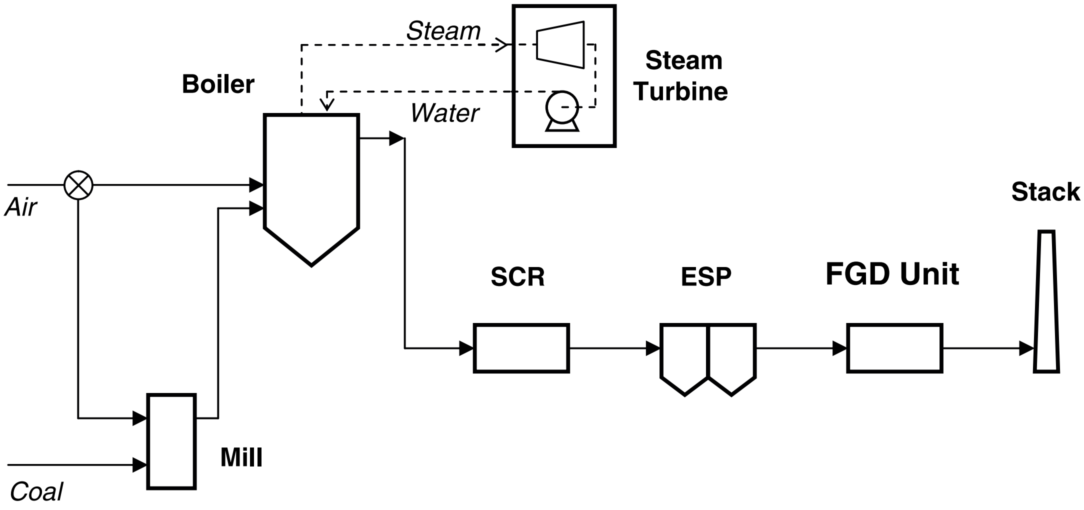


**Figure 8.1. Schematic diagram of the air-blown PC-fired power plant (Illinois coal)**

### *8.1.2. Process Simulation*

Process simulation software, CHEMCAD, developed by Chemstation Inc., was employed in this study. CHEMCAD can be used to perform steady-state simulations for mass and energy balances of the power plant.

The simulation was conducted for three main process areas in the power plant: a boiler system, a steam turbine system, and a gas cleaning system. Table 8.2 lists the performance of the main streams for the 533-MW (gross) plants burning Illinois and PRB coals. The detailed results of the mass/energy balance simulation are available in Appendix A1.

### Table 8.2. Main operating performance of the 533-MW (gross) plant

| | **Illinois coal** | | **PRB coal** |
|---|---|---|---|
| Combustion | | | |
| Air/coal equivalent ratio | 1.15 | | 1.15 |
| Air flow rate, lb/hr | 3,991,198 | | 4,029,474 |
| Coal feed rate, lb/hr | 360,611 | | 524,982 |
| | | | |
| Steam generation | | | |
| Hot reheat steam, lb/hr | 3,022,125 | | 3,022,125 |
| Superheat steam, lb/hr | 3,422,824 | | 3,422,824 |
| Steam condensate, lb/hr | 2,802,051 | | 2,802,051 |
| Main feed-water, lb/hr | 3,321,228 | | 3,321,228 |
| Heat duty in cooling tower, mBTU/hr | 2,178 | | 2,178 |
| | | | |
| Flue gas | **from boiler** | **from FGD** | **from boiler** |
| Flue gas volume, lb/h | 4,319,700 | 4,497,871 | 4,526,825 |
| Flue gas temperature, °F | 295 | 125 | 295 |
| Composition | | | |
| $N_2$, vol% | 75.69 | 70.73 | 71.73 |
| $O_2$, vol% | 2.73 | 2.47 | 2.49 |
| $CO_2$, vol% | 14.55 | 13.75 | 14.55 |
| $H_2O$, vol% | 6.78 | 13.04 | 11.15 |
| $SO_2$, vol% | 0.1774 | 0.0080 | 0.0257 |
| $NO_x$, vol% | 0.0325 | 0.0032 | 0.0444 |
| Fly ash flow rate, lb/hr | 25,676 | 257 | 22,049 |

Certain components of the power plant, such as pumps, fans, and conveyors, consume significant amounts of electricity. The auxiliary power uses of the coal handling, pulverizing, ash handling, and miscellaneous systems were scaled linearly based on coal consumption. For other components, energy usage was obtained from the process simulation. The results of the auxiliary power use analysis for 533-MW (gross) power plants are summarized in Table 8.3. For other plant capacities, auxiliary power uses are almost linearly proportional to the plant scale if the same coal is fired.

### Table 8.3. Auxiliary power use for 533-MW (gross) power plants

| | Illinois coal | PRB coal |
|---|---|---|
| **Auxiliary load, kW** | | |
| Coal handling | 233 | 339 |
| Pulverizers | 2,017 | 2,937 |
| Primary air fans | 1,199 | 1,212 |
| Forced draft fans | 1,142 | 1,154 |
| Induced draft fans | 4,921 | 5,122 |
| Seal air blowers | 45 | 46 |
| Steam turbine auxiliaries | 884 | 884 |
| Condensate pumps | 949 | 949 |
| Main feed pump[1] | 10,938 | 10,938 |
| Circulating water pumps | 4,187 | 4,187 |
| Cooling tower fans | 2,367 | 2,367 |
| Ash handling | 1,658 | 1,424 |
| Miscellaneous | 2,411 | 2,411 |
| Transformer loss | 1,215 | 1,215 |
| ESP | 1,259 | 1,319 |
| FGD | 7,500 | 0 |
| SCR | 2,750 | 2,750 |
| Subtotal | 34,737 | 28,316 |

[1] Driven by auxiliary steam turbine; electric equivalent not included in total.

### *8.1.3. Performance Summary*

The overall process performance for the reference air-blown PC plants without $CO_2$ capture is shown in Table 8.4. The power generation efficiency for the sub-critical PC plant without $CO_2$ capture is about 37 to 38%. The Illinois coal-fired power plant has a slightly higher generation efficiency than the PRB coal plant due to the lower moisture content of the Illinois coal.

### Table 8.4. Overall process performance of air-blown PC plants

| | Illinois coal | | | PRB coal | | |
|---|---|---|---|---|---|---|
| | **266 MW** | **533 MW** | **1,054 MW** | **266 MW** | **533 MW** | **1,054 MW** |
| Coal feed, as-received, lb/hr | 179,766 | 360,611 | 710,569 | 261,598 | 524,982 | 1,035,679 |
| Steam turbine power, MW | 266.4 | 533.2 | 1,053.6 | 266.4 | 533.2 | 1,053.6 |
| Auxiliary power use, MW | 17.6 | 34.7 | 67.9 | 14.4 | 28.3 | 55.1 |
| Net power output, MW | 248.7 | 498.5 | 985.7 | 251.9 | 504.9 | 998.5 |
| Net efficiency, % (HHV) | 37.8 | 37.8 | 37.9 | 37.3 | 37.3 | 37.4 |
| Heat rate, BTU/kWh (HHV) | 9,024 | 9,016 | 8,993 | 9,150 | 9,137 | 9,128 |

### *8.1.4. Cost Analysis*

The methodology of cost modeling and main financial assumptions are detailed in Appendix B.1. The cost model of conventional process areas in the PC power plant was that developed in U.S. DOE-funded studies (Gilbert/Commonwealth Inc., 1995; Office of Fossil Energy, U.S. DOE, 1999). Those of the LSFO unit and SCR unit were updated according to the latest EPA reports (Srivastava, 2000; Foerter and Jozewicz, 2001; U.S. EPA, 2002).

Table 8.5 summarizes the cost results of reference plants. The PRB coal-fired plant has a lower cost of electricity than the Illinois coal-fired plant mainly because no FGD unit was used for the power plant burning this coal (any new power plant burning PRB has to be equipped with a $SO_2$ sulfur removal process). As expected, the cost of electricity increases sharply with the decrease in plant capacity. In general, the estimated capital costs, operation and maintenance (O&M) costs, and costs of electricity are well within the range reported in references (Gilbert/Commonwealth Inc., 1995; Office of Fossil Energy, U.S. DOE, 1999).

**Table 8.5. Cost analysis of the air-blown PC reference plants**

| | **Illinois coal** | | | **PRB coal** | | |
|---|---|---|---|---|---|---|
| **Gross output (terminal)** | **266 MW** | **533 MW** | **1,054 MW** | **266 MW** | **533 MW** | **1,054 MW** |
| | | | | | | |
| Net output, MW | 248.7 | 498.5 | 985.7 | 251.9 | 504.9 | 998.5 |
| Capital cost, $/kW | | | | | | |
| Total plant cost | 1,403 | 1,119 | 914 | 1,339 | 1,019 | 877 |
| Total plant investment | 1,529 | 1,220 | 996 | 1,460 | 1,110 | 957 |
| Total capital requirement | 1,602 | 1,282 | 1,051 | 1,528 | 1,167 | 1,007 |
| | | | | | | |
| O&M costs, mills/kWh | | | | | | |
| Fixed O&M | 9.12 | 5.83 | 4.01 | 8.02 | 5.08 | 3.41 |
| Variable O&M | 4.91 | 3.14 | 2.16 | 4.32 | 2.74 | 1.84 |
| Consumables | 2.92 | 2.88 | 2.84 | 2.16 | 2.11 | 2.08 |
| Fuel cost | 10.84 | 10.85 | 10.81 | 10.38 | 10.40 | 10.37 |
| | | | | | | |
| Annual carrying charge, mills/kWh | 36.06 | 28.84 | 23.64 | 34.38 | 26.26 | 22.66 |
| | | | | | | |
| Levelized cost of electricity, mills/kWh | 63.84 | 51.54 | 43.47 | 59.27 | 46.58 | 40.36 |

## 8.2. Air-blown PC Plant with MEA Unit

### *8.2.1. Overall Process Descriptions*

The air-blown PC plant with $CO_2$ capture employs a sub-critical steam power cycle similar to that of the reference plant. The combustion and flue gas conditions are the same as the reference plant. The schematic diagram of the power plant equipped with an MEA plant is illustrated in Figure 8.2. The flue gas from the FGD unit flows vertically upward through the absorber countercurrent to the MEA solution. The cleaned gases then leave through the stack. The $CO_2$-rich MEA solution leaves the absorber and passes through a heat exchanger and then is further heated in a stripper using low-pressure steam to produce a concentrated $CO_2$ stream. The regenerated $CO_2$-lean solution is then cooled and re-circulates to the absorber. The $CO_2$ stream from the stripper is cooled in a condenser to remove moisture and then passes through a two-stage compressor with inter-stage coolers. The $CO_2$ stream is compressed to 83 bars through the three-stage compressors with inter-stage coolers. The high-pressure $CO_2$-rich product stream from the compressor is liquefied by cooling to ambient temperatures. Any residual trace gases are flashed off in a gas separator before transportation.

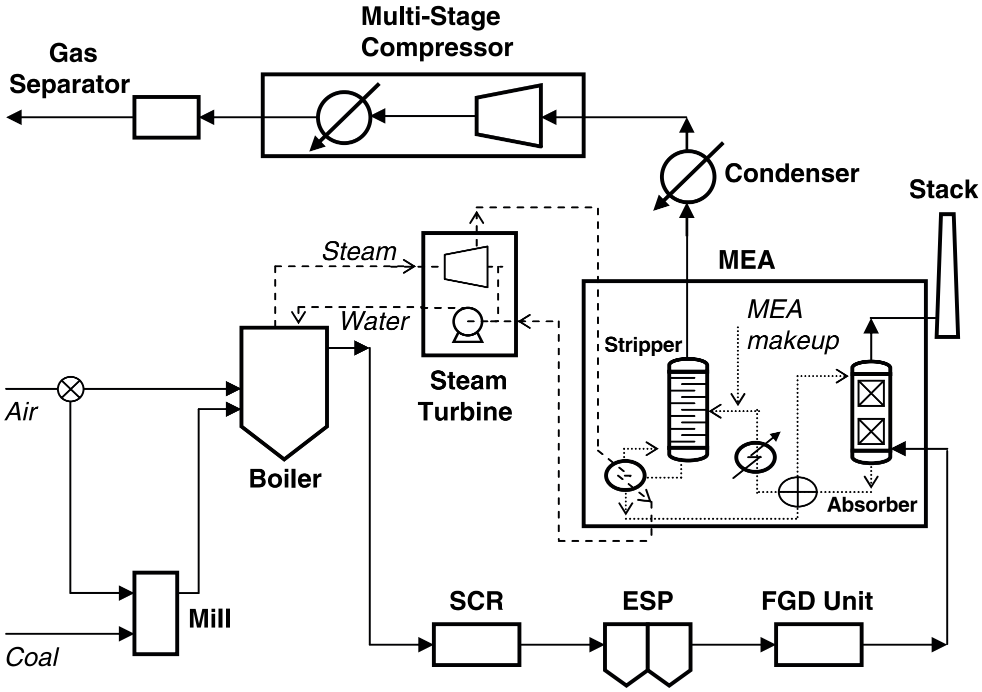


**Figure 8.2. Schematic diagram of the PC power plant with MEA unit**

Because steam is used for MEA regeneration, the operating condition of the steam turbine system is partly changed from that of the reference plant. The hot steam is drawn before entering the LP turbine at a pressure of 175 psi and then expands to 60 psi before reaching the stripper. The condensed water from the stripper is returned to a heat exchanger in the feed-water cycle.

MEA has strong affinity for absorbing various acid gases such as $SO_2$, $NO_x$, and HCl. Therefore, for cost-effective operation of an MEA unit, the flue gas should contain low levels of acid gases. In this regard, both for the Illinois coal and low-sulfur PRB coal, a FGD unit is required upstream of the MEA unit. The LSFO process was employed for the Illinois coal with an $SO_2$ removal efficiency of 95%. A lime spray dryer process was employed for the PRB coal with a removal efficiency of 90%.

### *8.2.2. Process Simulation*

Simulations of common process areas of the reference PC plant and of the PC + MEA plant were identical. The simulation of the MEA process was conducted using the K-value model and enthalpy model for amine. The chemical reactions in a $CO_2$-$H_2S$-amine system are described by the following reactions in CHEMCAD (Chemstations Inc., 2004):

$$RR'NH_2^+ \quad H^+ + RR'NH$$

$$RR'NCOO + H_2O \quad RR'NH + HCO_3^-$$

$$CO_2 + H_2O \quad HCO_2^- + H^+$$

$$HCO_3^- \leftrightarrow CO_2^{2-} + H^+$$

$H_2S \quad HS^- + H^+$

$HS^- \leftrightarrow S^{2-} + H^+$

$H_2O \quad H^+ + OH^-$

where R and R′ represent alcohol groups. The reaction equations are solved simultaneously to obtain the free concentration of $H_2S$ and $CO_2$. The partial pressure of $H_2S$ and $CO_2$ are calculated using the Henry's constants and free concentration in the liquid phase. The chemical reaction constants and Henry's constants are built into the software.

The typical operating parameters were referred to in some recent publications (White, 2002; Freguia and Rochelle, 2003; Alie et al., 2004). Main parameters used in simulation are listed:

Inlet flue gas temperature: ~129 $^{o}$F
MEA concentration: 30 wt%
Liquid/gas molar ratio: 5.0
Reboiler heat duty: ~3900 kJ/ kg of captured $CO_2$
LP steam: 60 psi, 475$^{o}$F
$CO_2$ capture efficiency: 90%
Inter-stage cooler temperature: 104$^{o}$F
Compressor efficiency: 85%
$CO_2$ product pressure: 83 bars

Table 8.6 lists the performance of main streams for the 533-MW (gross) plants fired with the Illinois and PRB coals, respectively. The detailed results of the mass/energy balance simulation are available in Appendix A2.

## Table 8.6. Main operating performance of the 533-MW (gross) plants

| | **Illinois coal** | **PRB coal** |
|---|---|---|
| Combustion | | |
| Air/coal equivalent ratio | 1.15 | 1.15 |
| Air flow rate, lb/hr | 3,991,198 | 4,029,474 |
| Coal feed rate, lb/hr | 360,611 | 524,982 |
| | | |
| Steam generation | | |
| Hot reheat steam, lb/hr | 3,022,125 | 3,022,125 |
| Superheat steam, lb/hr | 3,422,824 | 3,422,824 |
| Steam condensate, lb/hr | 1,421,687 | 1,331,136 |
| Main feed-water, lb/hr | 3,321,228 | 3,321,228 |
| Heat duty of cooling tower, mmBTU/hr | 1,104 | 1,034 |
| | | |
| MEA unit | | |
| Inlet flue gas volume, lb/hr | 4,472,743 | 4,683,939 |
| Gas inlet temperature, °F | 129 | 157 |
| MEA make-up, lb/hr | 2,794 | 1,827 |
| LP steam consumption, lb/hr | 1,380,364 | 1,470,914 |
| $CO_2$ captured, lb/hr | 834,748 | 894,354 |
| | | |
| $CO_2$ product | | |
| Flow rate, lb/hr | 835,083 | 894,680 |
| Temperature, °F | 68 | 68 |
| Composition | | |
| $O_2$, mol% | 0.001144 | 0.001172 |
| $N_2$, mol% | 0.018566 | 0.017991 |
| Ar, mol% | 0.000081 | 0.000068 |
| $CO_2$, mol% | 99.856110 | 99.859020 |
| $H_2O$, mol% | 0.120856 | 0.120859 |
| $SO_2$, mol% | 0.003230 | 0.000874 |
| $NO_x$, mol% | 0.000002 | 0.000003 |
| $NH_3$, mol% | 0.000011 | 0.000015 |

The MEA unit and $CO_2$ compressors consume considerable in-plant power. The results of the auxiliary power uses for 533-MW (gross) power plants with MEA unit are listed in Table 8.7. Induced draft fans and liquid pumps used in the MEA unit shared about 20% of the total auxiliary power use, while the $CO_2$ compression shared near half of the total.

### Table 8.7. Auxiliary power use of the 533-MW (gross) power plants with MEA unit

| | **Illinois coal** | **PRB coal** |
|---|---|---|
| Auxiliary load, kW | | |
| Coal handling | 233 | 339 |
| Pulverizers | 2,017 | 2,937 |
| Primary air fans | 1,199 | 1,212 |
| Forced draft fans | 1,142 | 1,154 |
| Induced draft fans | 4,921 | 5,122 |
| Seal air blowers | 45 | 46 |
| Steam turbine auxiliaries | 884 | 884 |
| Condensate pumps | 895 | 891 |
| Main feed pump[1] | 10,938 | 10,938 |
| Circulating water pumps | 2,124 | 1,989 |
| Cooling tower fans | 1,201 | 1,125 |
| Ash handling | 1,658 | 1,424 |
| Miscellaneous | 2,411 | 2,411 |
| Transformer loss | 1,215 | 1,215 |
| ESP | 1,259 | 1,319 |
| FGD | 7,500 | 3,500 |
| SCR | 2,750 | 2,750 |
| MEA: induced draft fans | 13,098 | 15,837 |
| pumps | 2,801 | 2,980 |
| $CO_2$ compressor | 35,423 | 37,951 |
| Subtotal | 82,777 | 85,086 |

[1] Driven by auxiliary steam turbine; electric equivalent not included in total.

### *8.2.3. Performance Summary*

The overall process performance for the PC + MEA unit for $CO_2$ capture is shown in Table 8.8. A large energy penalty exists due to the steam use for the MEA unit. The terminal power output of the steam turbine was reduced by about 17% compared to the reference plant without $CO_2$ capture. As mentioned before, the $CO_2$ compression and MEA unit also largely increased the auxiliary power use. As a result, the net efficiency of power generation was reduced to about 27% for the Illinois coal and 26% for the PRB coal, compared to 38% and 37% in the reference plants, respectively.

### Table 8.8. Overall process performance of air-blown PC plants with MEA

| | **Illinois coal** | | | **PRB coal** | | |
|---|---|---|---|---|---|---|
| | **266 MW** | **533 MW** | **1,054 MW** | **266 MW** | **533 MW** | **1,054 MW** |
| Coal feed as-received, lb/hr | 179,766 | 360,611 | 710,569 | 261,598 | 524,982 | 1,035,679 |
| Steam turbine power, MW | 220.3 | 440.9 | 871.6 | 217.4 | 434.9 | 859.4 |
| Auxiliary power use, MW | 41.6 | 82.8 | 163.6 | 42.8 | 85.1 | 168.2 |
| Net power output, MW | 178.7 | 358.1 | 708.0 | 174.6 | 349.8 | 691.2 |
| Net efficiency, % (HHV) | 27.2% | 27.2% | 27.3% | 25.8% | 25.8% | 25.9% |
| Heat rate, BTU/kWh (HHV) | 12,562 | 12,549 | 12,519 | 13,208 | 13,189 | 13,186 |

### *8.2.4. Cost Analysis*

*8.2.4.1. New Plant.* The cost model developed to estimate the MEA unit is detailed in Appendix B.1. The cost models used were those developed in recent DOE project reports (Parsons Energy and Chemicals Group Inc. and Wolk Integrated Technical Services, 2000; Parsons Infrastructure & Technology Group, Inc., 2002; Rao and Rubin, 2002; Rao et al., 2004). Cost models for other process areas are the same as for the reference plants.

The results of cost analyses for the power plants are summarized in Table 8.9. The definitions of the cost of $CO_2$ avoidance and the cost of $CO_2$ removal can be found in Appendix B1. From the table, the costs of electricity increase by about 77% for the Illinois coal and 95% for the PRB coal, compared to the cost of electricity for the reference plants, respectively. The $CO_2$ avoidance cost ranges from \$47 to \$67/ton for the selected plant capacities.

**Table 8.9. Cost analysis of the PC plants with MEA unit**

| | **Illinois coal** | | | **PRB coal** | | |
|---|---|---|---|---|---|---|
| | **266 MW** | **533 MW** | **1,054 MW** | **266 MW** | **533 MW** | **1,054 MW** |
| Net output, MW | 178.7 | 358.1 | 708.0 | 174.6 | 349.8 | 691.2 |
| Capital cost, \$/kW | | | | | | |
| Total plant cost | 2,414 | 1,938 | 1,586 | 2,461 | 2,015 | 1,627 |
| Total plant investment | 2,632 | 2,112 | 1,729 | 2,683 | 2,197 | 1,774 |
| Total capital requirement | 2,761 | 2,224 | 1,828 | 2,810 | 2,307 | 1,870 |
| | | | | | | |
| O&M[1] costs, mills/kWh | | | | | | |
| Fixed O&M | 14.65 | 9.56 | 6.69 | 14.67 | 9.60 | 6.55 |
| Variable O&M | 7.89 | 5.15 | 3.60 | 7.90 | 5.17 | 3.53 |
| Consumables | 11.07 | 11.20 | 11.14 | 9.29 | 9.22 | 9.19 |
| Fuel cost | 15.09 | 15.11 | 15.05 | 14.99 | 15.01 | 14.98 |
| | | | | | | |
| Annual carrying charge, mills/kWh | 62.14 | 50.05 | 41.13 | 63.24 | 51.91 | 42.08 |
| | | | | | | |
| Levelized cost | | | | | | |
| Cost of electricity, mills/kWh | 110.84 | 91.06 | 77.62 | 110.09 | 90.90 | 76.34 |
| Cost of $CO_2$ avoidance, \$/t $CO_2$ | 64.73 | 54.38 | 47.15 | 66.94 | 58.31 | 47.44 |
| Cost of $CO_2$ removal, \$/t $CO_2$ | 44.48 | 37.36 | 32.39 | 44.09 | 38.40 | 31.22 |

[1] Operating and maintenance.

## Table 8.10. Cost analysis of the MEA process

| | Illinois coal | | | PRB coal | | |
|---|---|---|---|---|---|---|
| | **267 MW** | **533 MW** | **1,054 MW** | **267 MW** | **533 MW** | **1,054 MW** |
| Total plant cost, $1,000 | 89,252 | 147,331 | 240,095 | 92,286 | 152,384 | 248,543 |
| O&M cost, $1,000/year | | | | | | |
| Fixed O&M | | | | | | |
| Operating labor | 701 | 701 | 701 | 701 | 701 | 701 |
| Maintenance | 2,231 | 3,683 | 6,002 | 2,307 | 3,810 | 6,214 |
| Admin. & support labor | 478 | 652 | 931 | 487 | 667 | 956 |
| Subtotal | 3,410 | 5,036 | 7,634 | 3,495 | 5,178 | 7,870 |
| Variable O&M | | | | | | |
| Amine make-up | 4,847 | 9,722 | 19,158 | 3,168 | 6,356 | 12,542 |
| Inhibitor | 969 | 1,944 | 3,832 | 634 | 1,271 | 2,508 |
| Acoustic soda | 301 | 603 | 1,189 | 320 | 643 | 1,268 |
| Activated carbon | 173 | 348 | 686 | 185 | 371 | 732 |
| Waste disposal | 52 | 518 | 1,020 | 30 | 61 | 121 |
| Water | 1,330 | 2,660 | 5,319 | 1,602 | 3,204 | 6,408 |
| Subtotal | 7,673 | 15,796 | 31,202 | 5,939 | 11,906 | 23,579 |

Table 8.10 presents the cost breakdown for the MEA process. For the example, for the 533-MW (gross) power plant, the installation of the MEA process resulted in an 26% increase in total capital costs and a 30% increase in operating and maintenance costs.

*8.2.4.2. MEA Retrofit.* In the retrofit case, all existing capital cost and operating and maintenance costs were assumed to be same before and after retrofit. Therefore, the cost analysis for retrofit only considers new expenditures in retrofit for $CO_2$ capture, which mainly consist of the costs related to the MEA and $CO_2$ compression facilities.

Three assumptions were adopted in this study. First, a retrofit factor of 1.2 was used to estimate the retrofit capital cost. Second, a 15-year remaining life was assumed for the retrofitted plant, and, based on the same financial criteria as for the new plant, a capital factor of 0.155 was adopted for amortizing the new capitals. Third, the reduced power generation due to $CO_2$ capture was assumed to be balanced with construction of a new NGCC plant. An electricity cost of 34 mills/kWh for NGCC was adopted in this study. The $CO_2$ emissions from the new NGCC plant were deducted in estimating the net $CO_2$ reduction.

The results are shown in the following table. The $CO_2$ avoidance cost in the MEA retrofit ranges from $35 to $45/t. The costs are lower than the new plants. This is a benefit of the construction of the new NGCC plant with electricity generation cost and $CO_2$ emissions both significantly lower than those of the coal-fired power plant.

**Table 8.11. Cost analysis of the PC plants with MEA retrofit**

| | **Illinois coal** | | | **PRB coal** | | |
|---|---|---|---|---|---|---|
| | **266 MW** | **533 MW** | **266 MW** | **533 MW** | **266 MW** | **533 MW** |
| Net output, MW | 249 | 498 | 986 | 252 | 505 | 998 |
| Retrofit TCR, $1,000 | 102,416 | 170,008 | 278,741 | 105,261 | 174,427 | 285,771 |
| Increase of levelized capital, $1,000/yr | 18,619 | 30,908 | 50,675 | 19,136 | 31,711 | 51,953 |
| Increase of O&M, $1,000/yr | 11,083 | 20,832 | 38,836 | 11,794 | 20,867 | 37,116 |
| Electricity loss, $1,000/yr | 14,602 | 29,269 | 57,889 | 16,134 | 32,347 | 64,064 |
| | | | | | | |
| $CO_2$ reduction, kt/yr | 1,157 | 2,320 | 4,572 | 1,232 | 2,472 | 4,878 |
| $CO_2$ emissions of NGCC, kt/yr | 156 | 314 | 620 | 173 | 347 | 686 |
| Cost of $CO_2$ avoided, $/t | 44.30 | 40.37 | 37.30 | 44.43 | 39.95 | 36.53 |

## 8.3. Oxy-combustion Process

### *8.3.1. Process Descriptions*

A sub-critical steam power cycle similar to the air-blown PC plant is employed in the oxy-combustion process (Figure 8.3). The steam turbine generation system is based on the nominal 2,415 psi throttle steam conditions. The boiler for the oxy-combustion is similar to that for the air-blown combustion because in the former process a portion (70 to 75%) of the $CO_2$-rich combustion flue gas exiting the boiler is recycled to the boiler to provide an oxygen-containing gas comparable to that of an air-blown PC.

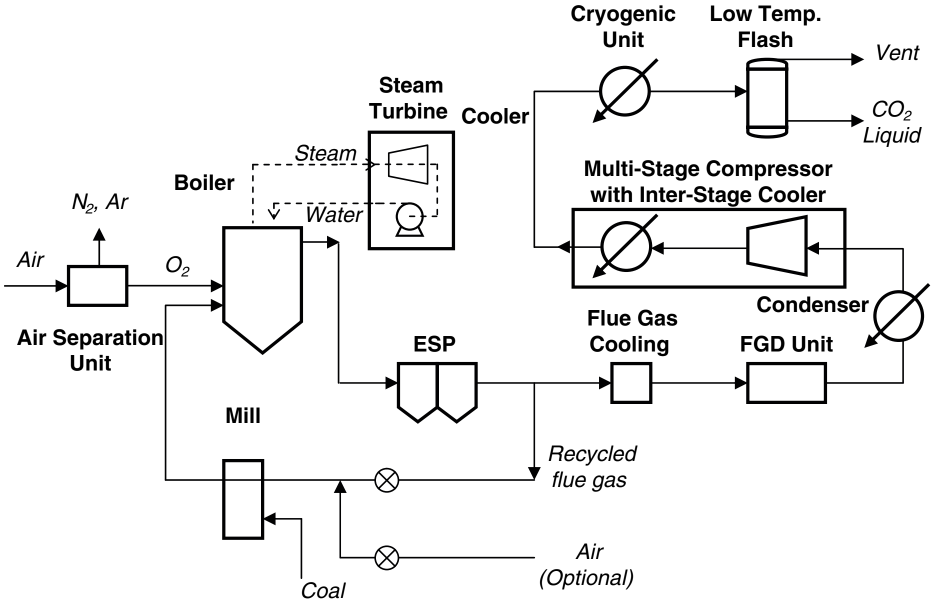


**Figure 8.3. Schematic diagram of the oxy-combustion power plant**

The replacement of combustion air with oxygen provides a nitrogen-free atmosphere in the boiler. A cryogenic ASU for separating nitrogen from oxygen is employed to produce a 95% purity oxygen stream. When oxygen instead of air is used, the volume of the flue gas is reduced fivefold. The flue gas generated in the OEC process is also highly enriched (>90% dry) with $CO_2$.

The $NO_x$ removal by the SCR process was not considered for the OEC process because $NO_x$ emissions as low as 0.15 lb/mmBTU can be obtained without $NO_x$ control (Châtel-Pélage et al., 2004).

A FGD unit is required for power plants burning both coals. An LSFO process is considered for the Illinois coal with a removal efficiency of 95%, and an LSD process is considered for the PRB coal with an efficiency of 90%. Because the flue gas temperature in the oxy-combustion process (~395°F) is higher than that of the air-blown process (~295°F), flue gas cooling is necessary before the gas enters the FGD unit.

The flue gas leaving the FGD unit is condensed to remove most moisture in the gas. The flu gas then passes through the three-stage compressors, where remaining moisture is also removed in the inter-stage coolers. The high-pressure gas from the compressor is further cooled to about –33°C in a cryogenic unit using ammonia as the cooling medium. Finally, the cold liquid $CO_2$ stream enters a low-temperature flash process to remove trace amounts of any gaseous impurity before transportation.

### *8.3.2. Process Simulation*

The process simulation was conducted for various process areas (i.e., the combustion system, steam turbine system, gas cleaning system, and $CO_2$ purification and compression system). The ASU was not simulated; its performance data were provided by American Air Liquide (personal communications, 2004)

The operating parameters selected for simulation were adopted using those reported in recent studies (Birkestad, 2002; Anderson and Maksinen, 2002; Singh et al., 2003):

$O_2$/fuel equivalent ratio: 1.03
Oxygen purity: 95%
Flue gas recirculation ratio: 72%
Flue gas temperature: 395°F
Condenser operating temperature: 104°F
Inter-stage cooler temperature: 104°F
$CO_2$ product pressure: 83 bars
Compression efficiency: 85%
Cryogenic temperature: –33°C
Cryogenic medium: ammonia
Low temperature flash (LTF): 83 bars, –33°C
$CO_2$ recovery in LTF: ~97%

Table 8.11 lists the main operating performance for the 533-MW (gross) oxy-combustion power plants. The detailed results of mass/energy balance calculations are listed in Appendix A3.

### Table 8.12. Main operating performance for the 533-MW (gross) oxy-combustion plants

| | **Illinois coal** | | **PRB coal** | |
|---|---|---|---|---|
| Combustion | | | | |
| $O_2$/fuel equivalent ratio, lb/hr | 1.03 | | 1.03 | |
| $O_2$ flow rate (95% purity), lb/hr | 843,599 | | 817,540 | |
| Coal feed rate, lb/hr | 348,375 | | 506,615 | |
| Flue gas recycle ratio, lb/hr | 0.72 | | 0.72 | |
| | | | | |
| Steam generation | | | | |
| Hot reheat steam, lb/hr | 3,022,125 | | 3,022,125 | |
| Superheat steam, lb/hr | 3,422,824 | | 3,422,824 | |
| Steam condensate, lb/hr | 2,802,051 | | 2,802,051 | |
| Main feedwater, lb/hr | 3,321,228 | | 3,321,228 | |
| Heat duty of cooling tower, mBTU/hr | 2,178 | | 2,178 | |
| | | | | |
| Gas stream | **Flue gas** | **$CO_2$ product** | **Flue gas** | **$CO_2$ product** |
| Flow rate, lb/hr | 1,161,068 | 874,274 | 1,297,622 | 927,390 |
| Temperature, °F | 395 | 68 | 395 | 68 |
| Composition | | | | |
| $O_2$, mol% | 2.93 | 0.0236 | 0.33 | 0.0232 |
| $N_2$, mol% | 0.57 | 0.0016 | 1.80 | 0.0020 |
| Ar, mol% | 3.34 | 0.0382 | 2.97 | 0.0353 |
| $CO_2$, mol% | 63.01 | 99.7366 | 53.74 | 99.7887 |
| $H_2O$, mol% | 29.38 | 0.1383 | 41.05 | 0.1318 |
| $SO_2$, mol% | 0.77 | 0.0609 | 0.10 | 0.0176 |
| $NO_x$, mol% | 0.04 | 0.0009 | 0.05 | 0.0015 |
| Fly ash flow rate, lb/hr | 24,804 | - | 21,278 | - |

The auxiliary power use in the oxy-combustion power plants is presented in Table 8.12. The auxiliary power use of the ASU is based on 0.394 kWh/Nm$^3$ pure $O_2$ and is almost independent of the $O_2$ purity, according to American Air Liquide. The ASU and $CO_2$ compression processes consume about 20% and 10% of the total gross output of the power plant, respectively. However, due to the reduced flue gas volume, the auxiliary power use by ESP, FGD, and induced draft fans decreases fourfold compared to the air-blown PC plant.

### Table 8.13. Auxiliary power usage in the 533-MW (gross) oxy-combustion power plants

| | **Illinois coal** | **PRB coal** |
|---|---|---|
| Auxiliary load summary, KWe | | |
| Coal handling | 225 | 328 |
| Pulverizers | 1,949 | 2,834 |
| Primary air fans | 1,133 | 1,221 |
| Forced draft fans | 1,079 | 1,163 |
| Induced draft fans | 1,353 | 1,470 |
| Seal air blowers | 43 | 46 |
| Steam turbine auxiliaries | 884 | 884 |
| Condensate pumps | 949 | 949 |
| Main feed pump[1] | 10,938 | 10,938 |
| Circulating water pumps | 4,187 | 4,187 |
| Cooling tower fans | 2,367 | 2,367 |
| Ash handling | 1,602 | 1,374 |
| Miscellaneous | 2,411 | 2,411 |
| Transformer loss | 1,215 | 1,215 |
| ESP | 338 | 376 |
| FGD | 2,016 | 998 |
| Oxy-combustion-specific | | |
| Flue gas recycle fan | 1,634 | 1,817 |
| Water pumps | 200 | 507 |
| Cryogenic unit | 13,064 | 13,710 |
| ASU | 100,230 | 101,081 |
| Compressor | 48,999 | 49,258 |
| **Subtotal** | 185,879 | 188,196 |

[1] Driven by auxiliary steam turbine; electric equivalent not included in total.

### *8.3.3. Performance Summary*

Table 8.13 lists the overall performances of the oxy-combustion power plants. The generation efficiency of the oxy-combustion process is about 26 to 27% compared to about 37 to 38% for the air-blown plant without $CO_2$ capture (due to ASU). However, the oxy-combustion process is slightly more efficient than the air-blown plant + MEA unit (Table 8.8).

## Table 8.14. Overall process performances of the oxy-combustion plants

| | Illinois coal | | | PRB coal | | |
|---|---|---|---|---|---|---|
| | **266 MW** | **533 MW** | **1054 MW** | **266 MW** | **533 MW** | **1054 MW** |
| Coal feed, as-received, lb/hr | 17,3558 | 348,375 | 686,672 | 252,670 | 506,615 | 991,497 |
| Steam turbine power, MW | 266.4 | 533.2 | 1,053.6 | 266.4 | 533.2 | 1,053.6 |
| ASU power, MWe | 49.9 | 100.2 | 197.6 | 50.4 | 101.2 | 197.8 |
| Other aux. power use, MW | 43.1 | 85.6 | 169.8 | 43.9 | 87.1 | 172.6 |
| Net power output, MW | 173.3 | 347.4 | 686.2 | 172.1 | 345.0 | 683.1 |
| Net efficiency, % (HHV) | 27.3% | 27.3% | 27.3% | 26.4% | 26.4% | 26.7% |
| Heat rate, BTU/kWh (HHV) | 12,512 | 12,490 | 12,483 | 12,921 | 12,919 | 12,772 |

### *8.3.4. Cost Analysis*

*8.3.4.1. New Plant.* The cost model developed to estimate the oxy-combustion process is detailed in Appendix B1. The cost data for the ASU was provided by American Air Liquide (personal communications, 2004; Singh et al., 2003; Châtel-Pélage et al., 2004; Sangras et al., 2004 ).

The cost was estimated for 266-MW, 533-MW, and 1,054-MW (gross) oxy-combustion power plants (Table 8.14). The cost of electricity increased by about 56% for the Illinois coal and about 70% for the PRB coal compared to the air-blown PC reference plants. The costs of $CO_2$ avoidance range from \$31 to \$43/t of $CO_2$ for the Illinois coal and from \$33 to \$46/t of $CO_2$ for the PRB coal. These costs are about 70% of those of the PC plant with a MEA unit.

## Table 8.15. Cost analysis of the oxy-combustion power plants

| | Illinois coal | | | PRB coal | | |
|---|---|---|---|---|---|---|
| | **266 MW** | **533 MW** | **1,054 MW** | **266 MW** | **533 MW** | **1,054 MW** |
| Net output, MW | 173.3 | 347.4 | 686.2 | 172.1 | 345.0 | 683.1 |
| Capital cost, \$/kW | | | | | | |
| Total plant cost | 2,221 | 1,836 | 1,528 | 2,230 | 1,863 | 1,549 |
| Total plant investment | 2,422 | 2,001 | 1,666 | 2,431 | 2,031 | 1,688 |
| Total capital requirement | 2,532 | 2,096 | 1,749 | 2,540 | 2,125 | 1,771 |
| | | | | | | |
| O& M costs, mills/kWh | | | | | | |
| Fixed O&M | 14.57 | 9.41 | 6.52 | 14.50 | 9.41 | 6.51 |
| Variable O&M | 7.85 | 5.07 | 3.51 | 7.81 | 5.07 | 3.51 |
| Consumables | 3.85 | 3.84 | 3.85 | 3.82 | 3.81 | 3.81 |
| Fuel cost | 15.02 | 15.04 | 15.01 | 14.68 | 14.68 | 14.51 |
| | | | | | | |
| Annual carrying charge, mills/kWh | 56.97 | 47.16 | 39.36 | 57.17 | 47.83 | 39.86 |
| | | | | | | |
| Levelized cost | | | | | | |
| Electricity, mills/kWh | 98.26 | 80.52 | 68.25 | 97.98 | 80.81 | 68.20 |
| $CO_2$ avoidance, \$/t of $CO_2$ | 43.00 | 36.19 | 31.05 | 46.04 | 40.66 | 33.14 |
| $CO_2$ removal, \$/t of $CO_2$ | 30.58 | 25.71 | 22.03 | 32.05 | 28.33 | 23.16 |

## Table 8.16. Cost analysis of specific components in oxy-combustion process

| | Illinois coal | | | PRB coal | | |
|---|---|---|---|---|---|---|
| | **266 MW** | **533 MW** | **1,054 MW** | **266 MW** | **533 MW** | **1,054 MW** |
| Total plant cost, $1,000 | | | | | | |
| ASU | 44,672 | 89,667 | 176,741 | 45,100 | 90,429 | 176,978 |
| Flue gas cooling | 1,528 | 2,403 | 3,733 | 507 | 3,185 | 4,934 |
| Condenser | 3,686 | 6,071 | 10,000 | 3,686 | 6,071 | 10,000 |
| Cryogenic unit | | | | | | |
| Flashs | 1,908 | 3,372 | 5,970 | 1,969 | 3,590 | 6,196 |
| Heat exchangers | 10,179 | 14,808 | 25,065 | 10,504 | 15,279 | 25,863 |
| $NH_3$ compressor | 5,836 | 10,401 | 14,161 | 6,017 | 10,736 | 14,608 |
| $NH_3$ expander | 251 | 457 | 773 | 261 | 475 | 803 |
| Compression | 18,721 | 33,463 | 45,628 | 18,769 | 33,549 | 45,748 |
| Subtotal | 86,781 | 160,641 | 282,070 | 86,814 | 163,314 | 285,131 |
| O&M cost | | | | | | |
| 1. FOM | | | | | | |
| Operating labor | 1,402 | 1,402 | 1,402 | 1,402 | 1,402 | 1,402 |
| Maintenance | 2,170 | 4,016 | 7,052 | 2,170 | 4,083 | 7,128 |
| Admin. & support labor | 681 | 902 | 1,267 | 681 | 910 | 1,276 |
| Subtotal | 4,252 | 6,320 | 9,720 | 4,253 | 6,395 | 9,806 |
| 2. VOM | | | | | | |
| Cooling water | 526 | 1,051 | 2,102 | 652 | 1,303 | 2,606 |

The cost breakdown of the process areas specific to the oxy-combustion process is listed in Table 8.15. For the example, for a 533-MW (gross) plant, the oxy-combustion specific components require a 29% increase in capital costs and a 12% increase in operating and maintenance costs.

*8.3.4.2. Retrofit with Oxy-combustion.* The cost analysis for oxy-combustion retrofit only considers new expenditure for $CO_2$ capture. The new costs are mainly related to the ASU, moisture condensation, and $CO_2$ compression facilities.

The assumptions same as for the MEA retrofit (i.e., the retrofit factor of 1.2) the remaining 15-year life time, and the new NGCC plant for supplement the generation, were used in the oxy-combustion retrofit.

The results are shown in Table 8.17. The $CO_2$ avoidance cost in the oxy-combustion retrofit ranges from $30 to $36/t. The costs for the Illinois coal are only slightly lower than those for the new plants. Despite the advantages of the new NGCC plant, the existing FGD unit wouldn't benefit from the reduced volume of flue gas in the retrofit case.

Table 8.17. Cost analysis of the oxy-combustion retrofit

| | **Illinois coal** | | | **PRB coal** | | |
|---|---|---|---|---|---|---|
| | **266 MW** | **533 MW** | **1,054 MW** | **266 MW** | **533 MW** | **1,054 MW** |
| Gross output, MW | 179 | 358 | 708 | 175 | 350 | 691 |
| Retrofit TCR, $1,000 | 97,452 | 180,290 | 316,540 | 97,529 | 183,363 | 319,297 |
| Increase of levelized capital, $1,000/yr | 17,717 | 32,777 | 57,547 | 17,731 | 33,335 | 58,048 |
| Increase of O&M, $1,000/yr | 4,777 | 7,371 | 11,822 | 4,904 | 7,698 | 9,806 |
| Electricity loss, $1,000/yr | 15,718 | 31,511 | 62,438 | 16,644 | 33,333 | 65,747 |
| Coal savings of OEC, kt/yr | –571 | –1,125 | –2,198 | –547 | –1,126 | –2,709 |
| | | | | | | |
| $CO_2$ reduction, kt/yr | 1,196 | 2,401 | 4,733 | 1,275 | 2,556 | 5,035 |
| $CO_2$ emissions of NGCC, kt/yr | 168 | 338 | 669 | 178 | 357 | 704 |
| Cost of $CO_2$ avoided, $/t | 36.63 | 34.18 | 31.89 | 35.33 | 33.31 | 30.23 |

## 8.4. IGCC + Physical absorption

### 8.4.1. Overall Process Descriptions

The IGCC plant is based on the General Electric H-type advanced turbine system combustion turbine coupled with a heat recovery unit that generates steam for a single steam turbine generator. An E-Gas gasifier is chosen as the basis for this IGCC configuration. The schematic diagram of the IGCC plant with a Selexol unit is displayed in Figure 8.4 (Parsons Energy and Chemicals Group Inc. and Wolk, 2000).

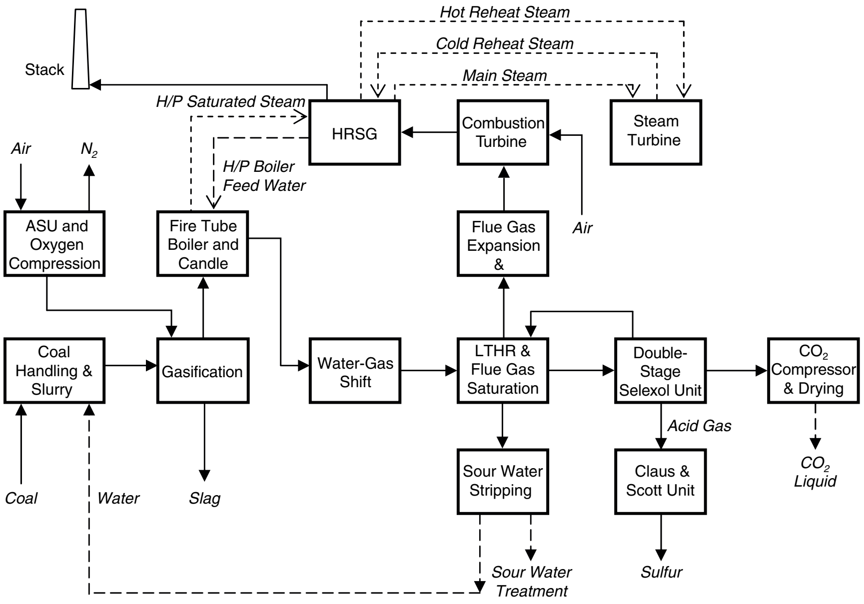


Figure 8.4. Schematic diagram of the Destec IGCC plant with Selexol unit

Coal-water slurry containing 63 wt% coal and a 95% oxygen stream from the ASU are fed to the gasifier. The E-gas gasifier is a two-stage, oxygen-blown, entrained flow, slagging gasifier. In the first stage, partial combustion of the coal maintains a temperature of about 2,500°F in the gasifier. Most of the coal reacts with steam to produce a raw fuel gas. Additional coal-water slurry added to the second stage undergoes de-volatilization, pyrolysis, and partial gasification to cool the raw fuel gas and enhance its heating value.

The fuel gas is further cooled using a fire tube boiler integrated with the gasifier and other heat exchangers to generate high-pressure, superheated steam for power generation. A candle filter is used to remove particulates, which are recycled to the gasifier. The fuel gas stream is mixed with steam and passes through the three-stage shift reactors at reduced entry temperatures. Heat exchangers between reaction stages help maintain the temperatures.

The fuel gas is treated in a double-stage Selexol unit to remove $H_2S$ and $CO_2$ (Figure 8.5). $H_2S$ is preferentially removed in the first absorber (705 psi, 105°F), and $CO_2$ is removed in the second absorber (90°F, 700 psi). The rich solvent from the first absorber enters two sets of flash where $CO_2$ is recovered at 50 psi in the first flash and the remainder at atmospheric pressure in the second flash with a total $CO_2$ removal efficiency of 98 to 99%. The $H_2S$-rich solvent is regenerated in a stripper through indirect heating by low-pressure steam in a reboiler. The resulting stripper acid gas stream is sent to a Claus unit.

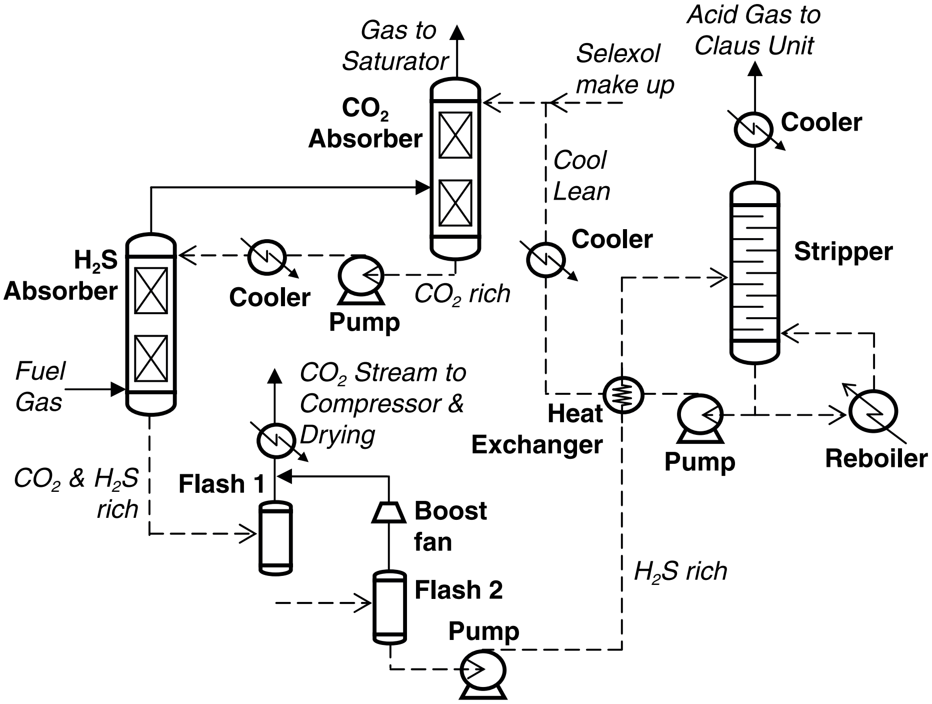


Figure 8.5. Schematic diagram of the Selexol unit for $CO_2$ removal

The cleaned gas passes to a moisture saturator to humidify the fuel gas and also increase its sensible heat content before entering an air-fired combustion turbine. The exhaust heat from the turbine is recovered in an HRSG to produce steam used in the steam turbines. The electricity generation is based on a combined cycle.

### *8.4.2. Process Simulation*

The IGCC simulation included two process blocks. One included the gasifier, the shift reactors, the saturators, the Selexol unit, and the gas combustion turbine, and the other included the steam turbine system. The heat exchanges involving in the combined cycle and other heat exchangers are balanced between the two blocks. The Selexol unit was not simulated in detail due to lack of thermodynamic data in CHEMCAD. However, a simple mass/heat balance was included assuming 99% and 98% recovery of $H_2S$ and $CO_2$, respectively, from the fuel gas after shift reactions.

The main operating parameters used in simulation were those reported in related recent studies (Parsons Energy and Chemicals Group Inc. and Wolk Integrated Technical Services, 2000; EG&G, 2000; IEA Greenhouse Gas R&D Programme, 2003). The overall $CO_2$ removal efficiency was assumed to be 90%. Table 8.18 lists the main performance of a 536-MW (gross) IGCC plants. The detailed results of the mass/energy balance simulations are available in Appendix A4.

**Table 8.18. Main operating performance of the 536-MW (gross) IGCC plants**

| | **Illinois coal** | | **PRB coal** | |
|---|---|---|---|---|
| | | | | |
| Coal feed rate, lb/h | 324,152 | | 466,262 | |
| $O_2$ (95% purity) flow rate, lb/h | 268,911 | | 288,533 | |
| Water make-up, lb/h | 584,349 | | 584,349 | |
| | | | | |
| Raw fuel gas, lb/h | 777,543 | | 934,997 | |
| Slag, lb/h | 28,850 | | 24,471 | |
| Steam to gas shift reactor, lb/h | 293,937 | | 293,937 | |
| Air to gas combustor, lb/h | 5,140,434 | | 5,022,433 | |
| $CO_2$ captured, lb/h | 738,801 | | 788,136 | |
| $CO_2$ capture efficiency, % | 89.6 | | 89.8 | |
| | | | | |
| **Gas composition** | **Fuel gas** | **Flue gas** | **Fuel gas** | **Flue gas** |
| $H_2O$, % | 17.06 | 21.86 | 25.42 | 22.39 |
| $O_2$, % | - | 10.88 | - | 10.74 |
| $N_2$, % | 0.56 | 65.32 | 0.37 | 64.93 |
| Ar, % | 0.97 | 1.00 | 0.88 | 1.01 |
| $H_2$, % | 31.21 | 0.00021 | 29.22 | 0.00024 |
| CO, % | 37.30 | 0.00003 | 29.63 | 0.000034 |
| $CO_2$, % | 10.96 | 0.93 | 13.84 | 0.92 |
| $CH_4$, % | 1.30 | - | 0.57% | - |
| $H_2S$, % | 0.5618 | - | 0.0742% | - |
| $NH_3$, % | 0.0087 | - | 0.0064% | - |

## Table 8.19. Auxiliary power use in IGCC plants with Selexol process

| | **Illinois coal** | **PRB coal** |
|---|---|---|
| Gross power summary, kW | | |
| Gas turbine power | 358,618 | 358,541 |
| Steam turbine power | 174,601 | 174,601 |
| Generator loss | –7,997 | –7,996 |
| Turbo-set power | 525,222 | 525,146 |
| Fuel gas expander power | 10,392 | 10,611 |
| Gross power | 535,614 | 535,757 |
| | | |
| Auxiliary load summary, kW | | |
| Coal handling and conveying | 366 | 526 |
| Coal milling | 843 | 1,213 |
| Coal slurry pumps | 168 | 235 |
| Slag handling and dewatering | 145 | 123 |
| Recycle gas blower | 113 | 127 |
| Air separation plant | 26,138 | 28,045 |
| Oxygen boost compressor | 15,153 | 16,259 |
| Selexol plant | 7,551 | 7,977 |
| Claus/TGTU[1] | 102 | 15 |
| Tail gas recycle | 1,016 | 154 |
| Humidification tower pump | 63 | 65 |
| Humidifier makeup pump | 141 | 141 |
| Low-pressure $CO_2$ compressor | 823 | 865 |
| High-pressure $CO_2$ compressor | 24,532 | 25,782 |
| Condensate pumps | 329 | 329 |
| High-pressure boiler feed pump | 3,294 | 3,294 |
| Low-pressure boiler feed pump | 61 | 61 |
| Miscellaneous balance of plant | 1,092 | 1,092 |
| Gas turbine auxiliaries | 623 | 623 |
| Steam turbine auxiliaries | 244 | 244 |
| Circulating water pumps | 1,636 | 1,636 |
| Cooling tower fans | 1,028 | 1,028 |
| Flash bottoms pump | 55 | 55 |
| Transformer loss | 1,660 | 1,661 |
| Subtotal | 87,175 | 91,549 |

[1] TGTU, tail gas treating unit.

The auxiliary power use for the 535 (gross) IGCC plants with $CO_2$ capture are summarized in Table 8.19. The total auxiliary power use amounts to about 17% of the gross power output, of which the Selexol unit consumes about 1.4 to 1.5%, and $CO_2$ compression consumes about 5%.

### *8.4.3. Performance Summary*

The overall performance data for IGCC plants equipped with Selexol units are listed in Table 8.20. The net generation efficiency for an Illinois coal is about 38% which is comparable to that for the PRB coal (~37%). The net generation efficiency of the IGCC plant decreases significantly due to the power use for $CO_2$ capture and compression compared to the net efficiency of 45 to 56% in the IGCC plant without $CO_2$ capture. However, the net efficiency of IGCC plant with Selexol unit is comparable to an air-blown PC plant without $CO_2$ removal (see Table 8.5).

**Table 8.20. Overall performance of an IGCC plant with $CO_2$ removal**

| | **Illinois coal** | | | **PRB coal** | | |
|---|---|---|---|---|---|---|
| | **268 MW** | **536 MW** | **1,071 MW** | **268 MW** | **536 MW** | **1,071 MW** |
| Coal feed as received, lb/hr | 162,097 | 324,152 | 648,133 | 233,163 | 466,262 | 932,271 |
| Oxygen feed, lb/hr | 134,473 | 268,911 | 537,680 | 144,286 | 288,533 | 576,910 |
| Water, lb/hr | 292,213 | 584,349 | 1,168,389 | 292,214 | 584,349 | 1,168,381 |
| Gross power, MWe | | | | | | |
| Steam turbine | 179.3 | 358.6 | 717.4 | 179.2 | 358.5 | 717.3 |
| Gas turbine | 87.3 | 174.6 | 349.3 | 87.3 | 174.6 | 349.3 |
| Generator loss | –4.0 | –8.0 | –16.0 | –4.0 | –8.0 | –16.0 |
| Fuel gas expander | 5.2 | 10.4 | 20.8 | 5.3 | 10.6 | 21.2 |
| Aux. power use, MWe | –43.6 | –87.2 | –174.4 | –45.8 | –91.5 | –183.1 |
| Net power, MWe | 224.2 | 448.4 | 897.1 | 222.1 | 444.2 | 888.7 |
| Net efficiency, HHV % | 37.80 | 37.81 | 37.83 | 36.90 | 36.91 | 36.93 |
| Heat rate, BTU/kWh (HHV) | 9,020 | 9,017 | 9,013 | 9,239 | 9,237 | 9,232 |

### *8.4.4. Cost Analysis*

The cost model developed to estimate the IGCC plant with Selexol unit is detailed in Appendix B.2. The estimation approach and financial assumptions were mainly referred to the studies on the E-gas gasifier-based IGCC technology (Parsons Energy and Chemicals Group Inc. and Wolk Integrated Technical Services, 2000; EG&G, 2000; Akunuri, 2000).

The cost was estimated for the 268-MW, 536-MW and 1,071-MW (gross) IGCC plants with the Selexol process. The results are listed in Table 8.21. The cost of electricity for the Illinois coal is about 10% lower than the PRB coal, mainly because the sulfur by-product credit for the Illinois coal is 50 mills/kWh more than for the PRB coal. The cost of $CO_2$ avoidance, for both the Illinois coal and the PRB coal, for an IGCC plant with Selexol process is the lowest among the three $CO_2$ capture processes examined in this study.

## Table 8.21. Cost of electricity generation of IGCC power plants

| | Illinois coal | | | PRB coal | | |
|---|---|---|---|---|---|---|
| | **268 MW** | **536 MW** | **1,071 MW** | **268 MW** | **536 MW** | **1,071 MW** |
| Capital cost, $/kW | | | | | | |
| Total plant cost | 1,967 | 1,610 | 1,318 | 2,196 | 1,795 | 1,467 |
| Total plant investment | 2,241 | 1,834 | 1,501 | 2,502 | 2,045 | 1,672 |
| Total capital requirement | 2,338 | 1,915 | 1,571 | 2,606 | 2,132 | 1,746 |
| | | | | | | |
| O&M costs, mills/kWh | | | | | | |
| Fixed O&M | 14.36 | 9.84 | 7.11 | 15.37 | 10.64 | 7.74 |
| Consumables | 1.09 | 1.09 | 1.09 | 1.04 | 1.04 | 1.04 |
| Fuel cost | 10.84 | 10.84 | 10.84 | 10.50 | 10.50 | 10.50 |
| Sulfur credit | 0.59 | 0.59 | 0.59 | 0.09 | 0.09 | 0.09 |
| | | | | | | |
| Annual carrying charge, mills/kWh | 52.61 | 43.10 | 35.35 | 58.65 | 47.98 | 39.29 |
| | | | | | | |
| Levelized cost | | | | | | |
| Electricity, mills/kWh | 78.32 | 64.29 | 53.80 | 85.47 | 70.06 | 58.48 |
| $CO_2$ avoided, $/t of $CO_2$ | 19.06 | 16.79 | 13.61 | 28.67 | 24.17 | 19.90 |
| $CO_2$ capture, $/t of $CO_2$ | 19.02 | 16.75 | 13.58 | 26.78 | 22.58 | 18.59 |

## 8.5. Cost Comparisons

### *8.5.1. General Comparisons*

Compared to the conventional PC power plant without $CO_2$ capture, the MEA unit and the oxy-combustion process significantly decreases the electricity generation efficiency (Figure 8.6). Installation of an MEA unit decreases the generation efficiency by about 28.2%. The oxy-combustion process decreases the generation efficiency by about 27.9%, just slightly lower than the MEA unit. The IGCC plant with $CO_2$ capture has a generation efficiency comparable to the PC plant without $CO_2$ capture. However, the IGCC plant without $CO_2$ capture could have a generation efficiency as high as 46%.

When a power generation plant is equipped with a $CO_2$ capture unit, the reduction of the net power output is largely due to energy use in the $CO_2$ capture process and the compression of $CO_2$ gas. Both the MEA unit and the oxy-combustion process consume about one third of the gross output of a power plant (Figure 8.7). The MEA unit and the ASU contribute to 50 to 60%, and $CO_2$ compression contributes to 20 to 25% of the total in-plant power use. In the IGCC plant the energy use of the Selexol unit is not significant because the bonding energy of $CO_2$ is not as strong with a physical absorbent as with a chemical absorbent. The IGCC plant also consumes much less oxygen than the oxy-combustion PC plant. In the IGCC plant, oxygen is only used for the gasification of coal rather than for combustion in the gas turbine.

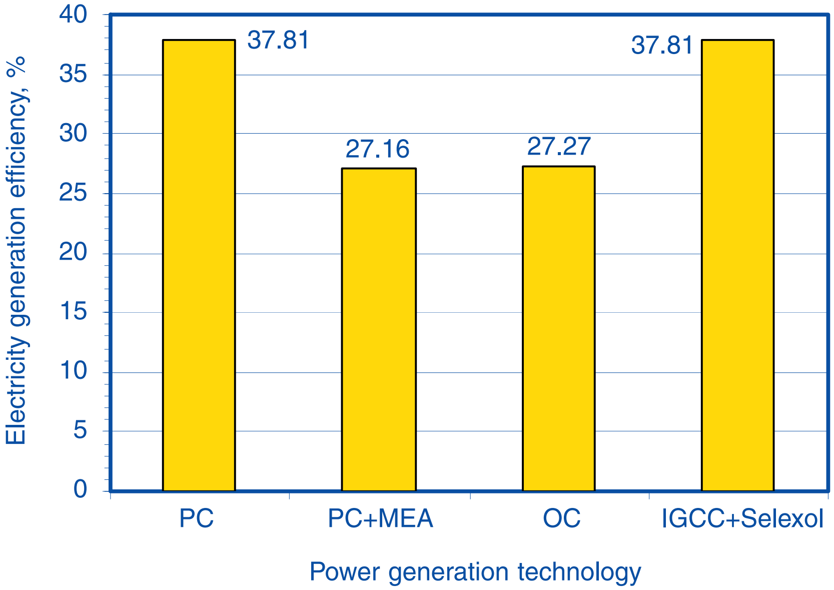


Figure 8.6. Net generation efficiency for 533-MW (gross) power plants (OC, oxy-combustion)

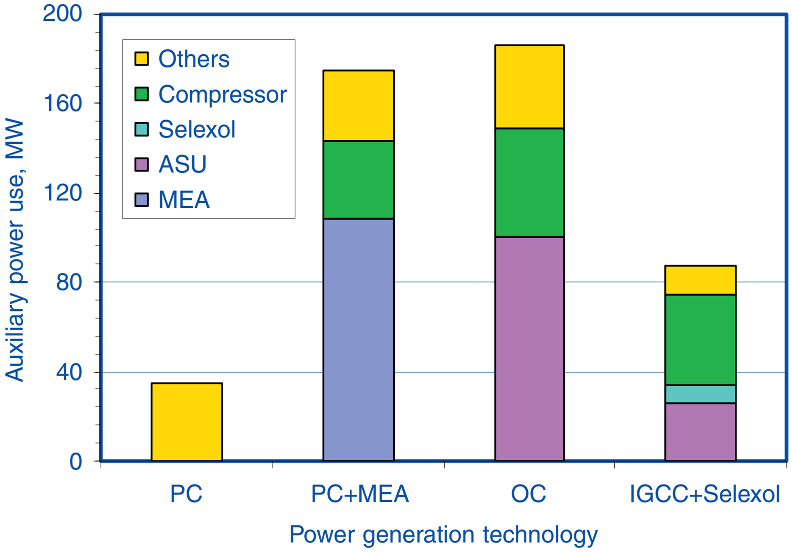


Figure 8.7. Auxiliary power uses for 533-MW (gross) power plants

The total plant cost for a PC plant without $CO_2$ capture is estimated to be $1,110 /kW (Figure 8.8). When the MEA unit is installed, capital costs per unit output of the plant increases by about 73%. The unit capital cost of an oxy-combustion process is about 64% higher than the reference PC plant without $CO_2$ capture. The increase is mainly due to the installations of the ASU and cryogenic equipment. The capital cost of the IGCC plant with a Selexol unit is about 40% higher than the reference PC plant. The cost of electricity increased by 77%, 56%, and 25% for the PC + MEA plant, oxy-combustion process, and the IGCC plant with Selexol process, respectively, compared to that of the reference plant (Figure 8.9). The increased cost of electricity for the IGCC plant is mainly due to the increase in capital cost. For the MEA and oxy-combustion processes, the increase is contributed to higher capital cost and an increase in fixed operating cost plus the expenses associated with MEA chemicals.

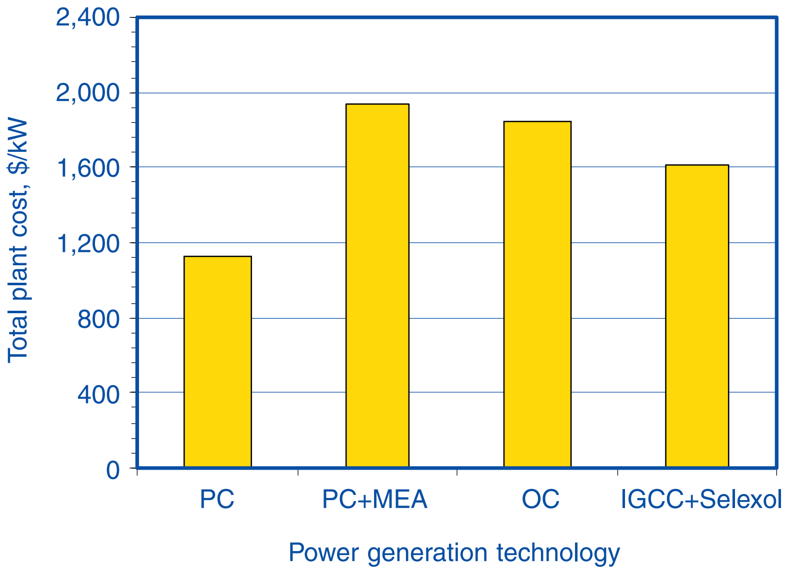


Figure 8.8. Capital costs for 533-MW (gross) power plants

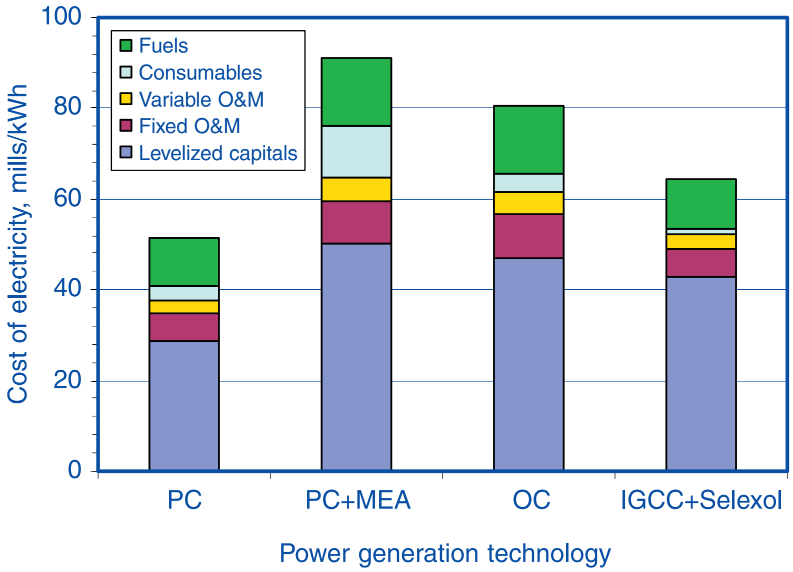


Figure 8.9. Costs of electricity for 533-MW (gross) power plants

The costs of $CO_2$ avoidance and $CO_2$ capture are based relative to the reference PC plant. For the PC + MEA power plant and the oxy-combustion process, the costs of $CO_2$ avoidance are much larger than the costs of $CO_2$ capture because the total amounts of the avoided $CO_2$ emissions are smaller than the captured $CO_2$ emissions. For the IGCC plant with Selexol process, the $CO_2$ capture and $CO_2$ avoidance are the same because $CO_2$ emissions per unit of electricity generation are comparable to those of the reference plant. The $CO_2$ avoidance costs for these three 533-MW (gross) power plants are $54/t, $36/t, and $17/t, respectively (Figure 8.10).

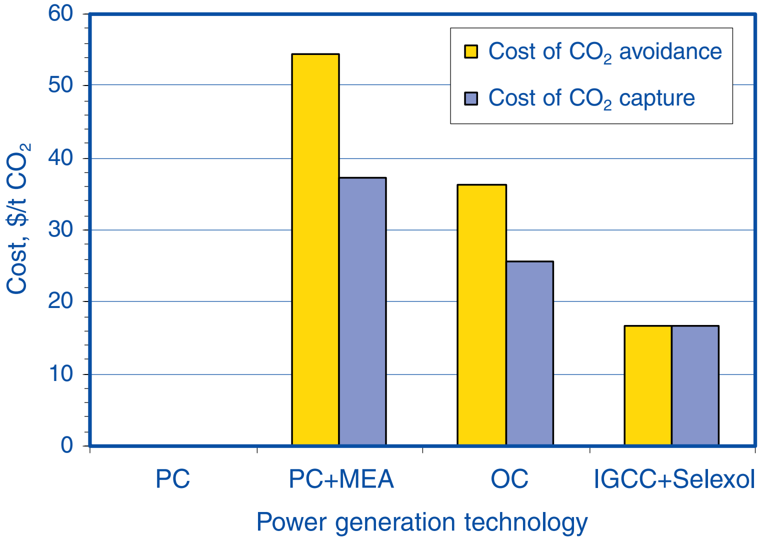


**Figure 8.10. Costs of $CO_2$ avoidance and capture for 533-MW (gross) power plants**

Three plant scales—266 MW, 533 MW, and 1,066 MW (gross)—were analyzed for all case studies. The costs of electricity increased as plant size decreased and increased more sharply as capacity decreased (Figure 8.11). The cost of electricity was well correlated to the plant size using a power rule with a factor of –0.28 to –0.26.

The cost of $CO_2$ avoidance and plant size (Figure 8.12) followed a similar trend. The power rule was also applied to correlate the costs to plant size. A power factor of –0.24 to –0.23 fit the data sets well for the three $CO_2$ capture processes studied.

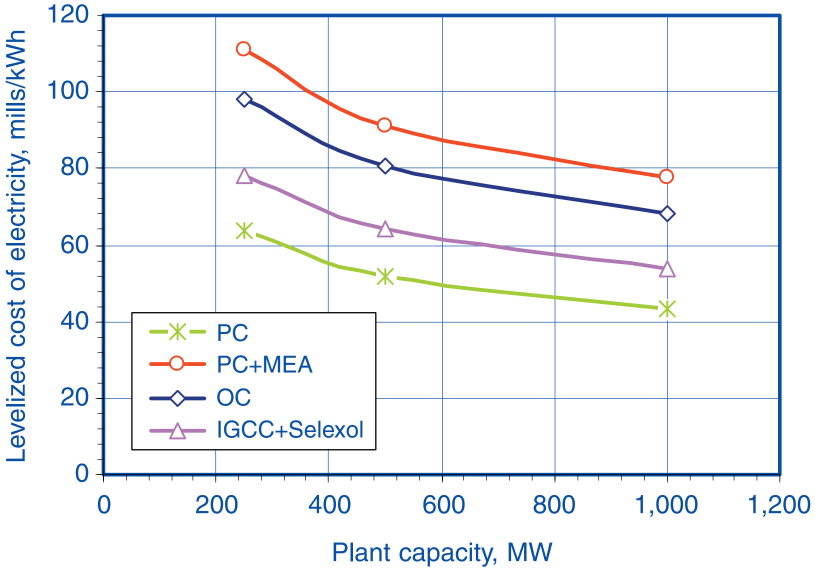


**Figure 8.11. Effect of plant size on electricity cost**

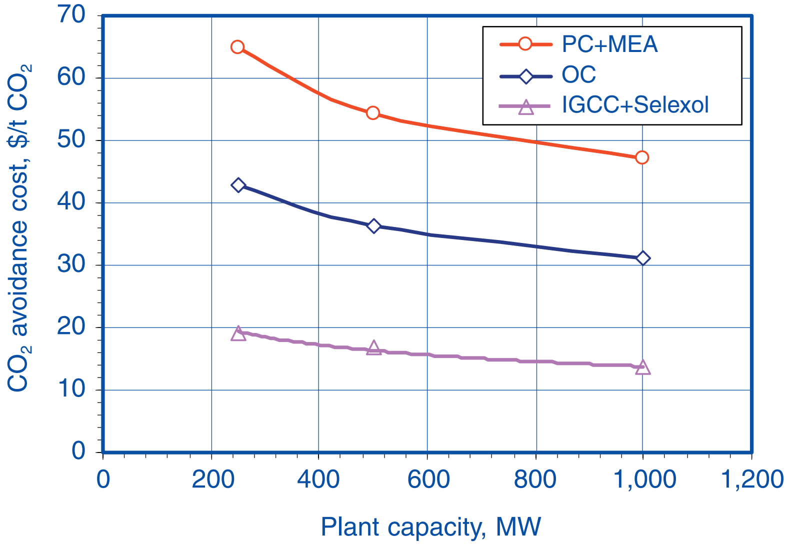


**Figure 8.12. Effect of plant size on $CO_2$ avoidance cost**

### *8.5.2. Economics of Illinois Coal and PRB Coal*

Additional techno-economic studies were conducted to compare the costs of electricity and $CO_2$ avoidance for power plants burning high-sulfur Illinois coal and PRB coal. Almost equal amounts of both coals are currently burned in power plants in the Illinois Basin. The reference PC plant burning the PRB coal has lower electricity costs than does the plant burning Illinois coal because no FGD installation was considered for the PC plant. However, for power plants that are equipped with the MEA units, the coal choice does not impact the cost of electricity. When an MEA unit is installed, an FGD unit is required to remove $SO_2$ emissions from the flue gas before it enters the MEA unit. Similarly, electricity costs for oxy-combustion plants burning Illinois coal and PRB coal were comparable. An IGCC plant + Selexol process burning Illinois high-sulfur is more attractive than burning PRB coal, mainly because sulfur in the former plant is recovered as a salable by-product (i.e., elemental sulfur) (Figure 8.13).

The calculated $CO_2$ avoidance costs favor burning high-sulfur coal rather than PRB coal in the three power generation plants studied. $CO_2$ avoidance cost is related to the cost of electricity. As shown in Figure 8.14, the $CO_2$ avoidance costs for the PRB coal are 7%, 12%, and 44% larger than the Illinois coal for the PC + MEA plant, the oxy-combustion process, and the IGCC plant, respectively.

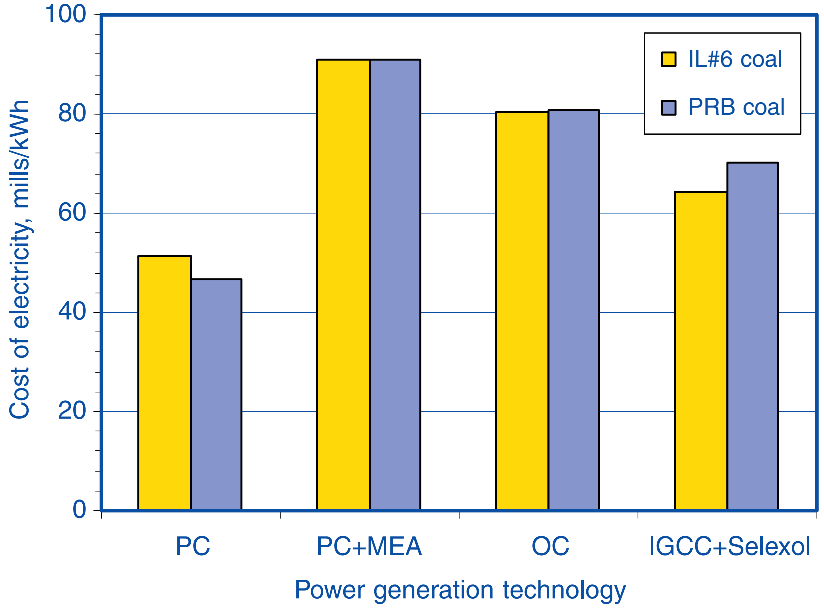


Figure 8.13. Costs of electricity for Illinois coal-fired and PRB coal-fired 533-MW plants

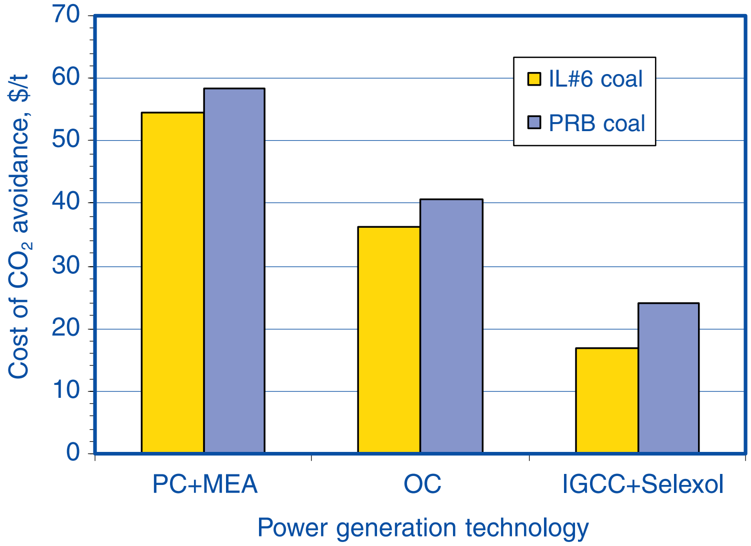


Figure 8.14. $CO_2$ avoidance costs for Illinois coal-fired and PRB coal-fired 533-MW plants

### 8.6. Process Optimization and Improvement

A sensitivity study was performed to evaluate the impact of absorption heat of MEA solvent used in chemical absorption process and energy consumption of the ASU in an oxy-combustion process on the cost of $CO_2$ avoidance. This type of analysis could provide some guide to scientists and engineers to improve existing, or develop advanced, $CO_2$ capture processes.

For the chemical absorption process, the sensitivity analysis was performed for the absorption heat of solvent. The sensitivity of absorption capacity of MEA was not studied because it would not greatly impact the process economics without improvement in absorption heat. The following parameters and assumptions were used in analysis:

1. Absorption heat of MEA reduces by 10%, 20%, 30%, 40%, and 50%.
2. Energy (or steam) requirement is proportional to absorption heat of solvent.
3. Absorption capacity of MEA remains the same as in the reference plant.
4. Capital cost of MEA does not change.
5. Reference plant is a 533-MW (gross) high-sulfur coal-fired power plant +MEA.

For the oxy-combustion process, the sensitivity analysis was performed considering the following parameters and assumptions:

1. Energy requirement of ASU reduces by 10%, 20%, 30%, 40%, 50%.
2. Capital cost of ASU does not change.
3. Reference plant is a 533 MW (gross) high-sulfur coal power plant.

The results of the sensitivity analysis (Figure 8.15) revealed that the $CO_2$ avoidance costs decreased for both cases studied. However, the $CO_2$ avoidance cost of the MEA process with 50% lower absorption heat requirement is still higher than the reference oxy-combustion power plant. The avoidance cost for the oxy-combustion power plant decreased by about 35% when energy requirement of the ASU is reduced by 50%. Additional sensitivity studies are required to evaluate the impact of chemical solvent absorption capacity, equipment size, cost of the solvent, and the size of power plant on the $CO_2$ avoidance cost.

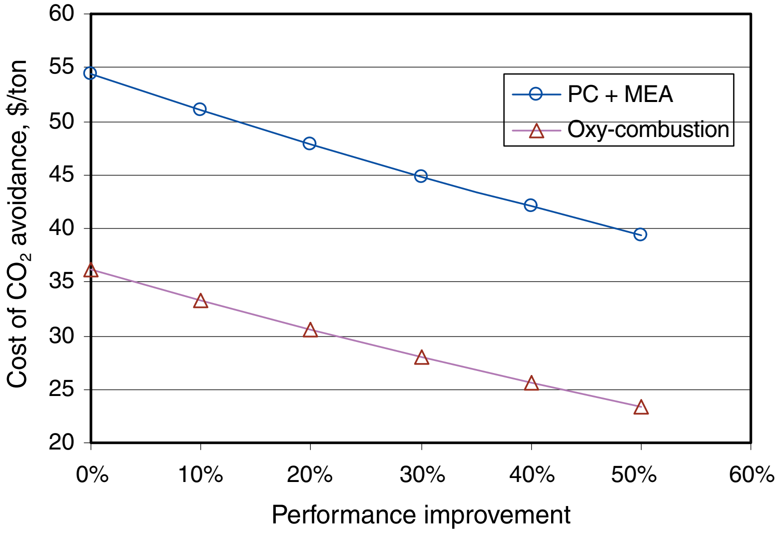


**Figure 8.15. The impact of process variables on $CO_2$ avoidance cost**

## 8.7. $CO_2$ Capture from Ethanol Plants

There are eight ethanol plants in the Illinois Basin generating about 3.7 million tons of $CO_2$ annually. Due to its high purity, much of the $CO_2$ is recovered and utilized for food and beverage industries. An average ethanol plant produces about 100 million gallons of ethanol per year. At this scale, the $CO_2$ emissions from an ethanol plant are comparable to the emissions from a 25-MW coal-fired power plant. But, because the concentration of the $CO_2$ is much higher in the product gas of an ethanol plant than in a coal combustion flue gas (87 vol% vs. 14 vol%), ethanol plants may offer a low-cost $CO_2$ source for sequestration research and testing, especially for small-scale field demonstration studies.

The concentration of $CO_2$ in the flue gas of ethanol plants is about 87% (120°F) on a wet basis and about 98% on a dry basis. Typical compositions of the $CO_2$ stream are listed in Table 8.22. Two major contaminants that exist in the gas stream are water and ethanol. Ethanol can be removed with water in a washing column operating at room temperature. The ethanol-water solution is then condensed, and high purity $CO_2$ is compressed to a desired pressure.

**Table 8.22. Flue gas composition from ethanol plants (at 120°F, 1 atm, $H_2O$ saturated)**

| Component | Concentration | Component | Concentration |
|---|---|---|---|
| $CO_2$ | 87.2% | Acetic acid | 10 PPMV |
| $H_2O$ | 11.1% | Amyl alcohol | 50 PPMV |
| Air | 1.2% | Isopropanol | 25 PPMV |
| Ethanol | 1,350 PPMV | Butanol | 25 PPMV |
| Methanol | 180 PPMV | Methane | 20 PPMV |
| Acetaldehyde | 270 PPMV | Ethyl acetate | 80 PPMV |
| Sulfur compounds ($H_2S$, $CS_2$) | 35 PPMV | | |

### *8.7.1. Overall Process Descriptions*

The flow chart used for process simulation using CHEMCAD is shown in Figure 8.16. Raw $CO_2$ stream from the ethanol plant enters a washing tower through a blower where ethanol and other water-soluble contaminants are dissolved into water. The washed gas stream is compressed and cooled. After removing the condensed water, the gas stream is compressed to 1,500 psi for transportation.

### *8.7.2. Cost of $CO_2$ Recovery from Ethanol Plants*

The cost of $CO_2$ recovery from ethanol plants was analyzed based on the following assumptions:

Scale: 800 tons of $CO_2$ per day
Life of the plant: 30 years
Product $CO_2$ pressure: 1,200 psi
Product $CO_2$ temperature: $68^0$F
Capacity usage: 85%

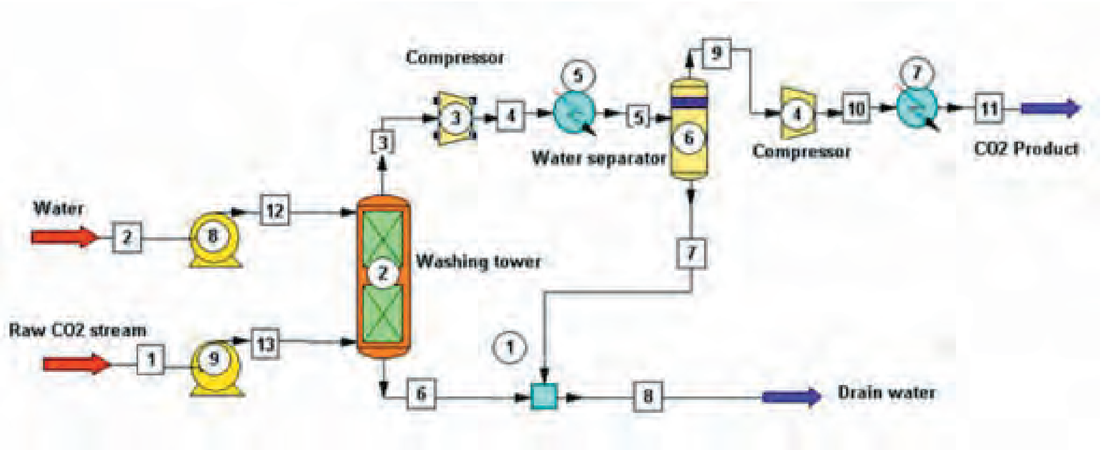


**Figure 8.16. Flow chart for $CO_2$ capture from ethanol plants**

The financial terms used are similar to those listed in Appendix B1 for the power generation plants. The equipment cost data were obtained using CHEMCAD, and costs for other items such as installation, piping, instrumentation, building, and land were chosen from literature values (Peter and Timmerhaus, 1991). The total cost for purchased equipment is about \$3.225 million, and the total plant cost is about \$7.1 million. The major equipment includes two compressors, and the major operating cost is associated with the compression of $CO_2$. The results of the cost analysis are presented in Table 8.23.

**Table 8.23. Cost analysis of $CO_2$ recovery from ethanol plants**

| **Item** | **Unit cost (\$1,000)** | **Units** | **Total cost (\$1,000)** |
|---|---|---|---|
| Purchased equipment cost | | | |
| Compressor | 1,200 | 2 | 2,400 |
| Heat exchangers | 150 | 2 | 300 |
| Washing tower | 350 | 1 | 350 |
| Water condenser | 25 | 1 | 25 |
| Fan | 150 | 1 | 150 |
| Total purchased equipment cost (TPEC) | | | 3,225 |
| Installation, % of TPEC | | 30% of TPEC | 968 |
| Piping, % of TPEC | | 30% of TPEC | 968 |
| Instrumentation, % of TPEC | | 20% of TPEC | 806 |
| Electrical, % of TPEC | | 10% of TPEC | 323 |
| Building, % of TPEC | | 20% of TPEC | 645 |
| Land, % of TPEC | | 10% of TPEC | 323 |
| Total plant cost (TPC) | | | 7,095 |
| Levelized capital cost | | 0.138 × 7,095 | 979 |
| O&M cost | | | |
| Fixed O&M | | | |
| Maintenance | 3% of TPC | | 213 |
| Admin. & support | | | 236 |
| Labor cost | 40 × 8,760 | 2 | 701 |
| Consumable | | | |
| Water | $\$0.06/\mathrm{m}^3$ | | 60 |
| Electricity | \$50/MWh | 3.166 MW | 1,266 |
| Total O&M cost | | | 2,476 |
| $CO_2$ capture cost, \$/metric ton | | | 13.92 |

The levelized $CO_2$ recovery cost is \$13.92/t. Since the electricity consumption for $CO_2$ compression is included in $CO_2$ recovery process cost, the recovery cost here equals the $CO_2$ avoidance cost. This cost is much lower than that of capturing of $CO_2$ from a coal-fired power plant + MEA, which is about \$47/t to \$67/t, depending on the plant size and type of coal burned (see Table 8.9).

## 9. Conclusions

1. There are 122 utility power plants in the Illinois Basin. They emit about 90% of 276 million t of $CO_2$ from stationary sources in the Illinois Basin annually. Power plant $CO_2$ emissions in the Illinois Basin contribute about 11.4% of total U.S. $CO_2$ emissions from electric power generation plants. The four largest power plants emit 22% of the total utility $CO_2$, the 13 largest plants emit 50% of total utility $CO_2$, and the 29 largest emit about 80% of the total utility $CO_2$.

2. Non-utility sources contribute to about 10% of total $CO_2$ emissions in the Illinois Basin. These sources include 5 refineries, 23 iron and steel plants, 8 cement plants, 1 lime plant, 2 aluminum plants, 1 ammonia plant, and 8 ethanol plants.

3. The most technically and economically viable commercial technology currently available for separating and capturing $CO_2$ from power plants is a post-combustion absorption process that uses MEA as a chemical solvent. A substantial increase in electricity cost results when an MEA process is installed to capture power plant $CO_2$. The results from a techno-economic study conducted revealed that for a 533-MW (gross) Illinois coal-fired plant, the cost of electricity is 91 mills/kWh for the MEA + PC plant and 52 mills/kWh for the conventional PC plant without $CO_2$ capture.

4. Burning coal in an oxygen-enriched flue gas is a potentially attractive technology for producing high-purity $CO_2$ flue gas for sequestration. The cost of electricity was estimated to be 81 mills/kWh for a 533-MW (gross) oxy-combustion power plant.

5. The cost of $CO_2$ avoidance was \$54/t for the MEA + PC plant and \$36/t for the 533-MW (gross) oxy-combustion plant.

6. The MEA process and the operation of the ASU in an oxy-combustion power plant are energy-intensive. Each process consumes about 20% of the gross electricity of a PC plant. Technological advances that result in lowering heat duty requirement during the regeneration of MEA solvent and electricity consumption in the ASU would substantially benefit the economics of $CO_2$ capture and separation by chemical absorption and oxy-combustion processes.

7. The cost of electricity for an IGCC + physical absorption (Selexol) plant is lower that a PC + MEA plant. For a 533-MW Illinois coal-fired IGCC + Selexol plant, the cost of electricity and of $CO_2$ avoidance are 64 mills/kWh and \$17/t, respectively.

8. The cost of electricity for the PC + MEA plant and the oxy-combustion plant was determined to be independent of the type of coal used (Illinois No. 6 coal or PRB coal). The $CO_2$ avoidance cost for the Illinois coal was slightly lower than that of the PRB coal. However, an IGCC plant + Selexol process burning Illinois high-sulfur coal was more attractive than burning PRB coal if the elemental sulfur by-product is considered a salable product.

9. The costs of electricity increase as plant size decreases from about 250 MW to around 1,070 MW (gross) in a power rule with an exponent of –0.28 to –0.26 for the examined power generation systems. The dependence of $CO_2$ avoidance cost on the plant size also fit the power rule, with an exponent of –0.24 to –0.23.

10. Ethanol plants are excellent sources of high-purity $CO_2$ for small-scale field demonstration sequestration tests in Illinois Basin. The cost of $CO_2$ capture from an ethanol plant with a production capacity of 100 million gallons/year was estimated to be about \$13.92/t.

11. The results of a sensitivity analysis revealed the impact on $CO_2$ avoidance cost of the heat of absorption of the solvent in an MEA plant and the energy consumption of the ASU plant in an oxy-

coal combustion plant. A 50% reduction in heat of absorption and 50% lower energy consumption in the ASU resulted in 27% and 37% reduction in $CO_2$ avoidance costs for the two processes, respectively.

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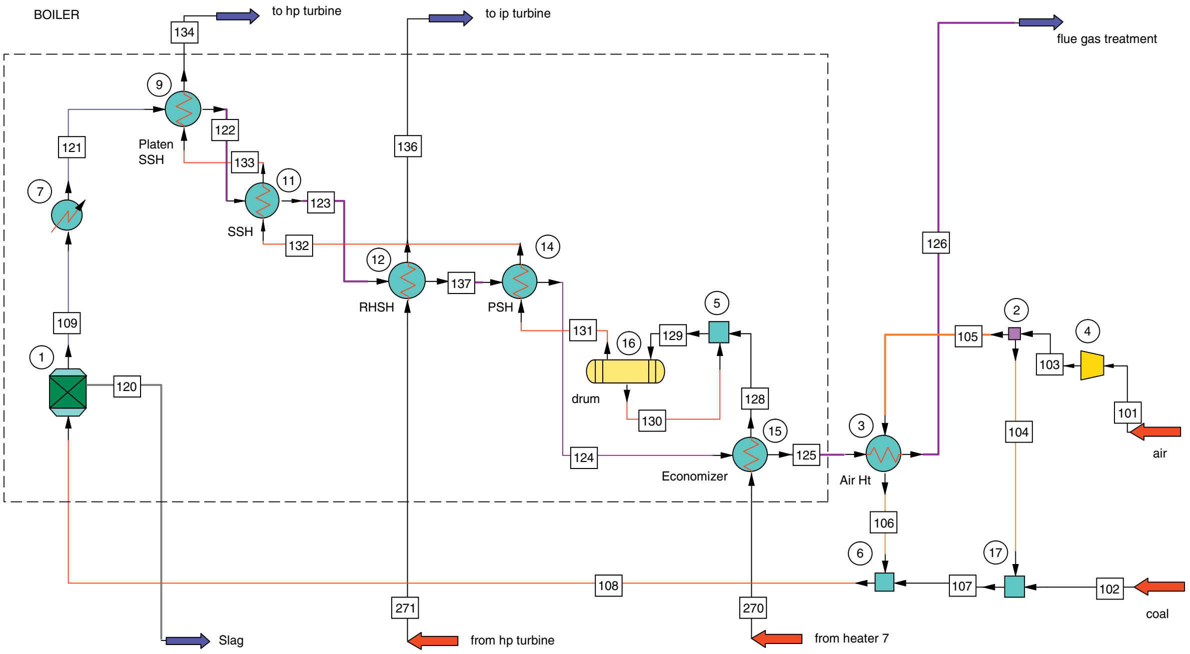
# Appendix A1. Reference Air-blown PC Power Plants with 533-MW (Gross) Output

## Figure A1.1. Boiler



## Table A1.1. Illinois coal

| Stream no. | 101 | 102 | 103 | 104 |
|---|---|---|---|---|
| Temp., °F | 63.0000 | 63.0000 | 81.6783 | 81.6783 |
| Pres., psia | 4.4000 | 15.8000 | 15.8000 | 15.8000 |
| Enth., mm BTU/hr | –13.910 | –432.10 | 4.0910 | 0.20455 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 | 1.0000 |
| Total, lb/hr | 3,991,198.00 | 360,610.00 | 3,991,198.00 | 199,560.00 |

| Stream no. | 105 | 106 | 107 | 108 |
|---|---|---|---|---|
| Temp., °F | 81.6783 | 510.0580 | 52.9506 | 457.8348 |
| Pres., psia | 15.8000 | 15.8000 | 15.8000 | 15.8000 |
| Enth., mm BTU/hr | 3.8865 | 401.70 | –431.90 | –30.193 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.38092 | 0.92903 |
| Total, lb/hr | 3,791,638.00 | 3,791,638.00 | 560,170.00 | 4,351,798.00 |

| Stream no. | 109 | 120 | 121 | 122 |
|---|---|---|---|---|
| Temp.,°F | 2,400.0000 | 2,400.0000 | 2,285.7006 | 1,900.0000 |
| Pres., psia | 15.3000 | 15.3000 | 15.3000 | 15.1000 |
| Enth., mm BTU/hr | –1,748.3 | –39.884 | –1,904.3 | –2,422.9 |
| Vapor mass fraction | 1.0000 | 0.00000 | 1.0000 | 1.0000 |
| Total, lb/h | 4,319,700.00 | 32,094.00 | 4,319,700.00 | 4,319,700.00 |

| Stream no. | 123 | 124 | 125 | 126 |
|---|---|---|---|---|
| Temp., °F | 1,700.0000 | 900.0000 | 650.0000 | 295.0000 |
| Pres., psia | 14.9000 | 14.5000 | 14.3000 | 14.1000 |
| Enth., mm BTU/hr | –2,686.3 | –3,689.5 | –3,982.3 | –4,380.1 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 4,319,700.00 | 4,319,700.00 | 4,319,700.00 | 4,319,700.00 |

| Stream no. | 128 | 129 | 130 | 131 |
|---|---|---|---|---|
| Temp.,°F | 557.8544 | 557.8544 | 665.1114 | 665.1114 |
| Pres., psia | 2,800.0000 | 2,450.0000 | 2,450.0000 | 2,450.0000 |
| Enth., mm BTU/hr | –21,590 | –21,590 | 0.00000 | –19,676 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | .0000 |
| Total, lb/h | 3,422,824.00 | 3,422,824.00 | 0.00 | 3,422,824.00 |

| Stream no. | 32 | 133 | 134 | 136 |
|---|---|---|---|---|
| Temp.,°F | 731.8905 | 802.0647 | 1,000.7155 | 1,000.0000 |
| Pres., psia | 2,440.0000 | 2,430.0000 | 2,414.7000 | 545.4000 |
| Enth., mm BTU/hr | –19,298 | –19,034 | –18,516 | –16,177 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 3,422,824.00 | 3,422,824.00 | 3,422,824.00 | 3,022,124.00 |

| Stream no. | 137 | 270 | 271 |
|---|---|---|---|
| Temp., °F | 1,210.4062 | 488.9000 | 634.5002 |
| Pres., psia | 14.7000 | 2,800.0000 | 603.6000 |
| Enth., mm BTU/hr | –3,311.3 | –21,883 | –16,802 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 |
| Total, lb/h | 4,319,700.00 | 3,422,824.00 | 3,022,124.00 |
| | | | |
| IL6# coal–new1 | 0.00 | 0.00 | 0.00 |
| Nitrogen dioxide | 7.90 | 0.00 | 0.00 |

## Table A1.2. PRB coal

| Stream no. | 101 | 102 | 103 | 104 |
|---|---|---|---|---|
| Temp., °F | 63.0000 | 63.0000 | 81.6783 | 81.6783 |
| Pres., psia | 14.4000 | 15.8000 | 15.8000 | 15.8000 |
| Enth., mm BTU/hr | –14.043 | –1,378.5 | 4.1302 | 0.20651 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 | 1.0000 |
| Total, lb/hr | 4,029,474.00 | 524,982.00 | 4,029,474.00 | 201,473.70 |

| Stream no. | 105 | 106 | 107 | 108 |
|---|---|---|---|---|
| Temp., °F | 81.6783 | 536.3796 | 57.3598 | 345.5448 |
| Pres., psia | 15.8000 | 15.8000 | 15.8000 | 15.8000 |
| Enth., mm BTU/hr | 3.9237 | 430.74 | –1,378.3 | –947.51 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.29099 | 0.92109 |
| Total, lb/hr | 3,828,000.25 | 3,828,000.25 | 726,456.00 | 4,554,397.50 |

| Stream no. | 109 | 120 | 121 | 122 |
|---|---|---|---|---|
| Temp., °F | 2,400.0000 | 2,400.0000 | 2,191.6589 | 1,900.0000 |
| Pres., psia | 15.3000 | 15.3000 | 15.3000 | 15.1000 |
| Enth., mm BTU/hr | –2,498.4 | –34.252 | –2,804.7 | –3,226.5 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 | 1.0000 |
| Total,lb/hr | 4,526,829.50 | 27,562.00 | 4,526,829.50 | 4,526,829.50 |

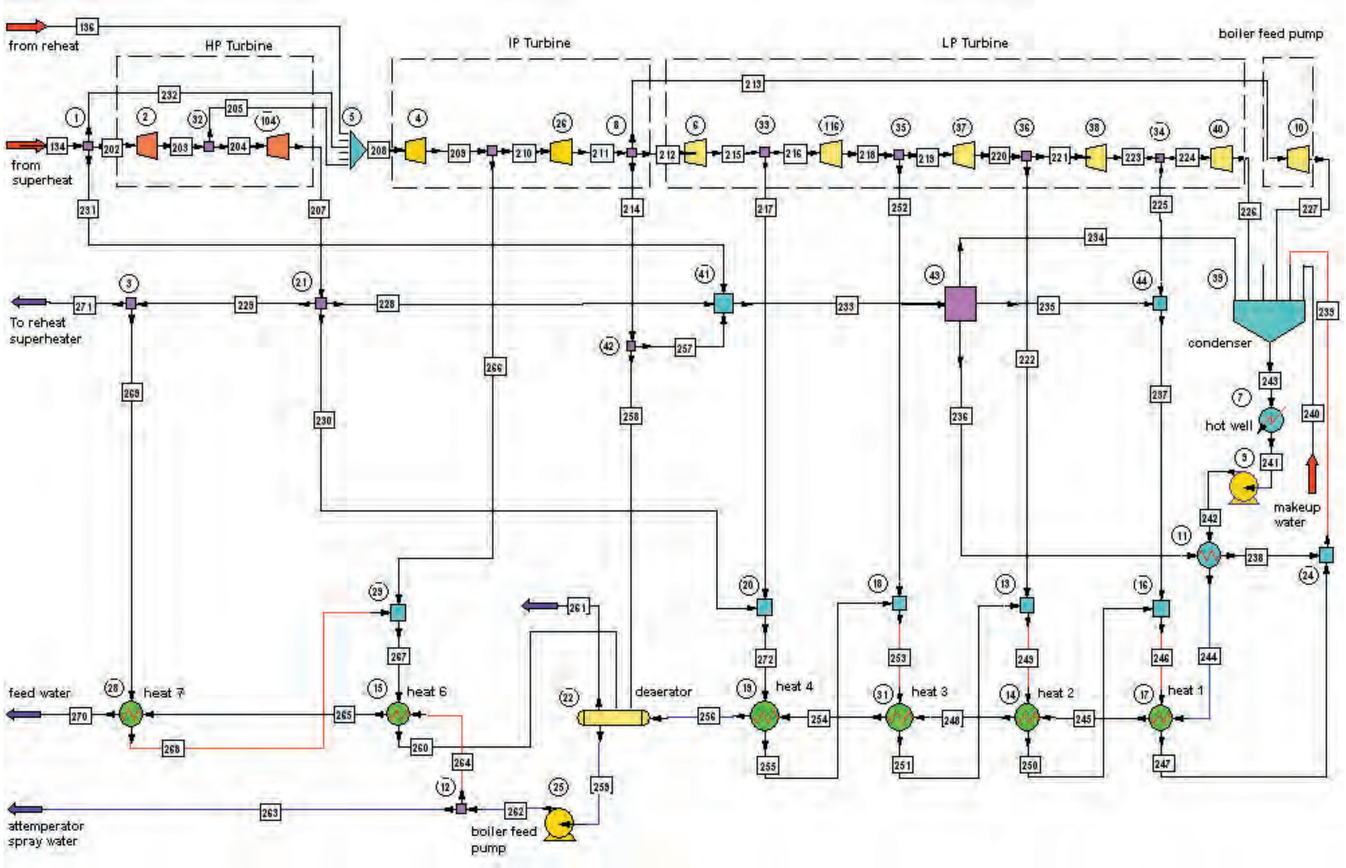
| Stream no. | 123 | 124 | 125 | 126 |
|---|---|---|---|---|
| Temp., °F | 1700.0000 | 900.0000 | 650.0000 | 295.0000 |
| Pres., psia | 14.9000 | 14.5000 | 14.3000 | 14.1000 |
| Enth., mm BTU/hr | –3,510.3 | –4,588.9 | –4,903.2 | –5,330.0 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 4,526,829.50 | 4,526,829.50 | 4,526,829.50 | 4,526,829.50 |

| Stream no. | 128 | 129 | 130 | 131 |
|---|---|---|---|---|
| Temp., °F | 562.6944 | 562.6944 | 665.1114 | 665.1114 |
| Pres., psia | 2,800.0000 | 2,450.0000 | 2,450.0000 | 2,450.0000 |
| Enth., mm BTU/hr | –21,568 | –21,568 | 0.00000 | –19,676 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 1.0000 |
| Total, lb/hr | 3,422,824.00 | 3,422,824.00 | 0.00 | 3,422,824.00 |

| Stream no. | 132 | 133 | 134 | 136 |
|---|---|---|---|---|
| Temp., °F | 749.9162 | 833.4616 | 1,000.2580 | 1,000.0000 |
| Pres., psia | 2,440.0000 | 2,430.0000 | 2,414.7000 | 545.4000 |
| Enth., mm BTU/hr | –19,222 | –18,938 | –18,517 | –16,177 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 3,422,824.00 | 3,422,824.00 | 3,422,824.00 | 3,022,124.00 |

| Stream no. | 137 | 270 | 271 |
|---|---|---|---|
| Temp., °F | 1,245.7663 | 488.9000 | 634.5002 |
| Pres., psia | 14.7000 | 2,800.0000 | 603.6000 |
| Enth., mm BTU/hr | –4,135.2 | –21,883 | –16,802 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 |
| Total, lb/hr | 4,526,829.50 | 3,422,824.00 | 3,022,124.00 |

## Figure A1.2. Steam cycle (same for Illinois coal and PRB coal)



## Table A1.3. Illinois coal and PRB coal

| Stream no. | 134 | 136 | 202 | 203 |
|---|---|---|---|---|
| Temp., °F | 1,000.0000 | 1,000.0000 | 1,000.0000 | 908.1930 |
| Pres., psia | 2,414.7000 | 545.4000 | 2,414.7000 | 1,775.0000 |
| Enth., mm BTU/hr | 5,000.4 | 4,587.5 | 4,991.5 | 4,863.3 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 3,422,824.00 | 3,022,125.00 | 3,416,724.50 | 3,416,724.50 |

| Stream no. | 204 | 205 | 207 | 208 |
|---|---|---|---|---|
| Temp., °F | 908.1930 | 908.1930 | 631.1807 | 997.5285 |
| Pres., psia | 1,775.0000 | 1,775.0000 | 603.6000 | 545.4000 |
| Enth., mm BTU/hr | 4,806.0 | 57.392 | 4,420.2 | 4,651.9 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 3,376,403.75 | 40,320.77 | 3,376,403.75 | 3,067,190.00 |

| Stream no. | 209 | 210 | 211 | 212 |
|---|---|---|---|---|
| Temp., °F | 817.2190 | 817.2190 | 702.8816 | 702.8816 |
| Pres., psia | 278.9000 | 278.9000 | 174.9000 | 174.9000 |
| Enth., mm BTU/hr | 4,386.6 | 4,240.6 | 4,080.4 | 3,626.8 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 3,067,190.00 | 2,965,172.00 | 2,965,172.00 | 2,635,575.50 |

| Stream no. | 213 | 214 | 215 | 216 |
|---|---|---|---|---|
| Temp., °F | 702.8816 | 702.8816 | 494.9297 | 494.9297 |
| Pres., psia | 174.9000 | 174.9000 | 66.5000 | 66.5000 |
| Enth., mm BTU/hr | 171.69 | 281.87 | 3,372.3 | 3,171.7 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 124,762.58 | 204,834.09 | 2,635,575.50 | 2,478,811.50 |

| Stream no. | 217 | 218 | 219 | 220 |
|---|---|---|---|---|
| Temp., °F | 494.9297 | 312.4631 | 312.4631 | 216.4566 |
| Pres., psia | 66.5000 | 24.0000 | 24.0000 | 12.8000 |
| Enth., mm BTU/hr | 200.58 | 2,965.7 | 2,864.0 | 2,760.2 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 156,764.03 | 2,478,811.50 | 2,393,808.00 | 2,393,808.00 |

| Stream no. | 221 | 222 | 223 | 224 |
|---|---|---|---|---|
| Temp., °F | 216.4566 | 216.4566 | 172.1188 | 172.1188 |
| Pres., psia | 12.8000 | 12.8000 | 6.3000 | 6.3000 |
| Enth., mm BTU/hr | 2,667.4 | 92.755 | 2,566.2 | 2,397.0 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.97448 | 0.97448 |
| Total, lb/hr | 2,313,364.25 | 80,443.91 | 2,313,364.25 | 2,160,797.75 |

| Stream no. | 225 | 226 | 227 | 228 |
|---|---|---|---|---|
| Temp., °F | 172.1188 | 101.5784 | 104.7907 | 631.1807 |
| Pres., psia | 6.3000 | 1.0000 | 1.1000 | 603.6000 |
| Enth., mm BTU/hr | 169.24 | 2,182.6 | 128.88 | 16.235 |
| Vapor, mass fraction | 0.97448 | 0.90786 | 0.92852 | 1.0000 |
| Total, lb/hr | 152,566.38 | 2,160,797.75 | 124,762.58 | 12,401.53 |

| Stream no. | 229 | 230 | 231 | 232 |
|---|---|---|---|---|
| Temp., °F | 631.1807 | 631.1807 | 1,000.0000 | 1,000.0000 |
| Pres., psia | 603.6000 | 603.6000 | 2,414.7000 | 2,414.7000 |
| Enth., mm BTU/hr | 4,395.0 | 8.9243 | 1.9801 | 6.9305 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 3,357,185.25 | 6,816.96 | 1,355.44 | 4,744.03 |

| Stream no. | 233 | 234 | 235 | 236 |
|---|---|---|---|---|
| Temp., °F | 626.4604 | 626.4604 | 626.4604 | 626.4604 |
| Pres., psia | 174.9000 | 174.9000 | 174.9000 | 174.9000 |
| Enth., mm BTU/hr | 24.561 | 4.0125 | 15.866 | 4.6827 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 18,368.40 | 3,000.79 | 11,865.62 | 3,501.99 |

| Stream no. | 237 | 238 | 239 | 240 |
|---|---|---|---|---|
| Temp., °F | 172.1188 | 200.0000 | 118.1270 | 70.0000 |
| Pres., psia | 6.3000 | 174.9000 | 6.0000 | 14.7000 |
| Enth., mm BTU/hr | 185.11 | 0.58921 | 42.801 | 0.62795 |
| Vapor, mass fraction | 0.99101 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 164,432.00 | 3,501.99 | 496,962.22 | 16,527.00 |

| Stream no. | 241 | 242 | 243 | 244 |
|---|---|---|---|---|
| Temp., °F | 96.4000 | 97.5560 | 98.0755 | 99.0167 |
| Pres., psia | 0.9000 | 330.0000 | 0.9000 | 330.0000 |
| Enth., mm BTU/hr | 180.43 | 183.67 | 2,358.9 | 187.77 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.74740 | 0.00000 |
| Total, lb/hr | 2,802,050.50 | 2,802,050.50 | 2,802,050.50 | 2,802,050.50 |

| Stream no. | 245 | 246 | 247 | 248 |
|---|---|---|---|---|
| Temp., °F | 167.4000 | 169.9850 | 117.5445 | 198.7000 |
| Pres., psia | 300.0000 | 6.0000 | 6.0000 | 270.0000 |
| Enth., mm BTU/hr | 379.63 | 234.07 | 42.212 | 467.78 |
| Vapor, mass fraction | 0.00000 | 0.33771 | 0.00000 | 0.00000 |
| Total, lb/hr | 2,802,050.50 | 493,460.25 | 493,460.25 | 2,802,050.50 |

| Stream no. | 249 | 250 | 251 | 252 |
|---|---|---|---|---|
| Temp., °F | 201.5301 | 180.6844 | 210.1221 | 312.4631 |
| Pres., psia | 11.9000 | 11.9000 | 22.4000 | 24.0000 |
| Enth., mm BTU/hr | 137.12 | 48.966 | 44.361 | 101.70 |
| Vapor, mass fraction | 0.25293 | 0.00000 | 0.00000 | 1.0000 |
| Total, lb/hr | 329,028.28 | 329,028.28 | 248,584.39 | 85,003.41 |

| Stream no. | 253 | 254 | 255 | 256 |
|---|---|---|---|---|
| Temp., °F | 234.0796 | 231.1000 | 241.2455 | 292.3000 |
| Pres., psia | 22.4000 | 240.0000 | 62.4000 | 160.0000 |
| Enth., mm BTU/hr | 136.04 | 559.46 | 34.344 | 734.62 |
| Vapor, mass fraction | 0.36067 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 248,584.39 | 2,802,050.50 | 163,581.00 | 2,802,050.50 |

| Stream no. | 257 | 258 | 259 | 260 |
|---|---|---|---|---|
| Temp., °F | 702.8816 | 702.8816 | 366.0413 | 399.4454 |
| Pres., psia | 174.9000 | 174.9000 | 164.8000 | 263.8000 |
| Enth., mm BTU/hr | 6.3458 | 275.53 | 1,167.9 | 164.23 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.00000 | 0.00000 |
| Total, lb/hr | 4,611.43 | 200,222.67 | 3,439,352.50 | 437,079.31 |

| Stream no. | 261 | 262 | 263 | 264 |
|---|---|---|---|---|
| Temp., °F | 366.0413 | 376.1441 | 376.1441 | 376.1441 |
| Pres., psia | 164.8000 | 2,903.3000 | 2,903.3000 | 2,903.3000 |
| Enth., mm BTU/hr | 0.00000 | 1,205.2 | 41.393 | 1,163.8 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 0.00 | 3,439,352.50 | 118,124.56 | 3,321,228.00 |

| Stream no. | 265 | 266 | 267 | 268 |
|---|---|---|---|---|
| Temp., °F | 407.5000 | 817.2190 | 410.8008 | 415.5000 |
| Pres., psia | 2,853.3000 | 278.9000 | 278.8000 | 588.5000 |
| Enth., mm BTU/hr | 1277.3 | 145.90 | 277.73 | 131.83 |
| Vapor, mass fraction | 0.00000 | 1.0000 | 0.30377 | 0.00000 |
| Total, lb/hr | 3,321,228.00 | 102,017.80 | 437,079.31 | 335,061.50 |

| Stream no. | 269 | 270 | 271 | 272 |
|---|---|---|---|---|
| Temp., °F | 631.1807 | 488.3002 | 631.1807 | 496.6450 |
| Pres., psia | 603.6000 | 2,800.0000 | 603.6000 | 62.4000 |
| Enth., mm BTU/hr | 438.64 | 1584.1 | 3956.4 | 209.51 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 | 1.0000 |
| Total, lb/hr | 335,061.50 | 3,321,228.00 | 3,022,123.75 | 163,581.00 |

**Figure A1.3. Flue gas cleaning (Illinois coal)**

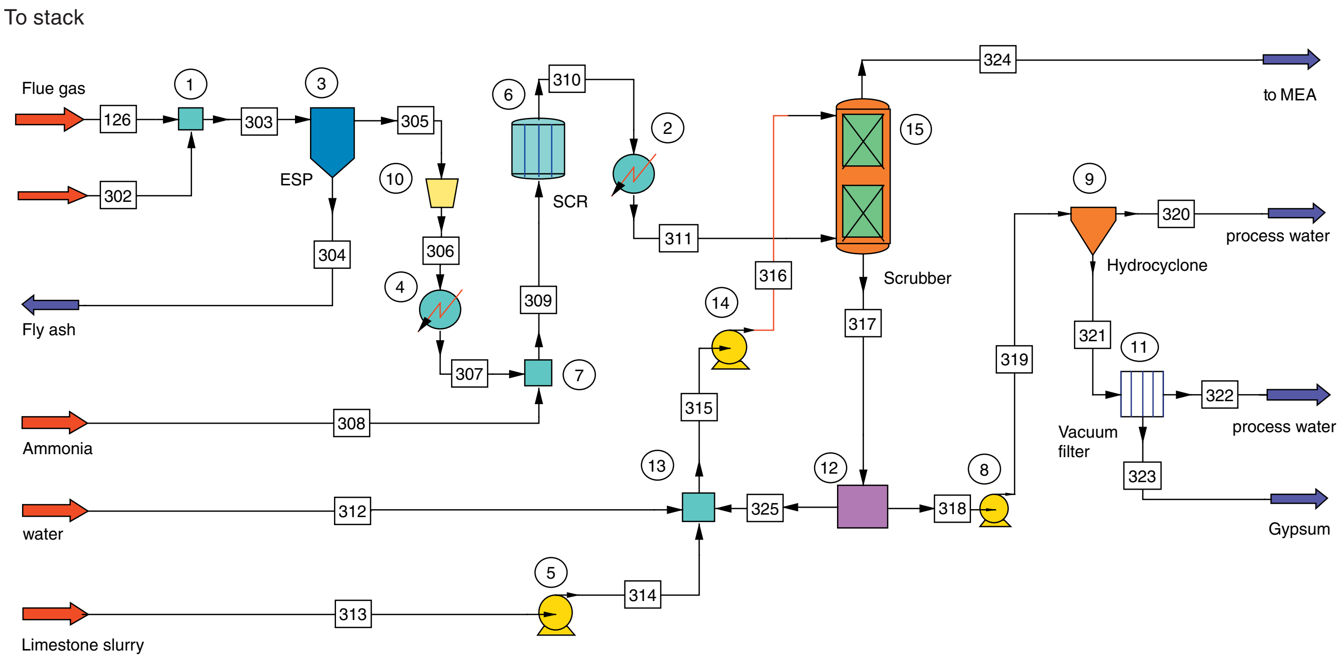


**Table A1.4. Illinois coal**

| Stream no. | 126 | 302 | 303 | 304 |
|---|---|---|---|---|
| Temp., °F | 295.0000 | 295.0000 | 295.0003 | 295.0003 |
| Pres., psia | 14.2000 | 14.2000 | 14.2000 | 14.1000 |
| Enth., mm BTU/hr | –4,374.9 | –54.992 | –4,429.9 | –54.440 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 | 0.00000 |
| Total, lb/hr | 4,319,700.00 | 25,675.20 | 4,345,375.00 | 25,417.41 |

| Stream no. | 305 | 306 | 307 | 308 |
|---|---|---|---|---|
| Temp., °F | 295.0003 | 374.6118 | 752.0000 | 70.0000 |
| Pres., psia | 14.1000 | 18.0000 | 18.0000 | 15.1000 |
| Enth., mm BTU/hr | –4,375.3 | –4,291.8 | –3,863.5 | –0.62266 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/h | 4,319,958.00 | 4,319,958.00 | 4,319,958.00 | 535.74 |

| Stream no. | 309 | 310 | 311 | 312 |
|---|---|---|---|---|
| Temp., °F | 749.3875 | 752.0000 | 295.0000 | 70.0000 |
| Pres., psia | 15.1000 | 17.8000 | 17.8000 | 60.0000 |
| Enth., mm BTU/hr | –3,864.1 | –3,869.2 | –4,385.2 | –1,145.8 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 0.00000 |
| Total, lb/hr | 4,320,493.50 | 4,320,493.50 | 4,320,493.50 | 167,803.00 |

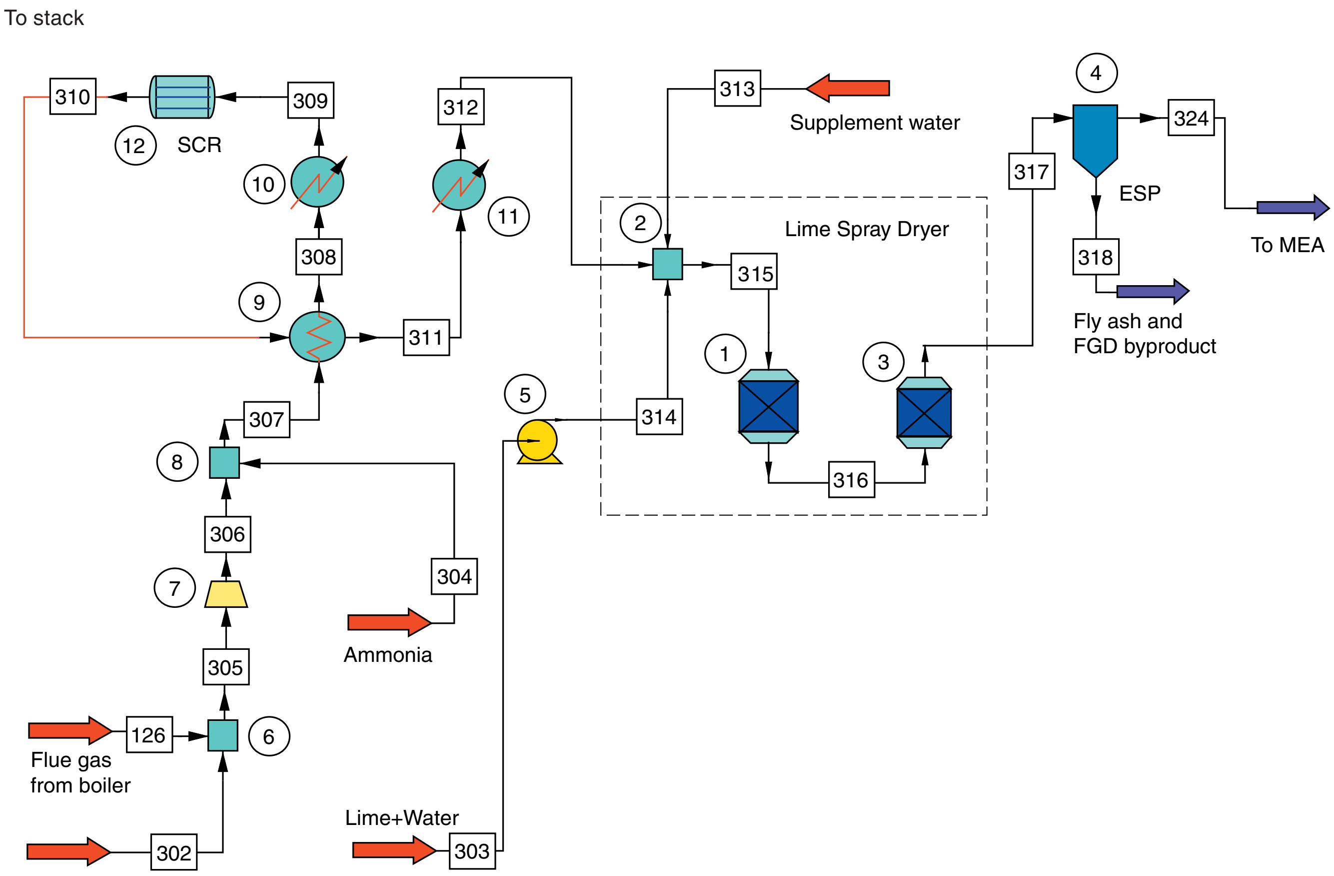
| Stream no. | 313 | 314 | 315 | 316 |
|---|---|---|---|---|
| Temp., °F | 70.0000 | 70.1574 | 129.1991 | 129.2440 |
| Pres., psia | 14.7000 | 60.0000 | 17.4000 | 30.0000 |
| Enth., mm BTU/hr | –1,177.2 | –1,177.2 | –6.1073E+005 | –6.1072E+005 |
| Ionic strength, molal | 0.0005 | 0.0005 | 0.0572 | 0.0572 |
| Total, lb/h | 179,427.23 | 179,427.25 | 96,026,952.00 | 96,026,952.00 |

| Stream no. | 317 | 318 | 319 | 320 |
|---|---|---|---|---|
| Temp., °F | 129.7292 | 129.7160 | 129.7508 | 129.7508 |
| Pres., psia | 17.4000 | 17.4000 | 30.0000 | 20.3720 |
| Enth., mm BTU/hr | –6.0962E+005 | –1,243.2 | –1,243.2 | –969.64 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 9,5874,696.00 | 195,514.00 | 195,514.00 | 144,110.03 |

| Stream no. | 321 | 322 | 323 | 324 |
|---|---|---|---|---|
| Temp., °F | 129.7508 | 129.7508 | 129.7508 | 129.2004 |
| Pres., psia | 20.3720 | 10.3720 | 10.3720 | 17.4000 |
| Enth., mm BTU/hr | –273.53 | –87.566 | –185.96 | –5,500.6 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 1.0000 |
| Total, lb/hr | 51,403.88 | 12,945.94 | 38,457.95 | 4,472,755.00 |

| Stream no. | 325 |
|---|---|
| Temp., °F | 129.7160 |
| Pres., psia | 17.4000 |
| Enth., mm BTU/hr | –6.0838E+005 |
| Vapor, mass fraction | 0.00000 |
| Total, lb/hr | 95,679,752.00 |

## Figure A1.4. Flue cleaning (PRB coal)



## Table A1.5. Flue cleaning (PRB coal)

| Stream no. | 126 | 302 | 303 | 304 |
|---|---|---|---|---|
| Temp., °F | 295.0000 | 295.0000 | 63.0000 | 63.0000 |
| Pres., psia | 14.1000 | 14.1000 | 15.0000 | 18.2000 |
| Enth., mm BTU/hr | –5,329.6 | –47.225 | –119.54 | –0.90991 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 0.00000 | 1.0000 |
| Total, lb/hr | 4,526,829.00 | 22,049.00 | 18,575.65 | 779.94 |

| Stream no. | 305 | 306 | 307 | 308 |
|---|---|---|---|---|
| Temp., °F | 295.0000 | 379.8514 | 379.7411 | 600.0000 |
| Pres., psia | 14.1000 | 18.2000 | 18.2000 | 18.2000 |
| Enth., mm BTU/hr | –5,376.8 | –5,281.0 | –5,281.9 | –5,015.4 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 4,548,878.00 | 4,548,878.00 | 4,549,658.00 | 4,549,658.00 |

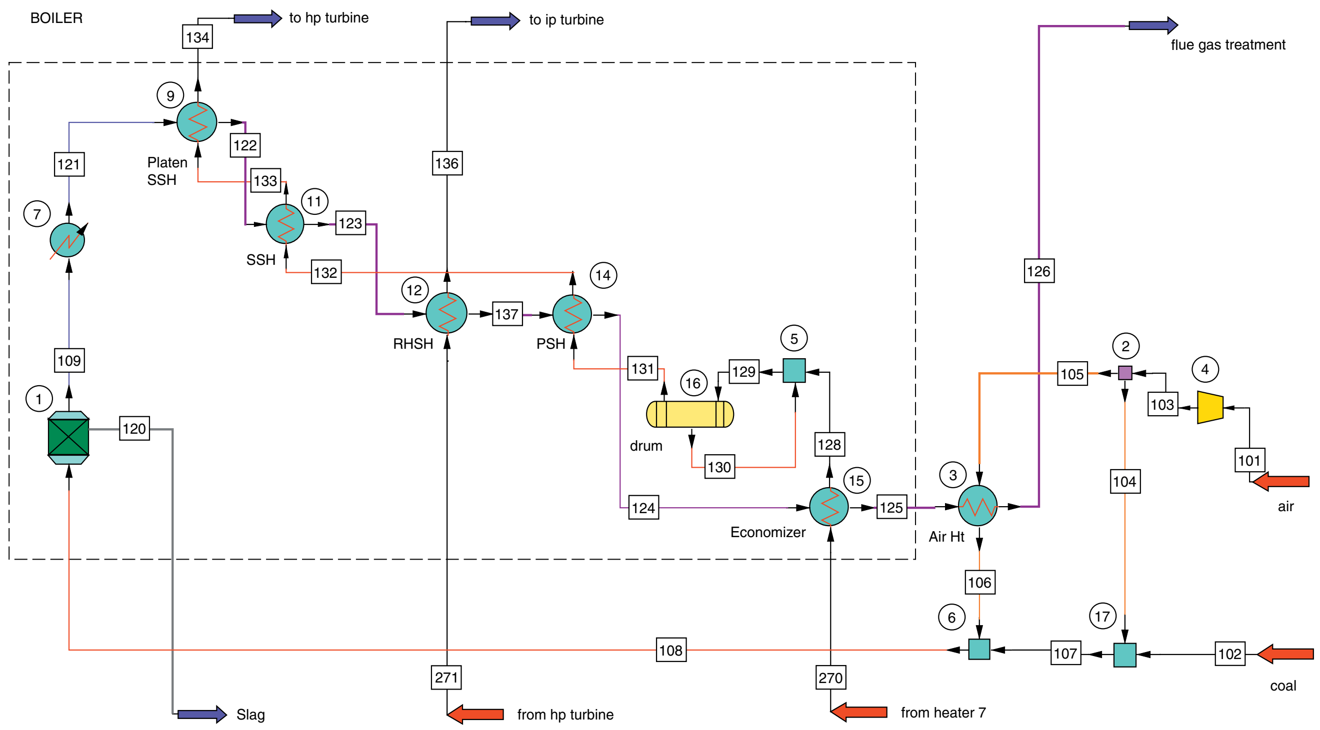
| Stream no. | 309 | 310 | 31 | 312 |
|---|---|---|---|---|
| Temp., °F | 752.0000 | 752.0000 | 536.8289 | 295.0000 |
| Pres., psia | 18.2000 | 17.9000 | 17.9000 | 17.9000 |
| Enth., mm BTU/hr | –4,826.2 | –4,833.9 | –5,100.4 | –5,389.9 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 4,549,658.00 | 4,549,658.00 | 4,549,658.00 | 4,549,658.00 |

| Stream no. | 313 | 314 | 315 | 316 |
|---|---|---|---|---|
| Temp., °F | 68.0000 | 63.0000 | 151.0242 | 146.9291 |
| Pres., psia | 30.0000 | 30.0000 | 17.9000 | 17.6000 |
| Enth., mm BTU/hr | –994.23 | –119.54 | –6,503.6 | –6,503.6 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.99624 | 0.99717 |
| Total, lb/hr | 145,560.00 | 18,575.65 | 4,713,792.50 | 4,713,791.00 |

| Stream no. | 317 | 318 | 324 |
|---|---|---|---|
| Temp., °F | 156.8155 | 156.8155 | 156.8155 |
| Pres., psia | 17.6000 | 17.4000 | 17.4000 |
| Enth., mm BTU/hr | –6,503.6 | –82.409 | –6,409.4 |
| Vapor, mass fraction | 0.99518 | 0.00000 | 1.0000 |
| Total, lb/hr | 4,713,985.50 | 29,743.08 | 4,684,241.50 |

# Appendix A2. Air-blown PC Power Plants with 533-MW (Gross) Output Installed with the MEA Process

## Figure A2.1. Boiler



## Table A2.1. Illinois coal

| Stream no. | 101 | 102 | 103 | 104 |
|---|---|---|---|---|
| Temp., °F | 63.0000 | 63.0000 | 81.6783 | 81.6783 |
| Pres., psia | 14.4000 | 15.8000 | 15.8000 | 15.8000 |
| Enth., mm BTU/hr | –13.910 | –432.10 | 4.0910 | 0.20455 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 | 1.0000 |
| Total, lb/h | 3,991,198.00 | 360,610.00 | 3,991,198.00 | 199,560.00 |

| Stream no. | 105 | 106 | 107 | 108 |
|---|---|---|---|---|
| Temp., °F | 81.6783 | 510.0580 | 52.9506 | 457.8348 |
| Pres., psia | 15.8000 | 15.8000 | 15.8000 | 15.8000 |
| Enth., mm BTU/hr | 3.8865 | 401.70 | –431.90 | –30.193 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.38092 | 0.92903 |
| Total, lb/hr | 3,791,638.00 | 3,791,638.00 | 560,170.00 | 4,351,798.00 |

| Stream no. | 109 | 120 | 121 | 122 |
|---|---|---|---|---|
| Temp., °F | 2,400.0000 | 2,400.0000 | 2,285.7006 | 1,900.0000 |
| Pres., psia | 15.3000 | 15.3000 | 15.3000 | 15.1000 |
| Enth., mm BTU/hr | –1,748.3 | –39.884 | –1,904.3 | –2,422.9 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 | 1.0000 |
| Total, lb/hr | 4,319,700.00 | 32,094.00 | 4,319,700.00 | 4,319,700.00 |

| Stream no. | 123 | 124 | 125 | 126 |
|---|---|---|---|---|
| Temp., °F | 1,700.0000 | 900.0000 | 650.0000 | 295.0000 |
| Pres., psia | 14.9000 | 14.5000 | 14.3000 | 14.1000 |
| Enth., mm BTU/hr | –2,686.3 | –3,689.5 | –3,982.3 | –4,380.1 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 4,319,700.00 | 4,319,700.00 | 4,319,700.00 | 4,319,700.00 |

| Stream no. | 128 | 129 | 130 | 131 |
|---|---|---|---|---|
| Temp., °F | 557.8544 | 557.8544 | 665.1114 | 665.1114 |
| Pres., psia | 2,800.0000 | 2,450.0000 | 2,450.0000 | 2,450.0000 |
| Enth., mm BTU/hr | –21,590 | –21,590 | 0.00000 | –19,676 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 1.0000 |
| Total, lb/hr | 3,422,824.00 | 3,422,824.00 | 0.00 | 3,422,824.00 |

| Stream no. | 132 | 133 | 134 | 136 |
|---|---|---|---|---|
| Temp., °F | 731.8905 | 802.0647 | 1,000.7155 | 1,000.0000 |
| Pres., psia | 2,440.0000 | 2,430.0000 | 2,414.7000 | 545.4000 |
| Enth., mm BTU/hr | –19,298 | –19,034 | –18,516 | –16,177 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 3,422,824.00 | 3,422,824.00 | 3,422,824.00 | 3,022,124.00 |

| Stream no. | 137 | 270 | 271 |
|---|---|---|---|
| Temp., °F | 1,210.4062 | 488.9000 | 634.5002 |
| Pres., psia | 14.7000 | 2800.0000 | 603.6000 |
| Enth., mm BTU/hr | –3,311.3 | –21,883 | –16,802 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 |
| Total, lb/hr | 4,319,700.00 | 3,422,824.00 | 3,022,124.00 |
| IL6# coal–new1 | 0.00 | 0.00 | 0.00 |
| Nitrogen dioxide | 7.90 | 0.00 | 0.00 |

## Table A2.2. PRB coal

| Stream no. | 101 | 102 | 103 | 104 |
|---|---|---|---|---|
| Temp., °F | 63.0000 | 63.0000 | 81.6783 | 81.6783 |
| Pres., psia | 14.4000 | 15.8000 | 15.8000 | 15.8000 |
| Enth., mm BTU/hr | –14.043 | –1,378.5 | 4.1302 | 0.20651 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 | 1.0000 |
| Total, lb/hr | 4,029,474.00 | 524,982.00 | 4,029,474.00 | 201,473.70 |

| Stream no. | 105 | 106 | 107 | 108 |
|---|---|---|---|---|
| Temp., °F | 81.6783 | 536.3796 | 57.3598 | 345.5448 |
| Pres., psia | 15.8000 | 15.8000 | 15.8000 | 15.8000 |
| Enth., mm BTU/hr | 3.9237 | 430.74 | –1,378.3 | –947.51 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.29099 | 0.92109 |
| Total, lb/hr | 3,828,000.25 | 3,828,000.25 | 726,456.00 | 4,554,397.50 |

| Stream no. | 109 | 120 | 121 | 122 |
|---|---|---|---|---|
| Temp., °F | 2,400.0000 | 2,400.0000 | 2,191.6589 | 1,900.0000 |
| Pres., psia | 15.3000 | 15.3000 | 15.3000 | 15.1000 |
| Enth., mm BTU/hr | –2,498.4 | –34.252 | –2,804.7 | –3,226.5 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 | 1.0000 |
| Total, lb/hr | 4,526,829.50 | 27,562.00 | 4,526,829.50 | 4,526,829.50 |

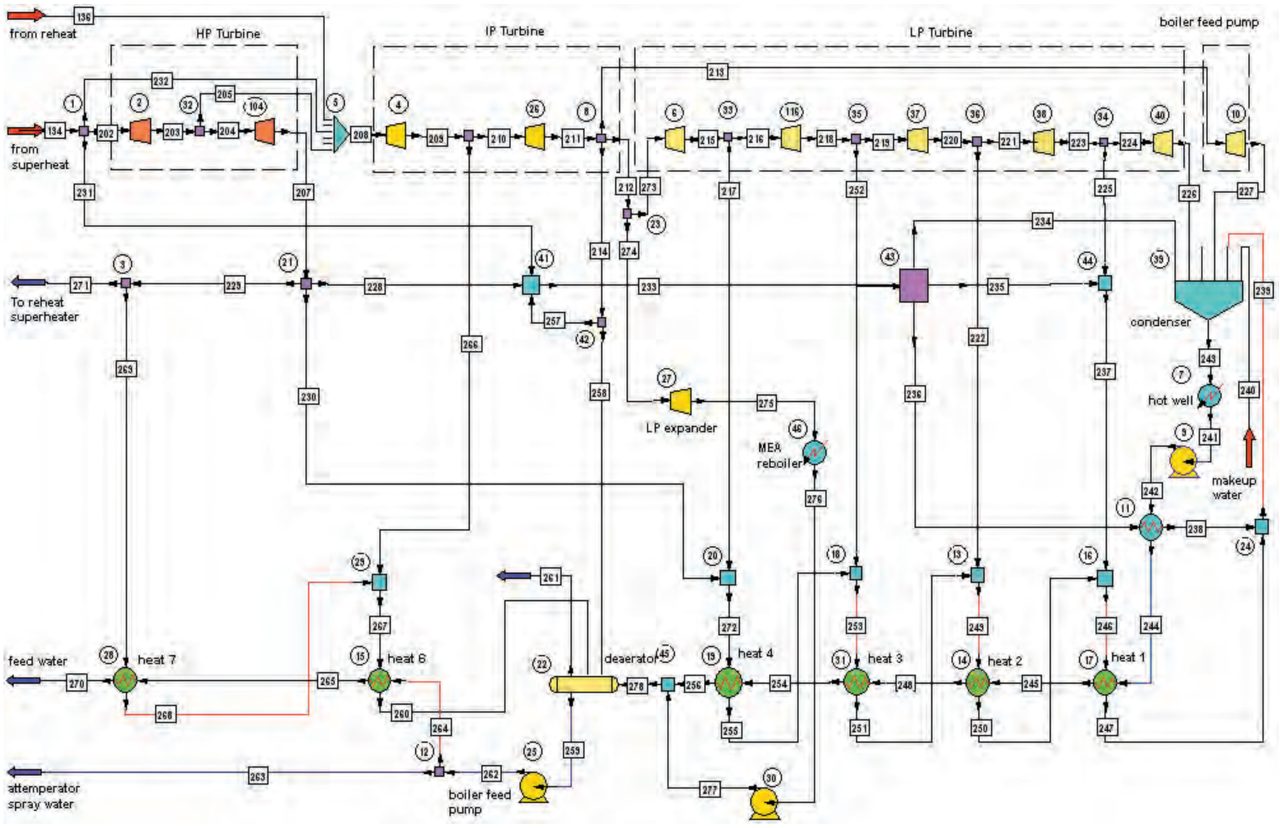
| Stream no. | 123 | 124 | 125 | 126 |
|---|---|---|---|---|
| Temp., °F | 1,700.0000 | 900.0000 | 650.0000 | 295.0000 |
| Pres., psia | 14.9000 | 14.5000 | 14.3000 | 14.1000 |
| Enth., mm BTU/hr | –3,510.3 | –4,588.9 | –4,903.2 | –5,330.0 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 4,526,829.50 | 4,526,829.50 | 4,526,829.50 | 4,526,829.50 |

| Stream no. | 128 | 129 | 130 | 131 |
|---|---|---|---|---|
| Temp., °F | 562.6944 | 562.6944 | 665.1114 | 665.1114 |
| Pres., psia | 2,800.0000 | 2,450.0000 | 2,450.0000 | 2,450.0000 |
| Enth., mm BTU/hr | –2,156 | –21,568 | 0.00000 | –19,676 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 1.0000 |
| Total, lb/hr | 3,422,824.00 | 3,422,824.00 | 0.00 | 3,422,824.00 |

| Stream no. | 132 | 133 | 134 | 136 |
|---|---|---|---|---|
| Temp., °F | 749.9162 | 833.4616 | 1,000.2580 | 1,000.0000 |
| Pres., psia | 2,440.0000 | 2,430.0000 | 2,414.7000 | 545.4000 |
| Enth., mm BTU/hr | –19,222 | –18,938 | –18,517 | –16,177 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 3,422,824.00 | 3,422,824.00 | 3,422,824.00 | 3,022,124.00 |

| Stream no. | 137 | 270 | 271 |
|---|---|---|---|
| Temp., °F | 1,245.7663 | 488.9000 | 634.5002 |
| Pres., psia | 14.7000 | 2,800.0000 | 603.6000 |
| Enth., mm BTU/hr | –4,135.2 | –21,883 | –16,802 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 |
| Total, lb/hr | 4,526,829.50 | 3,422,824.00 | 3,022,124.00 |

## Figure A2.2. Steam cycle (same for Illinois coal and PRN coal)



## Table A2.3. Illinois coal

| Stream no. | 134 | 136 | 202 | 203 |
|---|---|---|---|---|
| Temp., °F | 1,000.0000 | 1,000.0000 | 1,000.0000 | 908.1930 |
| Pres., psia | 2,414.7000 | 545.4000 | 2,414.7000 | 1,775.0000 |
| Enth., mm BTU/hr | 5,000.4 | 4,587.5 | 4,991.5 | 4,863.3 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 3,422,824.00 | 3,022,125.00 | 3,416,724.50 | 3,416,724.50 |

| Stream no. | 204 | 205 | 207 | 208 |
|---|---|---|---|---|
| Temp., °F | 908.1930 | 908.1930 | 631.1807 | 997.5285 |
| Pres., psia | 1,775.0000 | 1,775.0000 | 603.6000 | 545.4000 |
| Enth., mm BTU/hr | 4,806.0 | 57.392 | 4,420.2 | 4,651.9 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 3,376,403.75 | 40,320.77 | 3,376,403.75 | 3,067,190.00 |

| Stream no. | 209 | 210 | 211 | 212 |
|---|---|---|---|---|
| Temp., °F | 817.2190 | 817.2190 | 702.8818 | 702.8818 |
| Pres., psia | 278.9000 | 278.9000 | 174.9000 | 174.9000 |
| Enth., mm BTU/hr | 4,386.6 | 4,240.6 | 4,080.4 | 3,626.8 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 3,067,190.00 | 2,965,172.00 | 2,965,172.00 | 2,635,575.50 |

| Stream no. | 213 | 214 | 215 | 216 |
|---|---|---|---|---|
| Temp., °F | 702.8818 | 702.8818 | 494.9298 | 494.9298 |
| Pres., psia | 174.9000 | 174.9000 | 66.5000 | 66.5000 |
| Enth., mm BTU/hr | 171.69 | 281.87 | 1,606.1 | 1,510.5 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 124,762.58 | 204,834.09 | 1,255,211.50 | 1,180,551.50 |

| | | | | |
|---|---|---|---|---|
| Stream, °F | 494.9298 | 312.4632 | 312.4632 | 216.4570 |
| Pres., psia | 66.5000 | 24.0000 | 24.0000 | 12.8000 |
| Enth., mm BTU/hr | 95.529 | 1,412.4 | 1,364.0 | 1,314.5 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 74,660.00 | 1,180,551.50 | 1,140,068.13 | 1,140,068.13 |

| Stream no. | 221 | 222 | 223 | 224 |
|---|---|---|---|---|
| Temp., °F | 216.4570 | 216.4570 | 172.1188 | 172.1188 |
| Pres., psia | 12.8000 | 12.8000 | 6.3000 | 6.3000 |
| Enth., mm BTU/hr | 1,270.4 | 44.175 | 1,222.2 | 1,141.6 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.97448 | 0.97448 |
| Total, lb/hr | 1,101,756.13 | 38,312.00 | 1,101,756.13 | 1,029,095.38 |

| Stream no. | 225 | 226 | 227 | 228 |
|---|---|---|---|---|
| Temp., °F | 172.1188 | 101.5784 | 104.7907 | 631.1807 |
| Pres., psia | 6.3000 | 1.0000 | 1.1000 | 603.6000 |
| Enth., mm BTU/hr | 80.602 | 1,039.5 | 128.88 | 16.235 |
| Vapor, mass fraction | 0.97448 | 0.90786 | 0.92852 | 1.0000 |
| Total, lb/hr | 72,660.82 | 1,029,095.38 | 124,762.58 | 12,401.53 |

| Stream no. | 229 | 230 | 231 | 232 |
|---|---|---|---|---|
| Temp., °F | 631.1807 | 631.1807 | 1,000.0000 | 1,000.0000 |
| Pres., psia | 603.6000 | 603.6000 | 2,414.7000 | 2,414.7000 |
| Enth., mm BTU/hr | 4,395.0 | 8.9243 | 1.9801 | 6.9305 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 3,357,185.25 | 6,816.96 | 1,355.44 | 4,744.03 |

| Stream no. | 233 | 234 | 235 | 236 |
|---|---|---|---|---|
| Temp., °F | 626.4604 | 626.4604 | 626.4604 | 626.4604 |
| Pres., psia | 174.9000 | 174.9000 | 174.9000 | 174.9000 |
| Enth., mm BTU/hr | 24.561 | 4.0125 | 15.866 | 4.6827 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 18,368.40 | 3,000.79 | 11,865.62 | 3,501.99 |

| Stream no. | 237 | 238 | 239 | 240 |
|---|---|---|---|---|
| Temp., °F | 186.0220 | 200.0000 | 122.4025 | 70.0000 |
| Pres., psia | 6.3000 | 174.9000 | 6.0000 | 14.7000 |
| Enth., mm BTU/hr | 96.468 | 0.58921 | 22.447 | 0.62795 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 84,526.44 | 3,501.99 | 248,301.00 | 16,527.00 |

| Stream no. | 241 | 242 | 243 | 244 |
|---|---|---|---|---|
| Temp., °F | 96.4000 | 97.5560 | 98.0755 | 100.4350 |
| Pres., psia | 0.9000 | 330.0000 | 0.9000 | 330.0000 |
| Enth., mm BTU/hr | 91.548 | 93.191 | 1195.4 | 97.285 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.74644 | 0.00000 |
| Total, lb/hr | 1,421,686.50 | 1,421,686.50 | 1,421,686.50 | 1,421,686.50 |

| Stream no. | 245 | 246 | 247 | 248 |
|---|---|---|---|---|
| Temp., °F | 169.0000 | 169.9850 | 121.2895 | 198.7525 |
| Pres., psia | 300.0000 | 6.0000 | 6.0000 | 270.0000 |
| Enth., mm BTU/hr | 194.90 | 119.47 | 21.858 | 237.41 |
| Vapor, mass fraction | 0.00000 | 0.35145 | 0.00000 | 0.00000 |
| Total, lb/hr | 1,421,686.50 | 244,799.00 | 244,799.00 | 1,421,686.50 |

| Stream no. | 249 | 250 | 251 | 252 |
|---|---|---|---|---|
| Temp., °F | 201.5301 | 175.4000 | 206.7000 | 312.4632 |
| Pres., psia | 11.9000 | 11.9000 | 22.4000 | 24.0000 |
| Enth., mm BTU/hr | 65.519 | 23.001 | 21.343 | 48.435 |
| Vapor, mass fraction | 0.24480 | 0.00000 | 0.00000 | 1.0000 |
| Total, lb/hr | 160,272.41 | 160,272.41 | 121,960.42 | 40,483.48 |

| Stream no. | 253 | 254 | 255 | 256 |
|---|---|---|---|---|
| Temp., °F | 234.0796 | 229.4182 | 239.1000 | 289.7281 |
| Pres., psia | 22.4000 | 240.0000 | 62.4000 | 160.0000 |
| Enth., mm BTU/hr | 65.364 | 281.43 | 16.929 | 368.96 |
| Vapor, mass fraction | 0.34881 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 121,960.42 | 1,421,686.50 | 81,477.00 | 1,421,686.50 |

| Stream no. | 257 | 258 | 259 | 260 |
|---|---|---|---|---|
| Temp., °F | 702.8818 | 702.8818 | 366.0413 | 399.4454 |
| Pres., psia | 174.9000 | 174.9000 | 164.8000 | 263.8000 |
| Enth., mm BTU/hr | 6.3458 | 275.53 | 1,167.9 | 164.23 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.00000 | 0.00000 |
| Total, lb/hr | 4,611.43 | 200,222.67 | 3,439,352.50 | 437,079.31 |

| Stream no. | 261 | 262 | 263 | 264 |
|---|---|---|---|---|
| Temp., °F | 366.0413 | 376.1441 | 376.1441 | 376.1441 |
| Pres., psia | 164.8000 | 2,903.3000 | 2,903.3000 | 2,903.3000 |
| Enth., mm BTU/hr | 0.00000 | 1,205.2 | 41.393 | 1,163.8 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 0.00 | 3,439,352.50 | 118,124.56 | 3,321,228.00 |

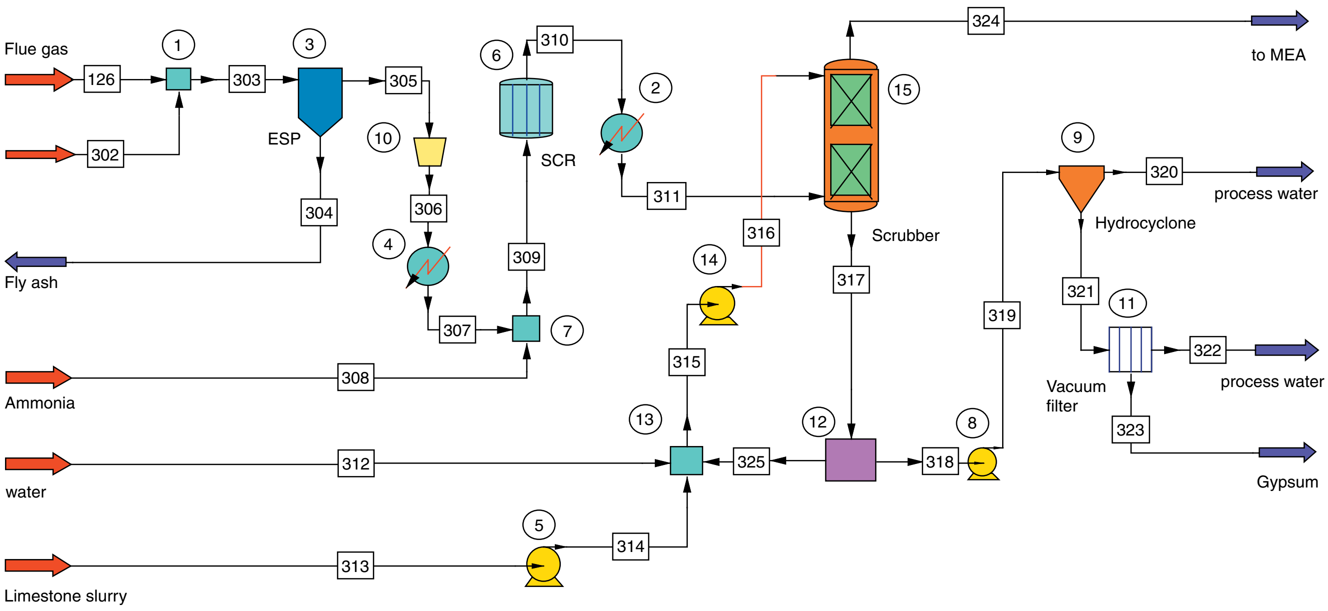
| Stream no. | 265 | 266 | 267 | 268 |
|---|---|---|---|---|
| Temp., °F | 407.5000 | 817.2190 | 410.8331 | 415.5000 |
| Pres., psia | 2,853.3000 | 278.9000 | 278.9000 | 588.5000 |
| Enth., mm BTU/hr | 1,277.3 | 145.90 | 277.73 | 131.83 |
| Vapor, mass fraction | 0.00000 | 1.0000 | 0.30374 | 0.00000 |
| Total, lb/hr | 3,321,228.00 | 102,017.80 | 437,079.31 | 335,061.50 |

| Stream no. | 269 | 270 | 271 | 272 |
|---|---|---|---|---|
| Temp., °F | 631.1807 | 488.3002 | 631.1807 | 499.1817 |
| Pres., psia | 603.6000 | 2,800.0000 | 603.6000 | 62.4000 |
| Enth., mm BTU/hr | 438.64 | 1,584.1 | 3,956.4 | 104.45 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 | 1.0000 |
| Total, lb/hr | 335,061.50 | 3,321,228.00 | 3,022,123.75 | 81,477.00 |

| Stream no. | 273 | 274 | 275 | 276 |
|---|---|---|---|---|
| Temp., °F | 702.8818 | 702.8818 | 475.1697 | 287.5000 |
| Pres., psia | 174.9000 | 174.9000 | 60.0000 | 60.0000 |
| Enth., mm BTU/hr | 1,727.3 | 1,899.5 | 1,753.7 | 355.07 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 0.00000 |
| Total, lb/hr | 1,255,211.50 | 1,380,364.00 | 1,380,364.00 | 1,380,364.00 |

| Stream no. | 277 | 278 |
|---|---|---|
| Temp., °F | 288.4902 | 289.1184 |
| Pres., psia | 330.0000 | 160.0000 |
| Enth., mm BTU/hr | 356.47 | 725.43 |
| Vapor, mass fraction | 0.00000 | 0.00000 |
| Total, lb/h | 1,380,364.00 | 2,802,050.50 |

## Figure A2.3. Flue gas cleaning (Illinois coal)



## Table A2.4. Illinois coal

| Stream no. | 126 | 302 | 303 | 304 |
|---|---|---|---|---|
| Temp., °F | 295.0000 | 295.0000 | 295.0003 | 295.0003 |
| Pres., psia | 14.2000 | 14.2000 | 14.2000 | 14.1000 |
| Enth., mm BTU/hr | –4,374.9 | –54.992 | –4,429.9 | –54.440 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 | 0.00000 |
| Total, lb/hr | 4,319,700.00 | 25,675.20 | 4,345,375.00 | 25,417.41 |

| Stream no. | 305 | 306 | 307 | 308 |
|---|---|---|---|---|
| Temp., °F | 295.0003 | 374.6118 | 752.0000 | 70.0000 |
| Pres., psia | 14.1000 | 18.0000 | 18.0000 | 15.1000 |
| Enth., mm BTU/hr | –4,375.3 | –4,291.8 | –3,863.5 | –0.62266 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 4,319,958.00 | 4,319,958.00 | 4,319,958.00 | 535.74 |

| Stream no. | 309 | 310 | 311 | 312 |
|---|---|---|---|---|
| Temp., °F | 749.3875 | 752.0000 | 295.0000 | 70.0000 |
| Pres., psia | 15.1000 | 17.8000 | 17.8000 | 60.0000 |
| Enth., mm BTU/hr | –3,864.1 | –3,869.2 | –4,385.2 | –1,145.8 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 0.00000 |
| Total, lb/hr | 4,320,493.50 | 4,320,493.50 | 4,320,493.50 | 167803.00 |

| Stream no. | 313 | 314 | 315 | 316 |
|---|---|---|---|---|
| Temp., °F | 70.0000 | 70.1574 | 129.1991 | 129.2440 |
| Pres., psia | 14.7000 | 60.0000 | 17.4000 | 30.0000 |
| Enth., mm BTU/hr | –1,177.2 | –1,177.2 | –6.1073E+005 | –6.1072E+005 |
| Ionic strength molal | 0.0005 | 0.0005 | 0.0572 | 0.0572 |
| Total, lb/hr | 179,427.23 | 179,427.25 | 96,026,952.00 | 96,026,952.00 |

| Stream no. | 317 | 318 | 319 | 320 |
|---|---|---|---|---|
| Temp., °F | 129.7292 | 129.7160 | 129.7508 | 129.7508 |
| Pres., psia | 17.4000 | 17.4000 | 30.0000 | 20.3720 |
| Enth., mm BTU/hr | –6.0962E+005 | –1,243.2 | –1,243.2 | –969.64 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 95,874,696.00 | 195,514.00 | 195,514.00 | 144,110.03 |

| Stream no. | 321 | 322 | 323 | 324 |
|---|---|---|---|---|
| Temp., °F | 129.7508 | 129.7508 | 129.7508 | 129.2004 |
| Pres., psia | 20.3720 | 10.3720 | 10.3720 | 17.4000 |
| Enth., mm BTU/hr | –273.53 | –87.566 | –185.96 | –5,500.6 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 1.0000 |
| Total, lb/hr | 51,403.88 | 12,945.94 | 38,457.95 | 4,472,755.00 |

| Stream no. | 325 |
|---|---|
| Temp., °F | 129.7160 |
| Pres., psia | 17.4000 |
| Enth., mm BTU/hr | –6.0838E+005 |
| Vapor, mass fraction | 0.00000 |
| Total, lb/hr | 95,679,752.00 |

**Figure A2.4. Flue gas cleaning (PRB coal)**

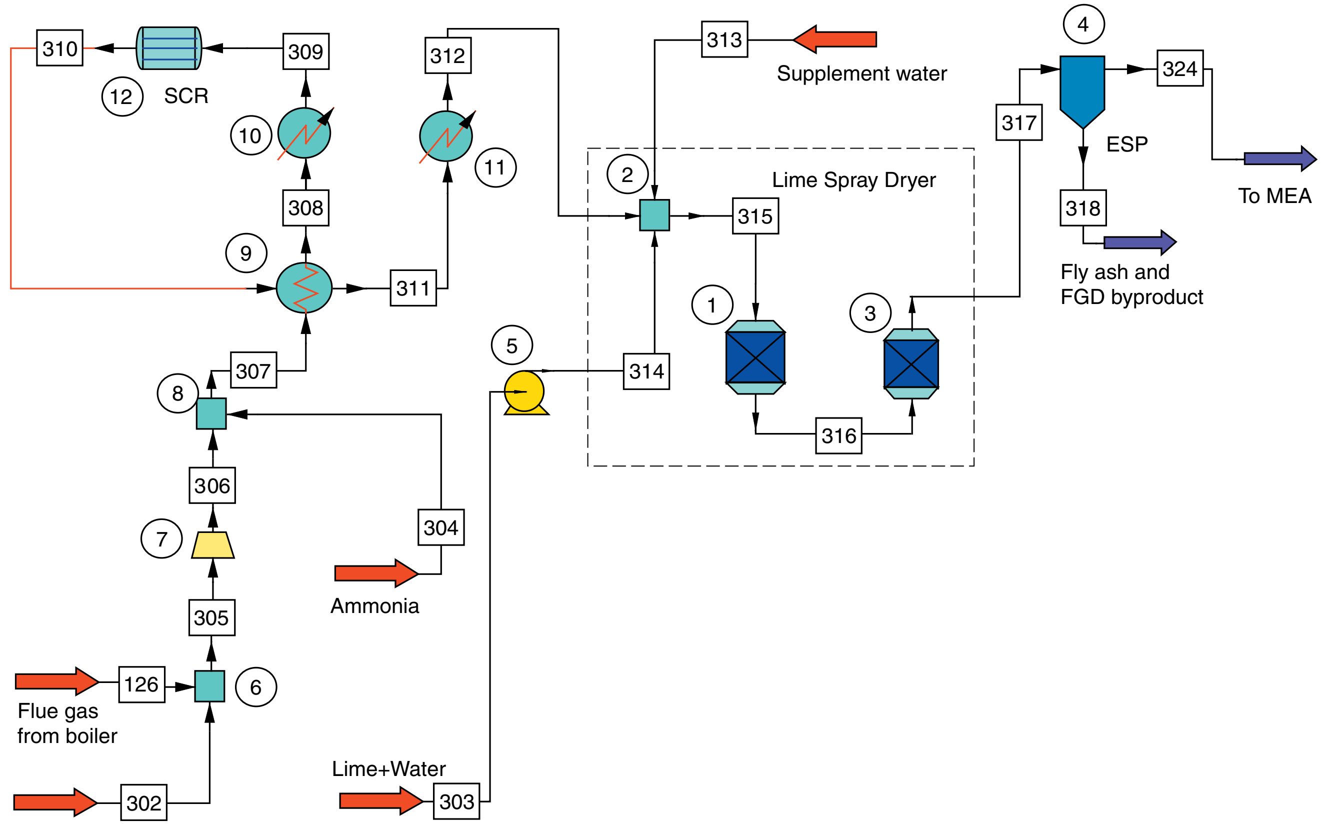


**Table A2.5. Illinois coal and PRB coal**

| Stream no. | 126 | 302 | 303 | 304 |
|---|---|---|---|---|
| Temp., °F | 295.0000 | 295.0000 | 63.0000 | 63.0000 |
| Pres., psia | 14.1000 | 14.1000 | 15.0000 | 18.2000 |
| Enth., mm BTU/hr | –5,329.6 | –47.225 | –119.54 | –0.90991 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 0.00000 | 1.0000 |
| Total, lb/hr | 4,526,829.00 | 22,049.00 | 18,575.65 | 779.94 |

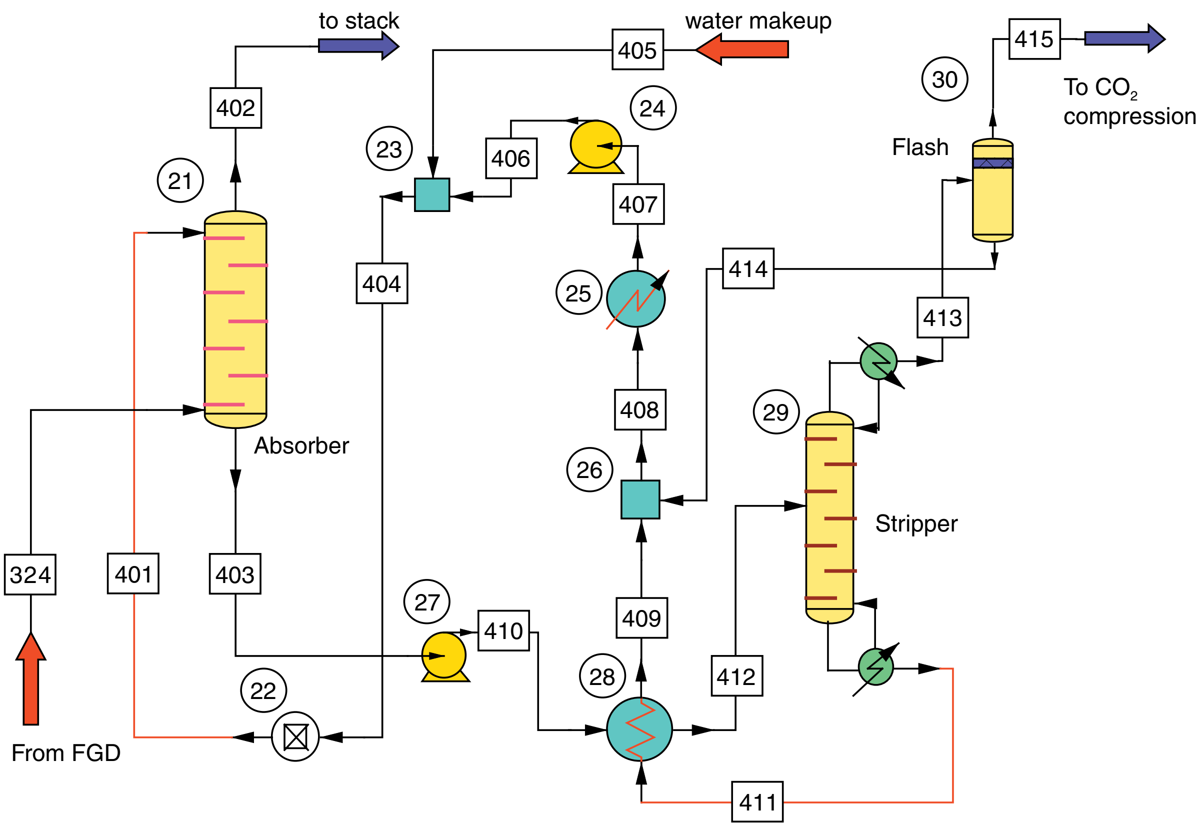
| Stream no. | 305 | 306 | 307 | 308 |
|---|---|---|---|---|
| Temp., °F | 295.0000 | 379.8514 | 379.7411 | 600.0000 |
| Pres., psia | 14.1000 | 18.2000 | 18.2000 | 18.2000 |
| Enth., mm BTU/hr | –5,376.8 | –5,281.0 | –5,281.9 | –5,015.4 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 4,548,878.00 | 4,548,878.00 | 4,549,658.00 | 4,549,658.00 |

| Stream no. | 309 | 310 | 311 | 312 |
|---|---|---|---|---|
| Temp., °F | 752.0000 | 752.0000 | 536.8289 | 295.0000 |
| Pres., psia | 18.2000 | 17.9000 | 17.9000 | 17.9000 |
| Enth., mm BTU/hr | –4,826.2 | –4,833.9 | –5,100.4 | –5,389.9 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 4,549,658.00 | 4,549,658.00 | 4,549,658.00 | 4,549,658.00 |

| Stream no. | 313 | 314 | 315 | 316 |
|---|---|---|---|---|
| Temp., °F | 68.0000 | 63.0000 | 151.0242 | 146.9291 |
| Pres., psia | 30.0000 | 30.0000 | 17.9000 | 17.6000 |
| Enth., mm BTU/hr | –994.23 | –119.54 | –6,503.6 | –6,503.6 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.99624 | 0.99717 |
| Total, lb/hr | 145,560.00 | 18,575.65 | 4,713,792.50 | 4,713,791.00 |

| Stream no. | 317 | 318 | 324 |
|---|---|---|---|
| Temp., °F | 156.8155 | 156.8155 | 156.8155 |
| Pres., psia | 17.6000 | 17.4000 | 17.4000 |
| Enth., mm BTU/hr | –6,503.6 | –82.409 | –6,409.4 |
| Vapor, mass fraction | 0.99518 | 0.00000 | 1.0000 |
| Total, lb/hr | 4,713,985.50 | 29,743.08 | 4,684,241.50 |

**Figure A2.5. MEA process**



## Table A2.6. Illinois coal

| Stream no. | 324 | 401 | 402 | 403 |
|---|---|---|---|---|
| Temp., °F | 129.2004 | 103.3656 | 126.7688 | 135.0667 |
| Pres., psia | 17.4000 | 25.0000 | 14.4000 | 15.0000 |
| Enth., mm BTU/hr | –5,506.2 | –91,894 | –3,688.1 | –93,673 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 | 0.00000 |
| Total, lb/hr | 4,472,743.00 | 17,251,364.00 | 3,879,624.75 | 17,840,666.00 |

| Stream no. | 404 | 405 | 406 | 407 |
|---|---|---|---|---|
| Temp., °F | 103.3656 | 68.0000 | 103.9998 | 104.0000 |
| Pres., psia | 25.0000 | 25.0000 | 25.0000 | 25.0000 |
| Enth., mm BTU/hr | –91,894 | –1,781.3 | –90,113 | –90,113 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 17,251,364.00 | 260,625.22 | 16,990,738.00 | 16,990,738.00 |

| Stream no. | 408 | 409 | 410 | 411 |
|---|---|---|---|---|
| Temp., °F | 163.5854 | 164.4966 | 135.1040 | 240.9159 |
| Pres., psia | 25.0000 | 25.0000 | 25.0000 | 25.0000 |
| Enth., mm BTU/hr | –89,242 | –87,705 | –93,673 | –86,584 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 16,990,738.00 | 16,763,741.00 | 17,840,666.00 | 16,763,468.00 |

| Stream no. | 412 | 413 | 414 | 415 |
|---|---|---|---|---|
| Temp., °F | 202.8000 | 202.8597 | 104.0000 | 104.0000 |
| Pres., psia | 25.0000 | 25.0000 | 25.0000 | 25.0000 |
| Enth., mm BTU/hr | –92,553 | –4,569.0 | –1,537.4 | –3,293.2 |
| Vapor, mass fraction | 0.0057181 | 1.0000 | 0.00000 | 1.0000 |
| Total, lb/hr | 17,840,666.00 | 1,077,198.38 | 226,998.00 | 850,201.00 |

## Table A2.7. PRB coal

| Stream no. | 324 | 401 | 402 | 403 |
|---|---|---|---|---|
| Temp., °F | 156.8155 | 103.6982 | 131.7120 | 143.0264 |
| Pres., psia | 17.4000 | 25.0000 | 14.4000 | 15.0000 |
| Enth., mm BTU/hr | –6,415.1 | –1.0319E+005 | –3,777.0 | –1.0589E+005 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 | 0.00000 |
| Total, lb/hr | 4,683,938.50 | 19,325,904.00 | 3,923,665.75 | 20,097,738.00 |

| Stream no. | 404 | 405 | 406 | 407 |
|---|---|---|---|---|
| Temp., °F | 103.6982 | 68.0000 | 103.9998 | 104.0000 |
| Pres., psia | 25.0000 | 25.0000 | 25.0000 | 25.0000 |
| Enth., mm BTU/hr | –1.0319E+005 | –950.09 | –1.0224E+005 | –1.0224E+005 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 19,325,904.00 | 139,007.44 | 19,186,896.00 | 19,186,896.00 |

| Stream no. | 408 | 409 | 410 | 411 |
|---|---|---|---|---|
| Temp., °F | 171.0819 | 172.1472 | 143.0638 | 241.0265 |
| Pres., psia | 25.0000 | 25.0000 | 25.0000 | 25.0000 |
| Enth., mm BTU/hr | –1.0113E+005 | –99,324 | –1.0589E+005 | –98,181 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 19,186,896.00 | 18,920,272.00 | 20,097,738.00 | 18,920,240.00 |

| Stream no. | 412 | 413 | 414 | 415 |
|---|---|---|---|---|
| Temp., °F | 205.0000 | 205.0469 | 104.0000 | 104.0000 |
| Pres., psia | 25.0000 | 25.0000 | 25.0000 | 25.0000 |
| Enth., mm BTU/hr | –1.0475E+005 | –5,027.7 | –1,805.3 | –3,528.2 |
| Vapor, mass fraction | 0.0042877 | 1.0000 | 0.00000 | 1.0000 |
| Total, lb/hr | 20,097,738.00 | 1,177,496.13 | 266,622.66 | 910,874.00 |

**Figure A2.6. CO$_2$ compression**

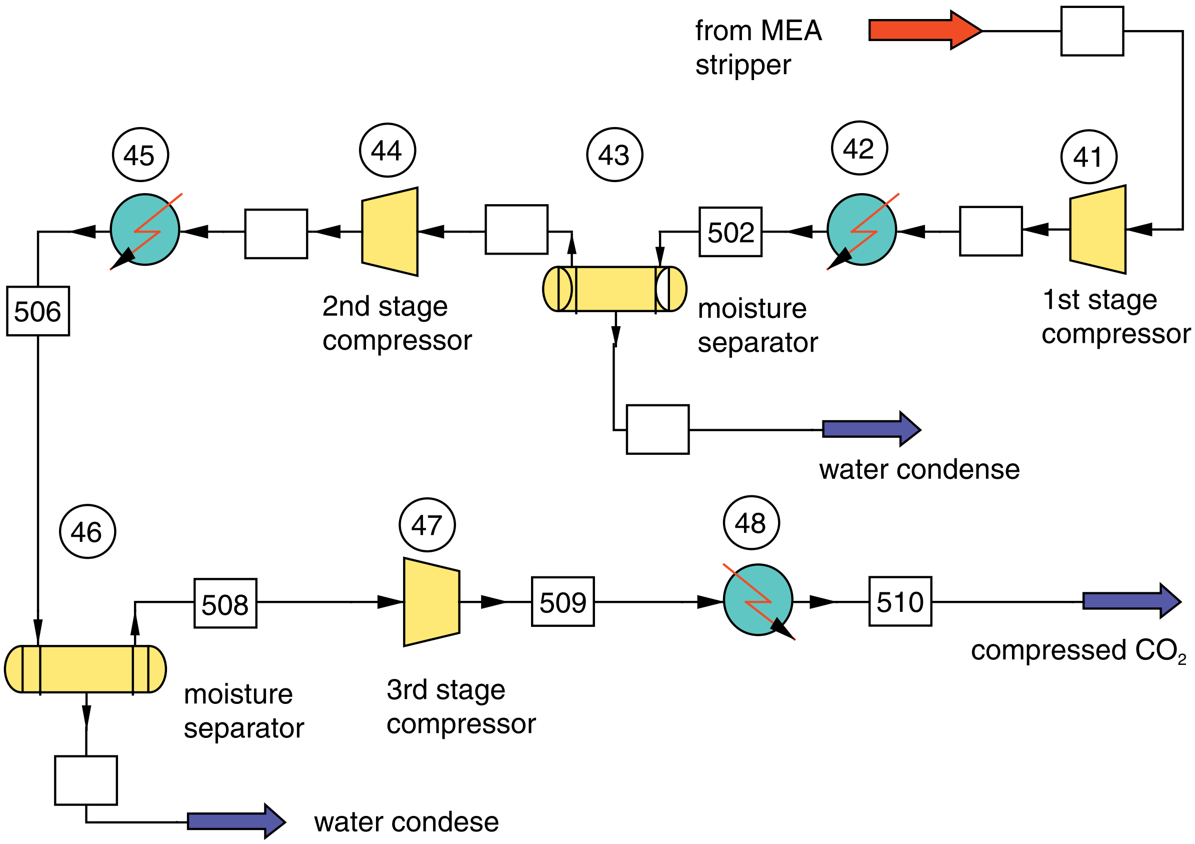


**Table A2.8. Illinois coal**

| Stream no. | 415 | 501 | 502 | 503 |
|---|---|---|---|---|
| Temp., °C | 40.0000 | 200.1252 | 40.0000 | 40.0000 |
| Pres., bar | 1.7237 | 10.0000 | 10.0000 | 10.0000 |
| Enth., mm BTU/hr | –3,293.2 | –3,238.1 | –3,308.9 | –87.230 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.98486 | 0.00000 |
| Total, lb/hr | 850,201.00 | 850,201.00 | 850,201.00 | 12,875.92 |

| Stream no. | 504 | 505 | 506 | 507 |
|---|---|---|---|---|
| Temp., °C | 40.0000 | 225.7489 | 40.0000 | 40.0000 |
| Pres., bar | 10.0000 | 60.0000 | 60.0000 | 60.0000 |
| Enth., mm BTU/hr | –3,221.7 | –3,162.0 | –3,242.4 | –14.919 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.99732 | 0.00000 |
| Total, lb/hr | 8,373,25.00 | 837,325.00 | 837,325.00 | 2,241.62 |

| Stream no. | 508 | 509 | 510 |
|---|---|---|---|
| Temp., °C | 40.0000 | 67.8599 | 20.0000 |
| Pres., bar | 60.0000 | 83.0000 | 83.0000 |
| Enth., mm BTU/hr | –3,227.4 | –3,221.3 | –3,260.7 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.00000 |
| Total, lb/hr | 835,083.00 | 835,083.00 | 835,083.00 |

# Table A2.9. PRB coal

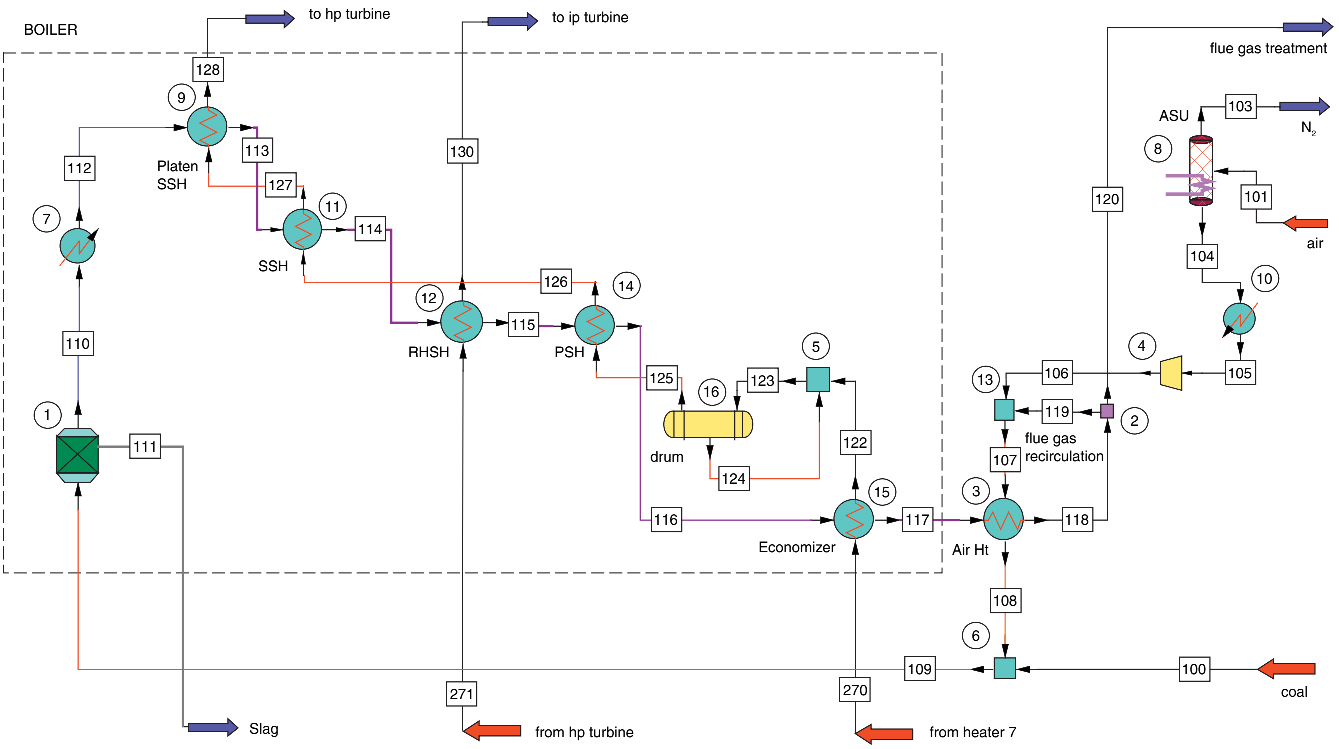
| Stream no. | 415 | 501 | 502 | 503 |
|---|---|---|---|---|
| Temp., °C | 40.0000 | 200.1241 | 40.0000 | 40.0000 |
| Pres., bar | 1.7237 | 10.0000 | 10.0000 | 10.0000 |
| Enth., mm BTU/hr | –3,528.2 | –3,469.2 | –3,545.1 | –93.438 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.98486 | 0.00000 |
| Total, lb/hr | 910,874.00 | 910,874.00 | 910,874.00 | 13,792.23 |

| Stream no. | 504 | 505 | 506 | 507 |
|---|---|---|---|---|
| Temp., °C | 40.0000 | 225.7489 | 40.0000 | 40.0000 |
| Pres., ba | 10.0000 | 60.0000 | 60.0000 | 60.0000 |
| Enth., mm BTU/hr | –3,451.7 | –3,387.7 | –3,473.8 | –15.984 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.99732 | 0.00000 |
| Total, lb/hr | 897,082.00 | 897,082.00 | 897,082.00 | 2,401.42 |

| Stream no. | 508 | 509 | 510 |
|---|---|---|---|
| Temp., °C | 40.0000 | 67.8589 | 20.0000 |
| Pres., bar | 60.0000 | 83.0000 | 83.0000 |
| Enth., mm BTU/hr | –3,457.8 | –3,451.3 | –3,493.4 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.00000 |
| Total, lb/hr | 894,680.00 | 894,680.00 | 8,946,80.00 |

# Appendix A3. Oxy-combustion Power Plants with 533-MW (Gross) Output

## Figure A3.1. Boiler



## Table A3.1. Illinois coal

| Stream no. | 100 | 101 | 103 | 104 |
|---|---|---|---|---|
| Temp., °F | 63.0000 | 63.0000 | –296.2000 | –298.0684 |
| Pres., psia | 15.8000 | 14.4000 | 14.4000 | 14.4000 |
| Enth., mm BTU/hr | –417.44 | –11.988 | –44.47 | –145.34 |
| Vapor, mass fraction | 0.00000 | 1.0000 | 1.0000 | 0.00000 |
| Total, lb/hr | 348,374.00 | 3,461,228.00 | 2,617,629.00 | 843,599.00 |

| Stream no. | 105 | 106 | 107 | 108 |
|---|---|---|---|---|
| Temp., °F | 68.0000 | 86.9714 | 335.7514 | 512.3862 |
| Pres., psia | 14.4000 | 15.8000 | 14.1000 | 14.1000 |
| Enth., mm BTU/hr | –1.7200 | 1.7064 | –11,013 | –10,841 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 843,599.00 | 843,599.00 | 3,770,798.50 | 3,770,824.00 |

| Stream no. | 109 | 110 | 111 | 112 |
|---|---|---|---|---|
| Temp., °F | 481.9835 | 2,400.0000 | 2,400.0000 | 2,222.8632 |
| Pres., psia | 15.8000 | 15.3000 | 15.3000 | 15.3000 |
| Enth., mm BTU/hr | –11,258 | –12,755 | –38.530 | –13,013 |
| Vapor, mass fraction | 0.92755 | 1.0000 | 0.00000 | 1.0000 |
| Total, lb/hr | 4,119,189.00 | 4,088,267.25 | 31,005.00 | 4,088,267.25 |

| Stream no. | 113 | 114 | 115 | 116 |
|---|---|---|---|---|
| Temp., °F | 1900.0000 | 1700.0000 | 1233.3990 | 900.0000 |
| Pres., psia | 15.1000 | 14.9000 | 14.7000 | 14.5000 |
| Enth., mm BTU/hr | –13,473 | –13,751 | –14,376 | –14,797 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 4,088,267.25 | 4,088,267.25 | 4,088,267.25 | 4,088,267.25 |

| Stream no. | 117 | 118 | 119 | 20 |
|---|---|---|---|---|
| Temp., °F | 550.0000 | 395.0000 | 395.0000 | 395.0000 |
| Pres., psia | 14.3000 | 14.1000 | 14.1000 | 14.1000 |
| Enth., mm BTU/hr | –15,211 | –15,384 | –11,015 | –4,369.1 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 4,088,267.25 | 4,088,268.25 | 2,927,199.50 | 1,161,067.88 |

| Stream no. | 122 | 123 | 124 | 125 |
|---|---|---|---|---|
| Temp., °F | 584.6858 | 584.6858 | 665.1114 | 665.1114 |
| Pres., psia | 2,800.0000 | 2,450.0000 | 2,450.0000 | 2,450.0000 |
| Enth., mm BTU/hr | –21,468 | –21,468 | 0.00000 | –19,676 |
| Vapor,, mass fraction | 0.00000 | 0.00000 | 0.00000 | 1.0000 |
| Total, lb/hr | 3,422,824.00 | 3,422,824.00 | 0.00 | 3,422,824.00 |

| Stream no. | 126 | 127 | 28 | 130 |
|---|---|---|---|---|
| Temp., °F | 741.9480 | 820.6823 | 1,000.5983 | 1,000.0000 |
| Pres., psia | 2,440.0000 | 2,430.0000 | 2,414.7000 | 545.4000 |
| Enth., mm BTU/hr | –19,255 | –18,976 | –18,516 | –16,177 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 3,422,824.00 | 3,422,824.00 | 3,422,824.00 | 3,022,124.00 |

| Stream no. | 270 | 271 |
|---|---|---|
| Temp., °F | 488.9000 | 634.5002 |
| Pres., psia | 2,800.0000 | 603.6000 |
| Enth., mm BTU/hr | –21,883 | –16,802 |
| Vapor, mass fraction | 0.00000 | 1.0000 |
| Total, lb/hr | 3,422,824.00 | 3,022,124.00 |

## Table A3.2. PRB coal

| Stream no. | 100 | 101 | 103 | 104 |
|---|---|---|---|---|
| Temp., °F | 63.0000 | 63.0000 | –296.2000 | –297.7483 |
| Pres., psia | 15.8000 | 14.4000 | 14.4000 | 14.4000 |
| Enth., mm BTU/hr | –1,330.2 | –12.029 | –246.55 | –142.66 |
| Vapor, mass fraction | 0.00000 | 1.0000 | 1.0000 | 0.00000 |
| Total, lb/hr | 506,615.00 | 3,457,391.00 | 2,639,851.00 | 817,540.00 |

| Stream no. | 105 | 106 | 107 | 108 |
|---|---|---|---|---|
| Temp., °F | 68.0000 | 86.7959 | 345.0483 | 527.6609 |
| Pres., psia | 14.4000 | 15.8000 | 14.1000 | 14.1000 |
| Enth., mm BTU/hr | –1.6947 | 1.6513 | –13,321 | –13,111 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 817,540.00 | 817,540.00 | 4,089,010.50 | 4,089,067.25 |

| Stream no. | 109 | 110 | 111 | 112 |
|---|---|---|---|---|
| Temp., °F | 384.8558 | 2,400.0000 | 2,400.0000 | 2,002.2980 |
| Pres., psia | 15.8000 | 15.3000 | 15.3000 | 15.3000 |
| Enth., mm BTU/hr | –14,441 | –15,414 | –33.052 | –16,115 |
| Vapor, mass fraction | 0.92455 | 1.0000 | 0.00000 | 1.0000 |
| Total, lb/hr | 4,595,661.00 | 4,569,093.00 | 26,597.00 | 4,569,093.00 |

| Stream no. | 113 | 114 | 115 | 116 |
|---|---|---|---|---|
| Temp., °F | 1,900.0000 | 1,700.0000 | 1,317.3363 | 900.0000 |
| Pres., psia | 15.1000 | 14.9000 | 14.7000 | 14.5000 |
| Enth., mm BTU/hr | –16,291 | –16,629 | –17,254 | –17,896 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 4,569,093.00 | 4,569,093.00 | 4,569,093.00 | 4,569,093.00 |

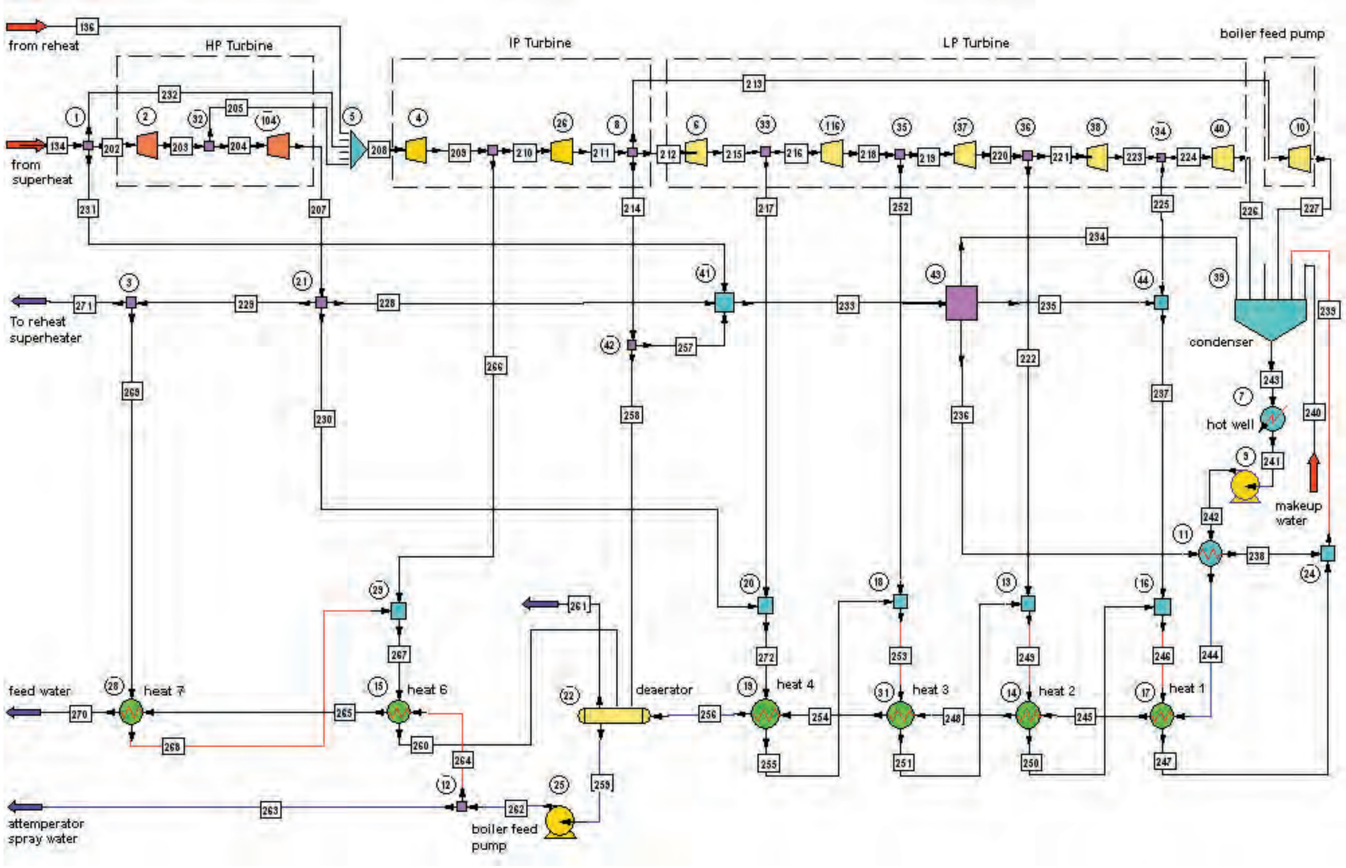
| Stream no. | 117 | 118 | 119 | 120 |
|---|---|---|---|---|
| Temp., °F | 550.0000 | 395.0000 | 395.0000 | 395.0000 |
| Pres., psia | 14.3000 | 14.1000 | 14.1000 | 14.1000 |
| Enth., mm BTU/hr | –18,397 | –18,607 | –13,322 | –5,284.3 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 4,569,093.00 | 4,569,093.00 | 3,271,470.75 | 1,297,622.50 |

| Stream no. | 122 | 123 | 124 | 125 |
|---|---|---|---|---|
| Temp., °F | 603.2884 | 603.2884 | 665.1114 | 665.1114 |
| Pres., psia | 2,800.0000 | 2,450.0000 | 2,450.0000 | 2,450.0000 |
| Enth., mm BTU/hr | –21,382 | –21,382 | 0.00000 | –19,676 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 1.0000 |
| Total, lb/hr | 3,422,824.00 | 3,422,824.00 | 0.00 | 3,422,824.00 |

| Stream no. | 126 | 127 | 128 | 130 |
|---|---|---|---|---|
| Temp., °F | 802.9302 | 925.1138 | 999.0877 | 1,000.0000 |
| Pres., psia | 2,440.0000 | 2,430.0000 | 2,414.7000 | 545.4000 |
| Enth., mm BTU/hr | –19,034 | –18,695 | –18,519 | –16,177 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 3,422,824.00 | 3,422,824.00 | 3,422,824.00 | 3,022,124.00 |

| Stream no. | 270 | 271 |
|---|---|---|
| Temp., °F | 488.9000 | 634.5002 |
| Pres., psia | 2,800.0000 | 603.6000 |
| Enth., mm BTU/hr | –21,883 | –16,802 |
| Vapor, mass fraction | 0.00000 | 1.0000 |
| Total, lb/hr | 3,422,824.00 | 3,022,124.00 |

## Figure A3.2. Steam cycle (same for Illinois coal and PRN coal)



# Table A3.3. Illinois coal and PRB coal

| Stream no. | 134 | 136 | 202 | 203 |
|---|---|---|---|---|
| Temp., °F | 1,000.0000 | 1,000.0000 | 1,000.0000 | 908.1930 |
| Pres., psia | 2,414.7000 | 545.4000 | 2,414.7000 | 1,775.0000 |
| Enth., mm BTU/hr | 5,000.4 | 4,587.5 | 4,991.5 | 4,863.3 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 3,422,824.00 | 3,022,125.00 | 3,416,724.50 | 3,416,724.50 |

| Stream no. | 204 | 205 | 207 | 208 |
|---|---|---|---|---|
| Temp., °F | 908.1930 | 908.1930 | 631.1807 | 997.5285 |
| Pres., psia | 1,775.0000 | 1,775.0000 | 603.6000 | 545.4000 |
| Enth., mm BTU/hr | 4,806.0 | 57.392 | 4,420.2 | 4,651.9 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 3,376,403.75 | 40,320.77 | 3,376,403.75 | 3,067,190.00 |

| Stream no. | 209 | 210 | 211 | 212 |
|---|---|---|---|---|
| Temp., °F | 817.2190 | 817.2190 | 702.8816 | 702.8816 |
| Pres., psia | 278.9000 | 278.9000 | 174.9000 | 174.9000 |
| Enth., mm BTU/hr | 4,386.6 | 4,240.6 | 4,080.4 | 3,626.8 |
| Vapor, mass fraction | 1.0000 | 0.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 3,067,190.00 | 2,965,172.00 | 2,965,172.00 | 2,635,575.50 |

| Stream no. | 213 | 214 | 215 | 216 |
|---|---|---|---|---|
| Temp., °F | 702.8816 | 702.8816 | 494.9297 | 494.9297 |
| Pres., psia | 174.9000 | 174.9000 | 66.5000 | 66.5000 |
| Enth., mm BTU/hr | 171.69 | 281.87 | 3,372.3 | 3,171.7 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 124,762.58 | 204,834.09 | 2,635,575.50 | 2,478,811.50 |

| Stream no. | 217 | 218 | 219 | 220 |
|---|---|---|---|---|
| Temp., °F | 494.9297 | 312.4631 | 312.4631 | 216.4566 |
| Pres., psia | 66.5000 | 24.0000 | 24.0000 | 12.800 |
| Enth., mm BTU/hr | 200.58 | 2,965.7 | 2,864.0 | 2,760.2 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 156,764.03 | 2,478,811.50 | 2,393,808.00 | 2,393,808.00 |

| Stream no. | 221 | 222 | 223 | 224 |
|---|---|---|---|---|
| Temp., °F | 216.4566 | 216.4566 | 172.1188 | 172.1188 |
| Pres., psia | 12.8000 | 12.8000 | 6.3000 | 6.3000 |
| Enth., mm BTU/hr | 2,667.4 | 92.755 | 2,566.2 | 2,397.0 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.97448 | 0.97448 |
| Total, lb/hr | 2,313,364.25 | 80,443.91 | 2,313,364.25 | 2,160,797.75 |

| Stream no. | 225 | 226 | 227 | 228 |
|---|---|---|---|---|
| Temp., °F | 172.1188 | 101.5784 | 104.7907 | 631.1807 |
| Pres., psia | 6.3000 | 1.0000 | 1.1000 | 603.6000 |
| Enth., mm BTU/hr | 169.24 | 2182.6 | 128.88 | 16.235 |
| Vapor, mass fraction | 0.97448 | 0.90786 | 0.92852 | 1.0000 |
| Total, lb/hr | 152,566.38 | 2,160,797.75 | 124,762.58 | 12,401.53 |

| Stream no. | 229 | 230 | 231 | 232 |
|---|---|---|---|---|
| Temp., °F | 631.1807 | 631.1807 | 1,000.0000 | 1,000.0000 |
| Pres., psia | 603.6000 | 603.6000 | 2,414.7000 | 2,414.7000 |
| Enth., mm BTU/hr | 4,395.0 | 8.9243 | 1.9801 | 6.9305 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 3,357,185.25 | 6,816.96 | 1,355.44 | 4,744.03 |

| Stream no. | 233 | 234 | 235 | 236 |
|---|---|---|---|---|
| Temp., °F | 626.4604 | 626.4604 | 626.4604 | 626.4604 |
| Pres., psia | 174.9000 | 174.9000 | 174.9000 | 174.9000 |
| Enth., mm BTU/hr | 24.561 | 4.0125 | 15.866 | 4.6827 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 18,368.40 | 3,000.79 | 11,865.62 | 3,501.99 |

| Stream no. | 237 | 238 | 239 | 240 |
|---|---|---|---|---|
| Temp., °F | 172.1188 | 200.0000 | 118.1270 | 70.0000 |
| Pres., psia | 6.3000 | 174.9000 | 6.0000 | 14.7000 |
| Enth., mm BTU/hr | 185.11 | 0.58921 | 42.801 | 0.62795 |
| Vapor, mass fraction | 0.99101 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 164,432.00 | 3,501.99 | 496,962.22 | 16,527.00 |

| Stream no. | 241 | 242 | 243 | 244 |
|---|---|---|---|---|
| Temp., °F | 96.4000 | 97.5560 | 98.0755 | 99.0167 |
| Pres., psia | 0.9000 | 330.0000 | 0.9000 | 330.0000 |
| Enth., mm BTU/hr | 180.43 | 183.67 | 2,358.9 | 187.77 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.74740 | 0.00000 |
| Total, lb/hr | 2,802,050.50 | 2,802,050.50 | 2,802,050.50 | 2,802,050.50 |

| Stream no. | 245 | 246 | 247 | 248 |
|---|---|---|---|---|
| Temp., °F | 167.4000 | 169.9850 | 117.5445 | 198.7000 |
| Pres., psia | 300.0000 | 6.0000 | 6.0000 | 270.0000 |
| Enth., mm BTU/hr | 379.63 | 234.07 | 42.212 | 467.78 |
| Vapor, mass fraction | 0.00000 | 0.33771 | 0.00000 | 0.00000 |
| Total, lb/hr | 2,802,050.50 | 493,460.25 | 493,460.25 | 2,802,050.50 |

| Stream no. | 249 | 250 | 251 | 252 |
|---|---|---|---|---|
| Temp., °F | 201.5301 | 180.6844 | 210.1221 | 312.4631 |
| Pres., psia | 11.9000 | 11.9000 | 22.4000 | 24.0000 |
| Enth., mm BTU/hr | 137.12 | 48.966 | 44.361 | 101.70 |
| Vapor, mass fraction | 0.25293 | 0.00000 | 0.00000 | 1.0000 |
| Total, lb/hr | 329,028.28 | 329,028.28 | 248,584.39 | 85,003.41 |

| Stream no. | 253 | 254 | 255 | 256 |
|---|---|---|---|---|
| Temp., °F | 234.0796 | 231.1000 | 241.2455 | 292.3000 |
| Pres., psia | 22.4000 | 240.0000 | 62.4000 | 160.0000 |
| Enth., mm BTU/hr | 136.04 | 559.46 | 34.344 | 734.62 |
| Vapor, mass fraction | 0.36067 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 248,584.39 | 2,802,050.50 | 163,581.00 | 2,802,050.50 |

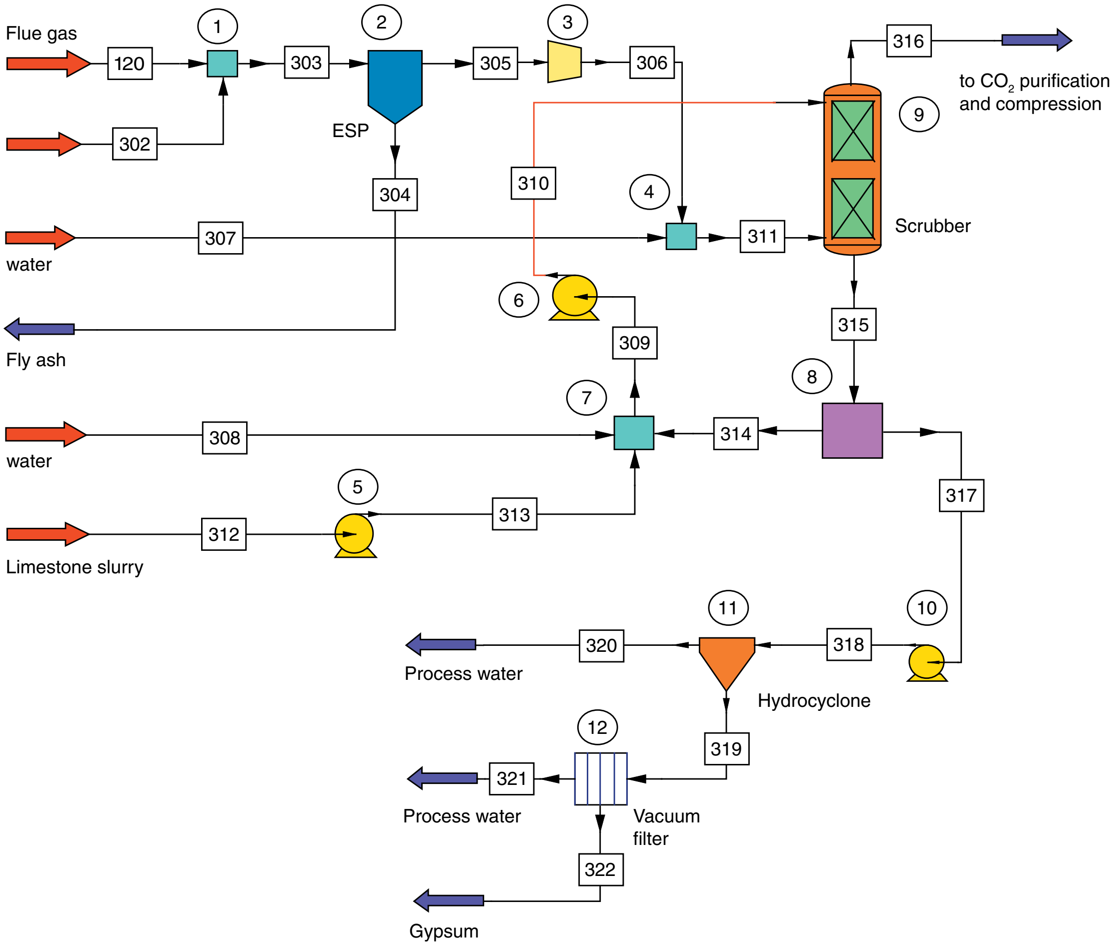
| Stream no. | 257 | 258 | 259 | 260 |
|---|---|---|---|---|
| Temp., °F | 702.8816 | 702.8816 | 366.0413 | 399.4454 |
| Pres., psia | 174.9000 | 174.9000 | 164.8000 | 263.8000 |
| Enth., mm BTU/hr | 6.3458 | 275.53 | 1,167.9 | 164.23 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.00000 | 0.00000 |
| Total, lb/hr | 4,611.43 | 200,222.67 | 3,439,352.50 | 437,079.31 |

| Stream no. | 261 | 262 | 263 | 264 |
|---|---|---|---|---|
| Temp., °F | 366.0413 | 376.1441 | 376.1441 | 376.1441 |
| Pres., psia | 164.8000 | 2,903.3000 | 2,903.3000 | 2,903.3000 |
| Enth., mm BTU/hr | 0.00000 | 1,205.2 | 41.393 | 1,163.8 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 0.00 | 3,439,352.50 | 118,124.56 | 3,321,228.00 |

| Stream no. | 265 | 266 | 267 | 268 |
|---|---|---|---|---|
| Temp., °F | 407.5000 | 817.2190 | 410.8008 | 415.5000 |
| Pres., psia | 2,853.3000 | 278.9000 | 278.8000 | 588.5000 |
| Enth., mm BTU/hr | 1,277.3 | 145.90 | 277.73 | 131.83 |
| Vapor, mass fraction | 0.00000 | 1.0000 | 0.30377 | 0.00000 |
| Total, lb/hr | 3,321,228.00 | 102,017.80 | 437,079.31 | 335,061.50 |

| Stream no. | 269 | 270 | 271 | 272 |
|---|---|---|---|---|
| Temp., °F | 631.1807 | 488.3002 | 631.1807 | 496.6450 |
| Pres., psia | 603.6000 | 2800.0000 | 603.6000 | 62.4000 |
| Enth., mm BTU/hr | 438.64 | ,584.1 | 3,956.4 | 209.51 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 | 1.0000 |
| Total, lb/hr | 335,061.50 | 3,321,228.00 | 3,022,123.75 | 163,581.00 |

## Figure A3.3. Flue gas cleaning (Illinois coal)



## Table A3.4. Illinois coal

| Stream no. | 120 | 302 | 303 | 304 |
|---|---|---|---|---|
| Temp., °F | 395.0000 | 395.0000 | 394.9973 | 394.9973 |
| Pres., psia | 14.2000 | 14.2000 | 14.2000 | 14.1000 |
| Enth., mm BTU/hr | –4,368.6 | –52.566 | –4,421.2 | –52.007 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 | 0.00000 |
| Total, lb/hr | 1,161,067.88 | 24,804.28 | 1,185,872.13 | 24,540.49 |

| Stream no. | 305 | 306 | 307 | 308 |
|---|---|---|---|---|
| Temp., °F | 394.9973 | 408.3329 | 50.0000 | 50.0000 |
| Pres., psia | 14.1000 | 14.8000 | 14.8000 | 17.4000 |
| Enth., mm BTU/hr | –4,369.2 | –4,365.1 | –203.16 | –504.30 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.00000 | 0.00000 |
| Total, lb/hr | 1,161,331.63 | 1,161,331.63 | 29,665.00 | 73,637.30 |

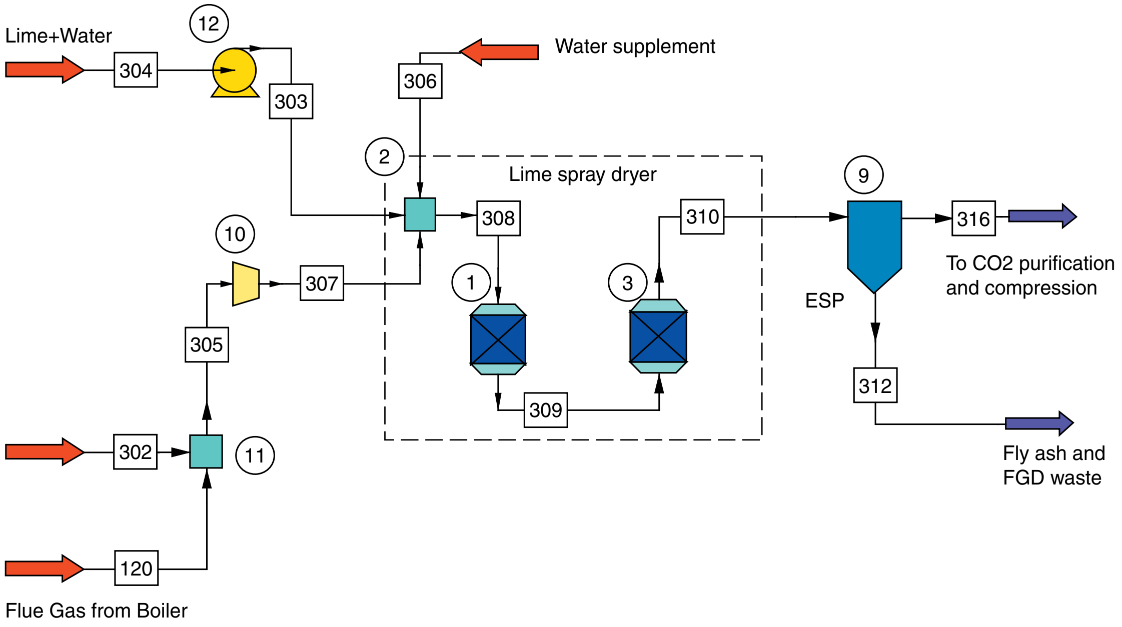
| Stream no. | 309 | 310 | 311 | 312 |
|---|---|---|---|---|
| Temp., °F | 168.2314 | 169.7365 | 295.0083 | 70.0000 |
| Pres., psia | 15.0000 | 435.1130 | 14.8000 | 15.0000 |
| Enth., mm BTU/hr | –1.4107E+005 | –1.4104E+005 | –4,568.2 | –1,137.4 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 1.0000 | 0.00000 |
| Total, lb/hr | 22,412,292.00 | 22,412,288.00 | 1,190,996.63 | 173,353.00 |

| Stream no. | 313 | 314 | 315 | 316 |
|---|---|---|---|---|
| Temp., °F | 70.1575 | 169.4717 | 169.4717 | 169.6657 |
| Pres., psia | 60.0000 | 15.0000 | 15.0000 | 14.5000 |
| Enth., mm BTU/hr | –1,137.3 | –1.3942E+005 | –1.4055E+005 | –5,057.6 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 1.0000 |
| Total, lb/hr | 173,353.00 | 22,165,302.00 | 22,344,162.00 | 1,259,120.75 |

| Stream no. | 317 | 318 | 319 | 320 |
|---|---|---|---|---|
| Temp., °F | 169.4717 | 169.5176 | 169.5176 | 169.5176 |
| Pres., psia | 15.0000 | 30.0000 | 21.1405 | 21.1405 |
| Enth., mm BTU/hr | –1,125.1 | –1,125.1 | –263.29 | –861.76 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 178,861.58 | 178,861.56 | 49,990.19 | 128,871.36 |

| Stream no. | 321 | 322 |
|---|---|---|
| Temp., °F | 169.5176 | 169.5176 |
| Pres., psia | 11.1405 | 11.1405 |
| Enth., mm BTU/hr | –78.181 | –185.11 |
| Vapor, mass fraction | 0.00000 | 0.00000 |
| Total, lb/hr | 11,624.74 | 38,365.45 |

**Figure A3.4. Flue gas cleaning (PRB coal)**



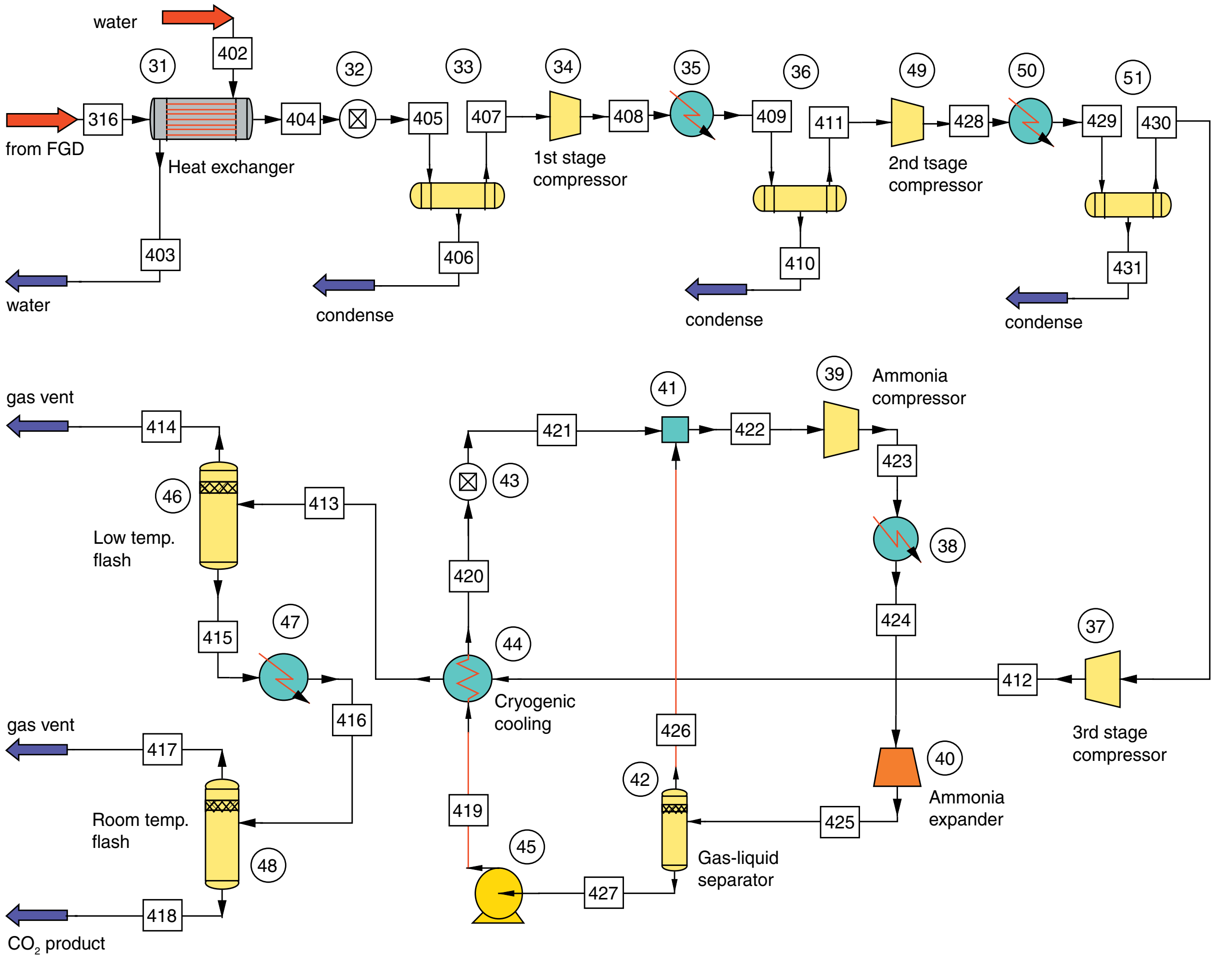
# Table A3.5. PRB coal

| Stream no. | 120 | 302 | 303 | 304 |
|---|---|---|---|---|
| Temp., °F | 395.0000 | 395.0000 | 63.0577 | 63.0000 |
| Pres., psia | 14.1000 | 14.1000 | 30.0000 | 14.7000 |
| Enth., mm BTU/hr | –5,283.7 | –45.093 | –115.15 | –115.15 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 1,297,622.50 | 21,278.00 | 17,929.77 | 17,929.77 |

| Stream no. | 305 | 306 | 307 | 308 |
|---|---|---|---|---|
| Temp., °F | 395.0000 | 63.0000 | 412.0364 | 196.6996 |
| Pres., psia | 14.1000 | 15.0000 | 15.0000 | 15.0000 |
| Enth., mm BTU/hr | –5,328.8 | –407.07 | –5,322.3 | –5,844.5 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 | 1.0000 |
| Total, lb/hr | 1,318,900.50 | 59,515.00 | 1,318,900.50 | 1,396,345.38 |

| Stream no. | 309 | 310 | 312 | 316 |
|---|---|---|---|---|
| Temp., °F | 191.0855 | 201.0469 | 201.0469 | 201.0469 |
| Pres., psia | 14.7000 | 14.7000 | 14.5000 | 14.5000 |
| Enth., mm BTU/hr | –5,844.5 | –5,844.5 | –79.371 | –5,765.1 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.00000 | 1.0000 |
| Total, lb/hr | 1,396,345.13 | 1,396,531.38 | 28,714.08 | 1,367,817.38 |

# Figure A3.5. $CO_2$ purification and compression



# Table A3.6. Illinois coal

| Stream no. | 316 | 402 | 403 | 404 |
|---|---|---|---|---|
| Temp., °C | 76.4809 | 15.0000 | 70.0000 | 40.0000 |
| Pres., bar | 0.9997 | 2.0684 | 1.0132 | 0.9997 |
| Enth., mm BTU/hr | –5,049.4 | –18,985 | –18,710 | –5,324.0 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 0.00000 | 0.80540 |
| Total, lb/hr | 1,259,107.50 | 2,777,197.25 | 2,777,197.25 | 1,259,107.50 |

| Stream no. | 405 | 406 | 407 | 408 |
|---|---|---|---|---|
| Temp., °C | 40.0000 | 40.0000 | 40.0000 | 294.8491 |
| Pres., bar | 0.9997 | 0.9997 | 0.9997 | 10.0000 |
| Enth., mm BTU/hr | –5,324.0 | –1,663.4 | –3,660.5 | –3,564.5 |
| Vapor, mass fraction | 0.80540 | 0.00000 | 1.0000 | 1.0000 |
| Total, lb/hr | 1,259,107.50 | 245,026.47 | 1,014,081.13 | 1,014,081.13 |

| Stream no. | 409 | 410 | 411 | 412 |
|---|---|---|---|---|
| Temp., °C | 40.0000 | 40.0000 | 40.0000 | 134.9481 |
| Pres., bar | 10.0000 | 10.0000 | 10.0000 | 83.0000 |
| Enth., mm BTU/hr | –3,703.6 | –202.40 | –3,501.2 | –3,512.7 |
| Vapor, mass fraction | 0.97051 | 0.00000 | 1.0000 | 1.0000 |
| Total, lb/hr | 1,014,081.13 | 29,909.95 | 984,172.00 | 981,498.00 |

| Stream no. | 413 | 414 | 415 | 416 |
|---|---|---|---|---|
| Temp., °C | –33.0000 | –33.0000 | –33.0000 | 20.0000 |
| Pres., bar | 83.0000 | 83.0000 | 83.0000 | 83.0000 |
| Enth., mm BTU/hr | –3,636.2 | –63.003 | –3,573.2 | –3,521.3 |
| Vapor, mass fraction | 0.084346 | 1.0000 | 0.00000 | 0.027192 |
| Total, lb/hr | 981,498.00 | 82,785.49 | 898,713.00 | 898,713.00 |

| Stream no. | 417 | 418 | 419 | 420 |
|---|---|---|---|---|
| Temp., °C | 20.0000 | 20.0000 | –33.4059 | –33.0000 |
| Pres., bar | 83.0000 | 83.0000 | 1.5000 | 1.0000 |
| Enth., mm BTU/hr | –70.105 | –3,451.2 | –377.19 | –253.62 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 0.00000 | 1.0000 |
| Total, lb/hr | 24,438.16 | 874,275.00 | 209,679.42 | 209,679.39 |

| Stream no. | 421 | 422 | 423 | 424 |
|---|---|---|---|---|
| Temp., °C | –33.0000 | –33.0809 | 180.7371 | 20.0000 |
| Pres., bar | 1.0000 | 1.0000 | 10.0000 | 10.000 |
| Enth., mm BTU/hr | –253.62 | –303.22 | –258.65 | –424.36 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 0.00000 |
| Total, lb/hr | 20,679.39 | 250,675.22 | 250,675.22 | 250,675.22 |

| Stream no. | 425 | 426 | 427 | 428 |
|---|---|---|---|---|
| Temp., °C | –33.4252 | –33.4252 | –33.4252 | 283.5626 |
| Pres., bar | 1.0132 | 1.0132 | 1.0132 | 60.0000 |
| Enth., mm BTU/hr | –426.80 | –49.604 | –377.19 | –3,437.5 |
| Vapor, mass fraction | 0.16354 | 1.0000 | 0.00000 | 1.0000 |
| Total, lb/hr | 250,675.23 | 40,995.81 | 209,679.42 | 984,172.00 |

| Stream no. | 429 | 430 | 431 |
|---|---|---|---|
| Temp., °C | 40.0000 | 40.0000 | 40.0000 |
| Pres., bar | 60.0000 | 60.0000 | 60.0000 |
| Enth., mm BTU/hr | –3,537.8 | –3,520.0 | –17.774 |
| Vapor, mass fraction | 0.99728 | 1.0000 | 0.00000 |
| Total, lb/hr | 984,172.00 | 981,498.00 | 2,673.3 |

# Table A3.7. PRB coal

| Stream no. | 316 | 402 | 403 | 404 |
|---|---|---|---|---|
| Temp., °C | 93.9150 | 15.0000 | 70.0000 | 40.0000 |
| Pres., bar | 0.9997 | 2.0684 | 1.0132 | 0.9997 |
| Enth., mm BTU/hr | –5,755.1 | –26,707 | –26,321 | –6,141.4 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 0.00000 | 0.75253 |
| Total, lb/hr | 1,367,817.38 | 3,906,867.50 | 3,906,867.50 | 1,367,817.38 |

| Stream no. | 405 | 406 | 407 | 08 |
|---|---|---|---|---|
| Temp., °C | 40.0000 | 40.0000 | 40.0000 | 290.3674 |
| Pres., bar | 0.9997 | 0.9997 | 0.9997 | 10.0000 |
| Enth., mm BTU/hr | –6,141.4 | –2,298.0 | –3,843.4 | –3,746.7 |
| Vapor, mass fraction | 0.75253 | 0.00000 | 1.0000 | 1.0000 |
| Total, lb/hr | 1,367,817.38 | 338,497.13 | 1,029,320.31 | 1,029,320.31 |

| Stream no. | 409 | 410 | 411 | 412 |
|---|---|---|---|---|
| Temp., °C | 40.0000 | 40.0000 | 40.0000 | 136.2390 |
| Pres., bar | 10.0000 | 10.0000 | 10.0000 | 83.0000 |
| Enth., mm BTU/hr | –3,887.4 | –204.77 | –3,682.6 | –3,694.5 |
| Vapor, mass fraction | 0.97060 | 0.00000 | 1.0000 | 1.0000 |
| Total, lb/hr | 1,029,320.31 | 30,259.94 | 999,061.00 | 996,355.00 |

| Stream no. | 413 | 414 | 415 | 416 |
|---|---|---|---|---|
| Temp., °C | –33.0000 | –33.0000 | –33.0000 | 20.0000 |
| Pres., bar | 83.0000 | 83.0000 | 83.0000 | 83.0000 |
| Enth., mm BTU/hr | –3,824.2 | –36.507 | –3,787.6 | –3,732.5 |
| Vapor, mass fraction | 0.045073 | 1.0000 | 0.00000 | 0.025284 |
| Total, lb/hr | 996,355.00 | 44,908.40 | 951,447.00 | 951,447.00 |

| Stream no. | 417 | 418 | 419 | 420 |
|---|---|---|---|---|
| Temp., °C | 20.0000 | 20.0000 | –33.4059 | –33.0000 |
| Pres., bar | 83.0000 | 83.0000 | 1.5000 | 1.0000 |
| Enth., mm BTU/hr | –70.468 | –3,662.0 | –395.83 | –266.15 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 0.00000 | 1.0000 |
| Total, lb/hr | 24,056.58 | 927,390.06 | 220,042.64 | 220,043.27 |

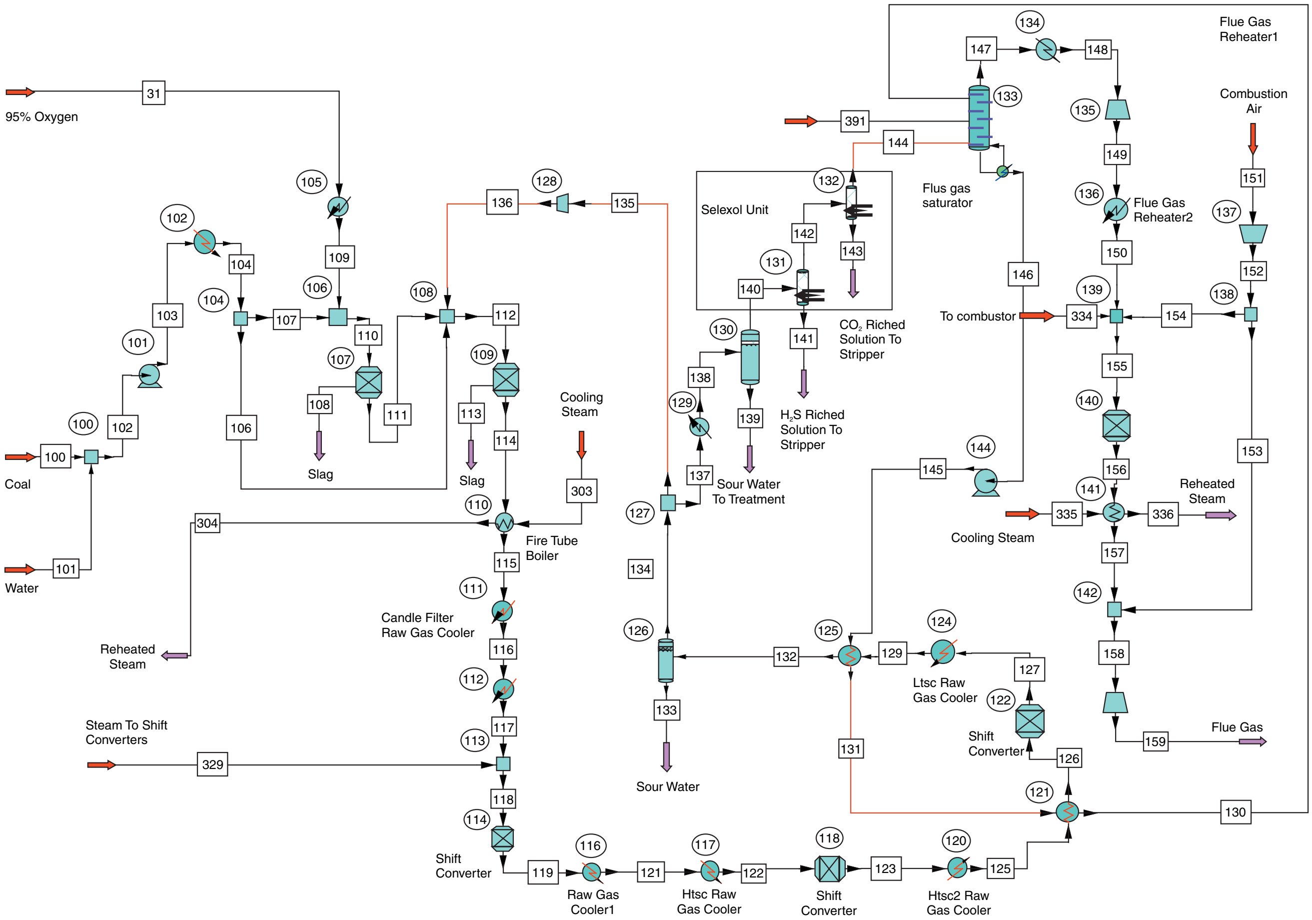
| Stream no. | 421 | 422 | 423 | 424 |
|---|---|---|---|---|
| Temp., °C | –33.0000 | –33.0805 | 180.7371 | 20.0000 |
| Pres., bar | 1.0000 | 1.0000 | 10.0000 | 10.0000 |
| Enth., mm BTU/hr | –266.15 | –318.21 | –271.43 | –445.33 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 0.00000 |
| Total, lb/hr | 220,043.27 | 263,064.63 | 263,064.63 | 263,064.63 |

| Stream no. | 425 | 426 | 427 | 428 |
|---|---|---|---|---|
| Temp., °C | –33.4252 | –33.4252 | –33.4252 | 278.7653 |
| Pres., bar | 1.0132 | 1.0132 | 1.0132 | 60.0000 |
| Enth., mm BTU/hr | –447.89 | –52.056 | –395.84 | –3,618.5 |
| Vapor, mass fraction | 0.16354 | 1.0000 | 0.00000 | 1.0000 |
| Total, lb/hr | 263,064.63 | 43,022.01 | 220,042.64 | 999,061.00 |

| Stream no. | 429 | 430 | 431 |
|---|---|---|---|
| Temp., °C | 40.0000 | 40.0000 | 40.0000 |
| Pres., bar | 60.0000 | 60.0000 | 60.0000 |
| Enth., mm BTU/hr | –3,719.7 | –3,701.8 | –17.992 |
| Vapor, mass fraction | 0.99729 | 1.0000 | 0.00000 |
| Total, lb/hr | 999,061.00 | 996,355.00 | 2,705.51 |

# Appendix A4. IGCC Plants with 533-MW (Gross) Output Installed with Selexol Process

## Figure A4.1. Gasifier, shift reactor, gas cleanup, and gas turbine



## Table A4.1. Illinois coal

| Stream no. | 31 | 100 | 101 | 102 |
|---|---|---|---|---|
| Temp., °F | 239.8713 | 59.0000 | 59.0000 | 59.0001 |
| Pres., psia | 957.0000 | 14.7000 | 14.7000 | 14.7000 |
| Enth., mm BTU/hr | 8.5202 | –243.90 | –1,302.9 | –1,546.8 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 268,910.69 | 324,152.00 | 190,375.00 | 514,527.00 |

| Stream no. | 103 | 104 | 106 | 107 |
|---|---|---|---|---|
| Temp., °F | 61.6427 | 300.0000 | 300.0000 | 300.0000 |
| Pres., psia | 860.0000 | 850.0000 | 850.0000 | 850.0000 |
| Enth., mm BTU/hr | –1,546.2 | –1,494.2 | –313.79 | –1,180.4 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 514,527.00 | 514,527.00 | 108,050.67 | 406,476.34 |

| Stream no. | 108 | 109 | 110 | 111 |
|---|---|---|---|---|
| Temp., °F | 2600.0000 | 305.0000 | 269.2364 | 2,600.0000 |
| Pres., psia | 823.0000 | 954.6000 | 840.0000 | 823.0000 |
| Enth., mm BTU/hr | –23.449 | 12.580 | –1,167.9 | –1,247.1 |
| Vapor, mass fraction | 0.00000 | 1.0000 | 0.42366 | 1.0000 |
| Total, lb/hr | 22,791.12 | 268,910.69 | 675,388.00 | 652,590.00 |

| Stream no. | 112 | 113 | 114 | 115 |
|---|---|---|---|---|
| Temp., °F | 2221.1989 | 1900.0000 | 1900.0000 | 650.0000 |
| Pres., psia | 813.0000 | 800.0000 | 800.0000 | 797.0000 |
| Enth., mm BTU/hr | –1,645.2 | –9.7538 | –1,532.9 | –1,954.1 |
| Vapor, mass fraction | 0.92697 | 0.00000 | 1.0000 | 1.0000 |
| Total, lb/hr | 783,602.00 | 6,058.40 | 777,544.00 | 777,544.00 |

| Stream no. | 116 | 117 | 118 | 119 |
|---|---|---|---|---|
| Temp., °F | 550.0000 | 545.0000 | 562.1572 | 753.3545 |
| Pres., psia | 795.0000 | 792.0000 | 792.0000 | 783.0000 |
| Enth., mm BTU/hr | –1,985.6 | –1,987.2 | –3,623.9 | –3,623.9 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 777,544.00 | 777,544.00 | 1,071,480.38 | 1,071,485.00 |

| Stream no. | 121 | 122 | 123 | 125 |
|---|---|---|---|---|
| Temp., °F | 690.0000 | 550.0000 | 694.9540 | 612.0000 |
| Pres., psia | 774.0000 | 765.0000 | 759.0000 | 750.0000 |
| Enth., mm BTU/hr | –3,654.5 | –3,722.0 | –3,722.0 | –3,762.4 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 1,071,485.00 | 1,071,485.00 | 1,071,488.50 | 1,071,488.50 |

| Stream no. | 126 | 127 | 129 | 130 |
|---|---|---|---|---|
| Temp., °F | 369.8771 | 458.5451 | 370.0000 | 450.0000 |
| Pres., psia | 741.0000 | 741.0000 | 732.0000 | 751.0000 |
| Enth., mm BTU/hr | –3,888.0 | –3,888.0 | –3,931.4 | –5,975.3 |
| Vapor, mass fraction | 0.99184 | 1.0000 | 1.0000 | 0.00033324 |
| Total, lb/hr | 1,071,488.50 | 1,071,490.88 | 1,071,490.88 | 928,248.00 |

| | | | | |
|---|---|---|---|---|
| Stream °F | 326.0000 | 303.3142 | 303.3023 | 303.3023 |
| Pres., psia | 761.0000 | 722.0000 | 722.0000 | 722.0000 |
| Enth., mm BTU/hr | –6,100.6 | –4,041.9 | –585.92 | –3,455.9 |
| Vapor, mass fraction | 0.00000 | 0.91622 | 0.00000 | 1.0000 |
| Total, lb/hr | 928,190.00 | 1,071,490.88 | 89,788.12 | 981,702.00 |

| Stream no. | 135 | 136 | 137 | 138 |
|---|---|---|---|---|
| Temp., °F | 303.3023 | 344.0381 | 303.3023 | 108.0000 |
| Pres., psia | 722.0000 | 850.0000 | 722.0000 | 714.0000 |
| Enth., mm BTU/hr | –84.670 | –84.257 | –3,371.3 | –3,537.4 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 0.91055 |
| Total, lb/hr | 24,051.69 | 24,052.37 | 957,650.00 | 957,650.00 |

| Stream no. | 139 | 140 | 141 | 142 |
|---|---|---|---|---|
| Temp., °F | 105.0000 | 105.0000 | 80.0000 | 90.0000 |
| Pres., psia | 704.0000 | 704.0000 | 700.0000 | 700.0000 |
| Enth., mm BTU/hr | –576.54 | –2,962.2 | –12.125 | –2,956.3 |
| Vapor, mass fraction | 0.00000 | 1.0000 | 0.98456 | 0.99961 |
| Total, lb/hr | 85,787.18 | 871,859.00 | 9,795.49 | 862,063.06 |

| Stream no. | 143 | 144 | 145 | 146 |
|---|---|---|---|---|
| Temp., °F | 90.0000 | 68.0000 | 209.8464 | 209.6072 |
| Pres., psia | 695.0000 | 695.0000 | 770.0000 | 695.0000 |
| Enth., mm BTU/hr | –2,853.7 | –115.00 | –6,211.0 | –6,211.2 |
| Vapor, mass fraction | 1.0000 | 0.99286 | 0.00000 | 0.00000 |
| Total, lb/hr | 738,801.00 | 123,262.51 | 928,190.00 | 928,190.00 |

| Stream no. | 147 | 148 | 149 | 150 |
|---|---|---|---|---|
| Temp., °F | 376.8052 | 520.0000 | 393.3941 | 535.0000 |
| Pres., psia | 690.0000 | 685.0000 | 385.0000 | 380.0000 |
| Enth., mm BTU/hr | –1,092.5 | –1,049.3 | –1,084.8 | –1,043.9 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 306,205.00 | 306,205.00 | 306,205.00 | 306205.00 |

,

| Stream no. | 151 | 152 | 153 | 154 |
|---|---|---|---|---|
| Temp., °F | 68.0000 | 946.0071 | 946.0071 | 946.0071 |
| Pres., psia | 14.4000 | 380.0000 | 380.0000 | 380.0000 |
| Enth., mm BTU/hr | –456.20 | 685.28 | 29.789 | 655.49 |
| Vapor, mass fraction | 0.99989 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 5,140,434.00 | 5,140,434.00 | 223,454.69 | 4,916,979.50 |

| Stream no. | 155 | 156 | 157 | 158 |
|---|---|---|---|---|
| Temp., °F | 861.7679 | 2,557.2514 | 2,449.7438 | 2,400.0109 |
| Pres., psia | 380.0000 | 353.0000 | 353.0000 | 353.0000 |
| Enth., mm BTU/hr | –938.29 | –938.30 | –1,138.4 | –1,108.6 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 5,322,013.00 | 5,322,005.50 | 5,322,005.50 | 5,545,460.50 |

| Stream no. | 159 | 303 | 304 | 329 |
|---|---|---|---|---|
| Temp., °F | 1095.6921 | 620.0002 | 631.9833 | 645.0002 |
| Pres., psia | 15.2000 | 2000.0000 | 1950.0000 | 800.0000 |
| Enth., mm BTU/hr | –3,473.7 | –6,402.5 | –5,981.3 | –1,636.7 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 0.80997 | 1.0000 |
| Total, lb/hr | 5,545,461.50 | 1,028,788.00 | 1,028,788.00 | 293,937.00 |

| Stream no. | 334 | 335 | 336 | 391 |
|---|---|---|---|---|
| Temp., °F | 598.5389 | 598.5389 | 1,000.0000 | 267.0000 |
| Pres., psia | 385.0000 | 385.0000 | 347.0000 | 855.0000 |
| Enth., mm BTU/hr | –549.87 | –5,138.1 | –4,938.0 | –1,213.4 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 0.00000 |
| Total, lb/hr | 98,828.41 | 923,480.00 | 923,480.00 | 182,884.48 |

## Table A4.2. PRB coal

| Stream no. | 31 | 100 | 101 | 102 |
|---|---|---|---|---|
| Temp., °F | 239.8713 | 59.0000 | 59.0000 | 58.9989 |
| Pres., psia | 957.0000 | 14.7000 | 14.7000 | 14.7000 |
| Enth., mm BTU/hr | 9.1419 | –1,056.9 | –1,216.6 | –2,273.5 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 288,533.41 | 466,261.41 | 177,761.59 | 644,023.13 |

| Stream no. | 103 | 104 | 106 | 107 |
|---|---|---|---|---|
| Temp., °F | 61.5821 | 300.0000 | 300.0000 | 300.0000 |
| Pres., psia | 860.0000 | 850.0000 | 850.0000 | 850.0000 |
| Enth., mm BTU/hr | –2,272.7 | –2,198.8 | –461.75 | –1,737.1 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 644,023.13 | 644,023.13 | 135,244.84 | 508,778.28 |

| Stream no. | 108 | 109 | 110 | 111 |
|---|---|---|---|---|
| Temp., °F | 2,600.0000 | 305.0000 | 274.2680 | 2,600.0000 |
| Pres., psia | 823.0000 | 954.6000 | 840.0000 | 823.0000 |
| Enth., mm BTU/hr | –19.890 | 13.498 | –1,723.6 | –1,806.8 |
| Vapor, mass fraction | 0.00000 | 1.0000 | 0.56388 | 1.0000 |
| Total, lb/hr | 19,331.82 | 288,533.41 | 797,312.06 | 777,974.00 |

| Stream no. | 112 | 113 | 114 | 115 |
|---|---|---|---|---|
| Temp., °F | 2,190.0837 | 1,900.0000 | 1,900.0000 | 650.0000 |
| Pres., psia | 813.0000 | 800.0000 | 800.0000 | 797.0000 |
| Enth., mm BTU/hr | –2,367.0 | –8.2734 | –2,256.2 | –2,772.9 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 | 1.0000 |
| Total, lb/hr | 940,136.00 | 5,138.84 | 934,997.00 | 934,997.00 |

| Stream no. | 116 | 117 | 118 | 119 |
|---|---|---|---|---|
| Temp., °F | 550.0000 | 545.0000 | 562.6477 | 718.0000 |
| Pres., psia | 795.0000 | 792.0000 | 792.0000 | 783.0000 |
| Enth., mm BTU/hr | –2,811.7 | –2,813 | –4,450.4 | –4,450.4 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 934,997.00 | 934,997.00 | 122,933.63 | 1,228,938.13 |

| Stream no. | 121 | 122 | 123 | 125 |
|---|---|---|---|---|
| Temp., °F | 665.0000 | 550.0000 | 668.0000 | 612.0000 |
| Pres., psia | 774.0000 | 765.0000 | 759.0000 | 750.0000 |
| Enth., mm BTU/hr | –4,479.8 | –4,544.1 | –4,544.1 | –4,575.4 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 1,228,938.13 | 1,228,938.13 | 1,228,941.38 | 1,228,941.38 |

| Stream no. | 126 | 127 | 129 | 130 |
|---|---|---|---|---|
| Temp., °F | 390.1741 | 473.6802 | 370.0000 | 450.0000 |
| Pres., psia | 741.0000 | 741.0000 | 732.0000 | 751.0000 |
| Enth., mm BTU/hr | –4,703.6 | –4,700.3 | –4,777.1 | –6,100.2 |
| Vapor, mass fraction | 0.99854 | 1.0000 | 0.98305 | 0.00035571 |
| Total, lb/hr | 1,228,941.25 | 1,228,943.50 | 1,228,941.00 | 947,713.00 |

| Stream no. | 131 | 132 | 133 | 134 |
|---|---|---|---|---|
| Temp., °F | 326.0000 | 335.2399 | 335.2399 | 335.2399 |
| Pres., psia | 761.0000 | 722.0000 | 722.0000 | 722.0000 |
| Enth., mm BTU/hr | –6,228.5 | –4,889.7 | –851.71 | –4,038.0 |
| Vapor, mass fraction | 0.00000 | 0.89407 | 0.00000 | 1.0000 |
| Total, lb/hr | 947,713.00 | 1,228,941.00 | 130,193.66 | 1,098,759.63 |

| Stream no. | 135 | 136 | 137 | 138 |
|---|---|---|---|---|
| Temp., °F | 335.2399 | 376.8267 | 335.2399 | 108.0000 |
| Pres., psia | 722.0000 | 850.0000 | 722.0000 | 714.0000 |
| Enth., mm BTU/hr | –98.932 | –98.441 | –939.1 | –4,192.5 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 0.85952 |
| Total, lb/hr | 26,919.61 | 26,917.07 | 1,071,840.00 | 1,071,840.00 |

| Stream no. | 139 | 140 | 141 | 142 |
|---|---|---|---|---|
| Temp., °F | 105.0000 | 105.0000 | 80.0000 | 90.0000 |
| Pres., psia | 704.0000 | 704.0000 | 700.0000 | 700.0000 |
| Enth., mm BTU/hr | –1,015.7 | –3,178.4 | –10.903 | –3,173.8 |
| Vapor, mass fraction | 0.00000 | 1.0000 | 0.96268 | 0.99962 |
| Total, lb/hr | 150,733.64 | 921,108.00 | 3,784.20 | 917,323.00 |

| Stream no. | 143 | 144 | 145 | 146 |
|---|---|---|---|---|
| Temp., °F | 90.0000 | 68.0000 | 209.9997 | 315.9250 |
| Pres., psia | 695.0000 | 695.0000 | 770.0000 | 695.0000 |
| Enth., mm BTU/hr | –3,044.3 | –142.39 | –6,341 | –6,833.2 |
| Vapor, mass fraction | 1.0000 | 0.99291 | 0.00000 | 0.00000 |
| Total, lb/hr | 788,137.00 | 129,186.65 | 947,713.00 | 1,038,595.06 |

| Stream no. | 147 | 148 | 149 | 150 |
|---|---|---|---|---|
| Temp., °F | 333.3868 | 520.0000 | 393.2608 | 535.0000 |
| Pres., psia | 690.0000 | 685.0000 | 385.0000 | 380.0000 |
| Enth., mm BTU/hr | –610.92 | –1,096.3 | –1,132.5 | –1,090.7 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 221,188.53 | 315,953.25 | 315,953.25 | 315,953.25 |

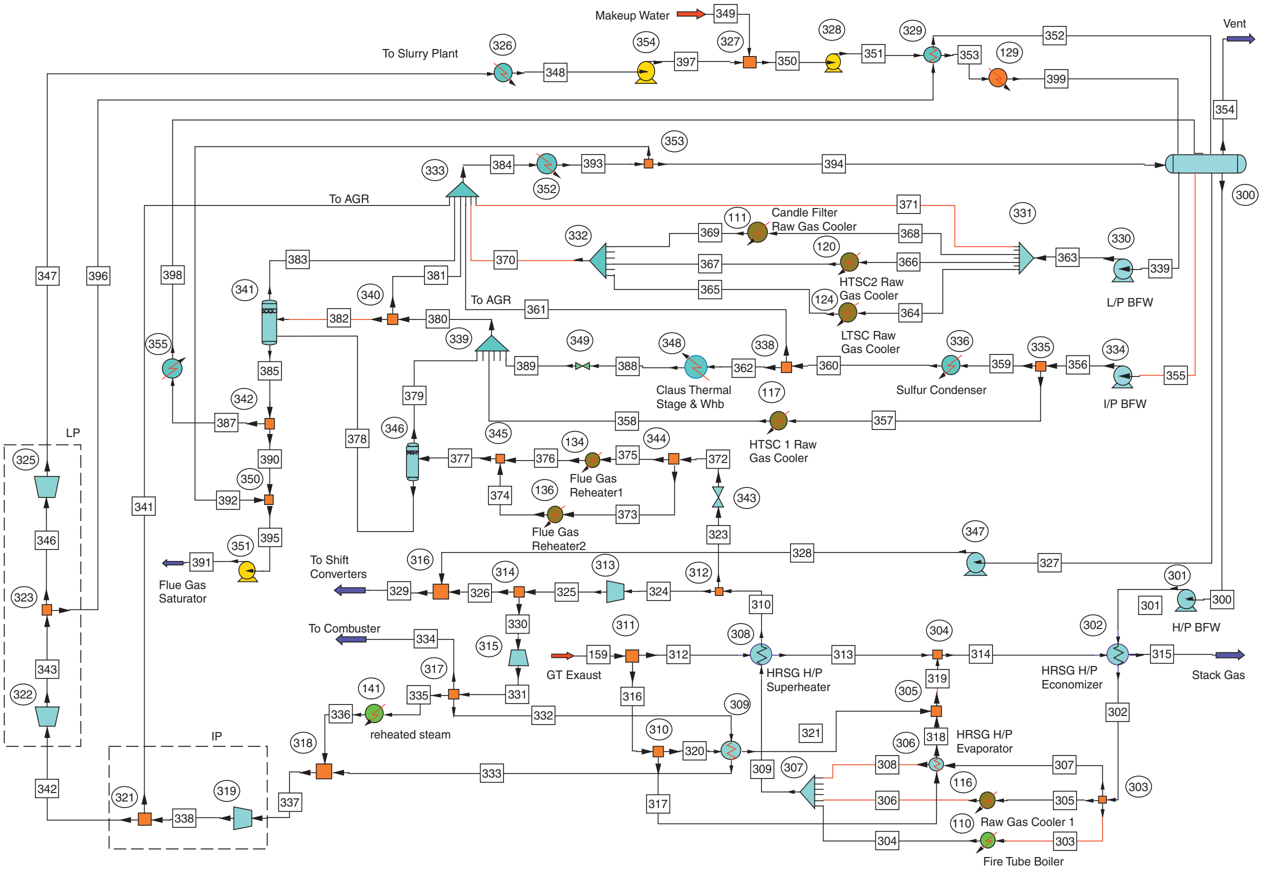
| Stream no. | 151 | 152 | 153 | 154 |
|---|---|---|---|---|
| Temp., °F | 68.0000 | 946.0071 | 946.0071 | 946.0071 |
| Pres., psia | 14.4000 | 380.0000 | 380.0000 | 380.0000 |
| Enth., mm BTU/hr | –445.73 | 669.55 | 29.105 | 640.44 |
| Vapor, mass fraction | 0.99989 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 5,022,433.00 | 5,022,433.00 | 218,325.16 | 4,804,108.00 |

| Stream no. | 155 | 156 | 157 | 158 |
|---|---|---|---|---|
| Temp., °F | 859.0961 | 2,574.4999 | 2,465.4452 | 2,415.5973 |
| Pres., psia | 380.0000 | 353.0000 | 353.0000 | 353.0000 |
| Enth., mm BTU/hr | –1,000.2 | –994.20 | –1,194.3 | –1,165.2 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 5,218,890.00 | 5,218,882.00 | 5,218,882.00 | 5,437,207.50 |

| Stream no. | 159 | 303 | 304 | 329 |
|---|---|---|---|---|
| Temp., °F | 1,106.4225 | 620.0002 | 632.3465 | 645.0002 |
| Pres., psia | 15.2000 | 2,000.0000 | 1,950.0000 | 800.0000 |
| Enth., mm BTU/hr | –3,503.9 | –6,402.5 | –5,885.9 | –1,636.7 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 1.0000 | 1.0000 |
| Total, lb/hr | 5,437,207.50 | 1,028,788.00 | 1,028,788.00 | 293,937.00 |

| Stream no. | 334 | 335 | 336 | 391 |
|---|---|---|---|---|
| Temp., °F | 598.5389 | 598.5389 | 1,000.0000 | 267.0000 |
| Pres., psia | 385.0000 | 385.0000 | 347.0000 | 855.0000 |
| Enth., mm BTU/hr | –549.87 | –5,138.1 | –4,938.0 | –1,213.4 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 0.00000 |
| Total, lb/hr | 98,828.41 | 923,480.00 | 923,480.00 | 182,884.48 |

## Figure A4.2. Steam cycle



## Table A4.3. Illinois coal

| Stream no. | 159 | 300 | 301 | 302 |
|---|---|---|---|---|
| Temp., °F | 1,095.6921 | 240.1082 | 247.5038 | 620.0000 |
| Pres., psia | 15.2000 | 25.0000 | 2,300.0000 | 2,000.0000 |
| Enth., mm BTU/hr | –3,474.0 | –9,714.1 | –9,703.1 | –9,074.5 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 5,545,462.00 | 1,458,127.00 | 1,458,127.00 | 1,458,127.00 |

| Stream no. | 303 | 304 | 305 | 306 |
|---|---|---|---|---|
| Temp., °F | 620.0000 | 632.2300 | 620.0000 | 632.1550 |
| Pres., psia | 2,000.0000 | 1,950.0000 | 2,000.0000 | 1,950.0000 |
| Enth., mm BTU/hr | –6,402.5 | –5,886.1 | –378.99 | –348.43 |
| Vapor, mass fraction | 0.00000 | 1.0000 | 0.00000 | 1.0000 |
| Total, lb/hr | 1,028,788.00 | 1,028,788.00 | 60,898.00 | 60,898.00 |

| Stream no. | 307 | 308 | 309 | 310 |
|---|---|---|---|---|
| Temp., °F | 620.0000 | 632.2000 | 631.9833 | 1,004.0000 |
| Pres., psia | 2,000.0000 | 2,000.0000 | 1,950.0000 | 1,840.0000 |
| Enth., mm BTU/hr | –2,293.0 | –2,286.5 | –8,521.0 | –7,858.0 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.74815 | 1.0000 |
| Total, lb/hr | 368,441.00 | 368,441.00 | 1,458,127.00 | 1,458,127.00 |

| Stream no. | 312 | 313 | 314 | 315 |
|---|---|---|---|---|
| Temp., °F | 1095.6921 | 665.2483 | 680.4660 | 277.0513 |
| Pres., psia | 15.2000 | 15.0000 | 15.0000 | 14.8000 |
| Enth., mm BTU/hr | –3,265.6 | –3,928.7 | –4,155.2 | –4,783.8 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 5,212,733.50 | 5,212,733.50 | 5,545,462.00 | 5,545,462.00 |

| Stream no. | 316 | 317 | 318 | 319 |
|---|---|---|---|---|
| Temp., °F | 1,095.6921 | 1,095.6921 | 880.4981 | 915.0102 |
| Pres., psia | 15.2000 | 15.2000 | 15.2000 | 15.0000 |
| Enth., mm BTU/hr | –208.44 | –62.532 | –68.970 | –226.50 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 332,728.00 | 99,818.31 | 99,818.31 | 332,728.00 |

| Stream no. | 320 | 321 | 323 | 324 |
|---|---|---|---|---|
| Temp., °F | 1,095.6921 | 929.7526 | 1,004.0000 | 1,004.0000 |
| Pres., psia | 15.2000 | 15.0000 | 18,40.0000 | 1,840.0000 |
| Enth., mm BTU/hr | –145.91 | –157.53 | –567.52 | –7,290.4 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 232,909.38 | 232,909.38 | 105,308.85 | 1,352,818.25 |

| Stream no. | 325 | 326 | 327 | 328 |
|---|---|---|---|---|
| Temp., °F | 772.5801 | 772.5801 | 240.1082 | 242.1404 |
| Pres., psia | 800.0000 | 800.0000 | 25.0000 | 650.0000 |
| Enth., mm BTU/hr | –7,425.0 | –1,519.7 | –113.65 | –113.62 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.00000 | 0.00000 |
| Total, lb/hr | 1,352,818.25 | 276,877.25 | 17,060.00 | 17,060.00 |

| Stream no. | 329 | 330 | 331 | 332 |
|---|---|---|---|---|
| Temp., °F | 645.0892 | 772.5801 | 598.5389 | 598.5389 |
| Pres., psia | 650.0000 | 800.0000 | 385.0000 | 385.0000 |
| Enth., mm BTU/hr | –1,633.3 | –5,905.4 | –5,986.4 | –298.41 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 293,937.25 | 1,075,941.00 | 1,075,941.00 | 53,633.00 |

| Stream no. | 333 | 334 | 335 | 336 |
|---|---|---|---|---|
| Temp., °F | 1,000.0000 | 598.5389 | 598.5389 | 1,000.0000 |
| Pres., psia | 347.0000 | 385.0000 | 385.0000 | 347.0000 |
| Enth., mm BTU/hr | –286.79 | –549.87 | –5,138.1 | –4,938.0 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 53,633.00 | 98,828.41 | 923,480.00 | 923,480.00 |

| Stream no. | 337 | 338 | 339 | 341 |
|---|---|---|---|---|
| Temp., °F | 1,000.0000 | 569.0683 | 240.1082 | 569.0683 |
| Pres., psia | 347.0000 | 58.0000 | 25.0000 | 58.0000 |
| Enth., mm BTU/hr | –5,224.8 | –5,427.2 | –1,076.1 | –1,799.7 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.00000 | 1.0000 |
| Total, lb/hr | 977,113.00 | 977,113.00 | 161,529.19 | 324,028.13 |

| Stream no. | 342 | 343 | 346 | 347 |
|---|---|---|---|---|
| Temp., °F | 569.0683 | 467.1291 | 467.1291 | 101.5784 |
| Pres., psia | 58.0000 | 35.0000 | 35.0000 | 1.0000 |
| Enth., mm BTU/hr | –3,627.4 | –3,658.4 | –3,621.8 | –3,768.6 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 0.93869 |
| Total, lb/hr | 653,085.00 | 653,085.00 | 646,554.00 | 646,554.00 |

| Stream no. | 348 | 349 | 350 | 351 |
|---|---|---|---|---|
| Temp., °F | 80.0000 | 63.0000 | 71.9504 | 71.9816 |
| Pres., psia | 1.0000 | 14.4000 | 14.4000 | 25.0000 |
| Enth., mm BTU/hr | –4,411.3 | –3,996.8 | –8,408.1 | –8,408.1 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 646,554.00 | 584,349.19 | 1,230,902.88 | 1,230,902.88 |

| Stream no. | 352 | 353 | 354 | 355 |
|---|---|---|---|---|
| Temp., °F | 240.0000 | 77.6091 | 240.1082 | 240.1082 |
| Pres., psia | 27.0000 | 25.0000 | 25.0000 | 25.0000 |
| Enth., mm BTU/hr | –43.509 | –8,401.2 | –49.681 | –629.95 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 1.0000 | 0.00000 |
| Total, lb/hr | 6,530.84 | 1,230,902.88 | 8,699.87 | 94,559.00 |

| Stream no. | 356 | 357 | 358 | 359 |
|---|---|---|---|---|
| Temp., °F | 242.7635 | 242.7635 | 481.7281 | 242.7635 |
| Pres., psia | 841.0000 | 841.0000 | 260.0000 | 841.0000 |
| Enth., mm BTU/hr | –629.70 | –432.63 | –365.09 | –197.07 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 1.0000 | 0.00000 |
| Total, lb/hr | 94,559.00 | 64,966.57 | 64,966.57 | 29,592.43 |

| Stream no. | 360 | 361 | 362 | 363 |
|---|---|---|---|---|
| Temp., °F | 284.3000 | 284.3000 | 284.3000 | 241.3838 |
| Pres., psia | 741.0000 | 741.0000 | 741.0000 | 417.0000 |
| Enth., mm BTU/hr | –195.81 | –85.902 | –109.91 | –1,075.9 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 29,592.43 | 12,982.20 | 16,610.23 | 161,529.19 |

| Stream no. | 364 | 365 | 366 | 367 |
|---|---|---|---|---|
| Temp., °F | 241.3838 | 331.4878 | 241.3838 | 331.4878 |
| Pres., psia | 417.0000 | 105.0000 | 417.0000 | 105.0000 |
| Enth., mm BTU/hr | –340.26 | –296.87 | –323.95 | –283.55 |
| Vapor, mass fraction | 0.0000 | 0.85495 | 0.00000 | 0.83394 |
| Total, lb/hr | 51,084.00 | 51,084.00 | 48,636.50 | 48,636.50 |

| Stream no. | 368 | 369 | 370 | 371 |
|---|---|---|---|---|
| Temp., °F | 241.3838 | 331.4878 | 331.4878 | 241.3838 |
| Pres., psia | 417.0000 | 105.0000 | 105.0000 | 417.0000 |
| Enth., mm BTU/hr | –319.24 | –287.78 | –868.20 | –92.449 |
| Vapor, mass fraction | 0.00000 | 0.63684 | 0.77723 | 0.00000 |
| Total, lb/hr | 47,928.90 | 47,928.90 | 147,649.39 | 13,879.80 |

| Stream no. | 372 | 373 | 374 | 375 |
|---|---|---|---|---|
| Temp., °F | 951.8254 | 951.8254 | 407.9199 | 951.8254 |
| Pres., psia | 870.0000 | 870.0000 | 270.0000 | 870.0000 |
| Enth., mm BTU/hr | –567.52 | –283.76 | –324.65 | –283.76 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.39199 | 1.0000 |
| Total, lb/hr | 105,308.85 | 52,654.43 | 52,654.43 | 52,654.43 |

| Stream no. | 376 | 377 | 378 | 379 |
|---|---|---|---|---|
| Temp., °F | 407.9199 | 406.2503 | 404.5561 | 404.5561 |
| Pres., psia | 270.0000 | 265.0000 | 260.0000 | 260.0000 |
| Enth., mm BTU/hr | –326.88 | –651.53 | –431.68 | –219.91 |
| Vapor, mass fraction | 0.34012 | 0.36757 | 0.00000 | 1.0000 |
| Total, lb/hr | 52,654.43 | 105,308.84 | 66,520.34 | 38,788.51 |

| | | | | |
|---|---|---|---|---|
| Stream °F | 404.5561 | 404.5561 | 404.5561 | 300.0000 |
| Pres., psia | 260.0000 | 260.0000 | 260.0000 | 67.0000 |
| Enth., mm BTU/hr | –691.02 | –37.315 | –653.71 | 0.00000 |
| Vapor, mass fraction | 0.91279 | 0.91279 | 0.91279 | 1.0000 |
| Total, lb/hr | 120,365.32 | 6,499.73 | 113,865.59 | 0.00 |

| Stream no. | 384 | 385 | 387 | 388 |
|---|---|---|---|---|
| Temp., °F | 300.0941 | 300.0000 | 300.0000 | 497.4000 |
| Pres., psia | 67.0000 | 67.0000 | 67.0000 | 665.0000 |
| Enth., mm BTU/hr | –2,883.6 | –1,190.7 | –725.5 | –106.02 |
| Vapor, mass fraction | 0.97986 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 505,039.22 | 180,386.00 | 109,913.30 | 16,610.23 |

| Stream no. | 389 | 390 | 391 | 392 |
|---|---|---|---|---|
| Temp., °F | 404.5561 | 300.0000 | 267.0836 | 242.0000 |
| Pres., psia | 260.0000 | 67.0000 | 855.0000 | 67.0000 |
| Enth., mm BTU/hr | –106.02 | –465.17 | –1,213.4 | –748.67 |
| Vapor, mass fraction | 0.12990 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 16,610.23 | 70,472.64 | 182,883.7 | 112,411.13 |

| Stream no. | 393 | 394 | 395 | 396 |
|---|---|---|---|---|
| Temp., °F | 242.0000 | 242.0000 | 264.4763 | 467.1291 |
| Pres., psia | 67.0000 | 67.0000 | 67.0000 | 35.0000 |
| Enth., mm BTU/hr | –3,363.6 | –2,614.9 | –1,213.8 | –36.584 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 1.0000 |
| Total, lb/hr | 505,039.22 | 392,628.09 | 182,883.77 | 6,530.84 |

| Stream no. | 397 | 398 | 399 |
|---|---|---|---|
| Temp., °F | 80.0000 | 240.0000 | 240.1082 |
| Pres., psia | 14.0000 | 35.0000 | 25.0000 |
| Enth., mm BTU/hr | –4,411.3 | –732.26 | –8,145.6 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.046730 |
| Total, lb/hr | 646,554.00 | 109,913.30 | 1,230,902.88 |

## Table A4.4. PRB coal

| Stream no. | 159 | 300 | 301 | 302 |
|---|---|---|---|---|
| Temp., °F | 1,106.4220 | 240.1082 | 247.5038 | 620.0000 |
| Pres., psia | 15.2000 | 25.0000 | 2,300.0000 | 2,000.0000 |
| Enth., mm BTU/hr | –3,504.2 | –9,714.1 | –9,703.1 | –9,074.5 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 5,437,207.50 | 1,458,127.00 | 1,458,127.00 | 1,458,127.00 |

| Stream no. | 303 | 304 | 305 | 306 |
|---|---|---|---|---|
| Temp., °F | 620.0000 | 632.2300 | 620.0000 | 632.2850 |
| Pres., psia | 2,000.0000 | 1,950.0000 | 2,000.0000 | 1,950.0000 |
| Enth., mm BTU/hr | –6,402.5 | –5,886.1 | –364.86 | –335.43 |
| Vapor, mass fraction | 0.00000 | 1.0000 | 0.00000 | 1.0000 |
| Total, lb/hr | 1,028,788.00 | 1,028,788.00 | 58,628.00 | 58,628.00 |

| Stream no. | 307 | 308 | 309 | 310 |
|---|---|---|---|---|
| Temp., °F | 620.0000 | 632.2000 | 631.9833 | 1,004.0000 |
| Pres., psia | 2,000.0000 | 2,000.0000 | 1,950.0000 | 1,840.0000 |
| Enth., mm BTU/hr | –2,307.1 | –2,300.6 | –8,522.1 | –7,858.0 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.74661 | 1.0000 |
| Total, lb/hr | 370,711.03 | 370,711.03 | 1,458,127.00 | 1,458,127.00 |

| Stream no. | 312 | 313 | 314 | 315 |
|---|---|---|---|---|
| Temp., °F | 1,106.4220 | 668.4714 | 683.9567 | 273.7841 |
| Pres., psia | 15.2000 | 15.0000 | 15.0000 | 14.8000 |
| Enth., mm BTU/hr | –3,293.9 | –3,958.1 | –4,186.4 | –4,815.1 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 5,110,975.00 | 5,110,975.00 | 5,437,207.50 | 5,437,207.50 |

| Stream no. | 316 | 317 | 318 | 319 |
|---|---|---|---|---|
| Temp., °F | 1,106.4220 | 1,106.4220 | 886.5415 | 922.5463 |
| Pres., psia | 15.2000 | 15.2000 | 15.2000 | 15.0000 |
| Enth., mm BTU/hr | –210.25 | –63.075 | –69.552 | –228.35 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 326,232.44 | 97,869.73 | 97,869.73 | 326,232.44 |

| Stream no. | 320 | 321 | 323 | 324 |
|---|---|---|---|---|
| Temp., °F | 1,106.4220 | 937.9244 | 1,004.0000 | 1,004.0000 |
| Pres., psia | 15.2000 | 15.0000 | 1,840.0000 | 1,840.0000 |
| Enth., mm BTU/hr | –147.17 | –158.80 | –567.52 | –7,290.4 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 228,362.69 | 228,362.69 | 105,308.85 | 1,352,818.25 |

| Stream no. | 325 | 326 | 327 | 328 |
|---|---|---|---|---|
| Temp., °F | 772.5801 | 772.5801 | 240.1082 | 242.1404 |
| Pres., psia | 800.0000 | 800.0000 | 25.0000 | 650.0000 |
| Enth., mm BTU/hr | –7,425.0 | –1,519.7 | –113.65 | –113.62 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.00000 | 0.00000 |
| Total, lb/hr | 1,352,818.25 | 276,877.25 | 17,060.00 | 17,060.00 |

| Stream no. | 329 | 330 | 331 | 332 |
|---|---|---|---|---|
| Temp., °F | 645.0892 | 772.5801 | 598.5389 | 598.5389 |
| Pres., psia | 650.0000 | 800.0000 | 385.0000 | 385.0000 |
| Enth., mm BTU/hr | –1,633.3 | –5,905.4 | –5,986.4 | –298.41 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 293,937.25 | 1,075,941.00 | 1,075,941.00 | 53,633.00 |

| Stream no. | 333 | 334 | 335 | 336 |
|---|---|---|---|---|
| Temp., °F | 1,000.0000 | 598.5389 | 598.5389 | 1,000.0000 |
| Pres., psia | 347.0000 | 385.0000 | 385.0000 | 347.0000 |
| Enth., mm BTU/hr | –286.79 | –549.87 | –5,138.1 | –4,938.0 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
| Total, lb/hr | 53,633.00 | 98,828.41 | 923,480.00 | 923,480.00 |

| Stream no. | 337 | 338 | 339 | 341 |
|---|---|---|---|---|
| Temp., °F | 1,000.0000 | 569.0683 | 240.1082 | 569.0683 |
| Pres., psia | 347.0000 | 58.0000 | 25.0000 | 58.0000 |
| Enth., mm BTU/hr | –5,224.8 | –5,427.2 | –1,075.5 | –1,799.7 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.00000 | 1.0000 |
| Total, lb/hr | 977,113.00 | 977,113.00 | 161,444.00 | 324,028.13 |

| Stream no. | 342 | 343 | 346 | 347 |
|---|---|---|---|---|
| Temp., °F | 569.0683 | 467.1291 | 467.1291 | 101.5784 |
| Pres., psia | 58.0000 | 35.0000 | 35.0000 | 1.0000 |
| Enth., mm BTU/hr | –3,627.4 | –3,658.4 | –3,621.8 | –3,768.6 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 1.0000 | 0.93869 |
| Total, lb/hr | 653,085.00 | 653,085.00 | 646,554.00 | 646,554.00 |

| Stream no. | 348 | 349 | 350 | 351 |
|---|---|---|---|---|
| Temp., °F | 80.0000 | 63.0000 | 71.9504 | 71.9816 |
| Pres., psia | 1.0000 | 14.4000 | 14.4000 | 25.0000 |
| Enth., mm BTU/hr | –4,411.3 | –3,996.8 | –8,408.1 | –8,408.1 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 646,554.00 | 584,349.19 | 1,230,902.88 | 1,230,902.88 |

| Stream no. | 352 | 353 | 354 | 355 |
|---|---|---|---|---|
| Temp., °F | 240.0000 | 77.6091 | 240.1082 | 240.1082 |
| Pres., psia | 27.0000 | 25.0000 | 25.0000 | 25.0000 |
| Enth., mm BTU/hr | –43.509 | –8,401.2 | –49.679 | –629.95 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 1.0000 | 0.00000 |
| Total, lb/hr | 6,530.84 | 1,230,902.88 | 8,699.44 | 94,559.00 |

| Stream no. | 356 | 357 | 358 | 359 |
|---|---|---|---|---|
| Temp., °F | 242.7635 | 242.7635 | 404.5561 | 242.7635 |
| Pres., psia | 841.0000 | 841.0000 | 260.0000 | 841.0000 |
| Enth., mm BTU/hr | –629.70 | –432.63 | –368.35 | –197.07 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.99974 | 0.00000 |
| Total, lb/hr | 94,559.00 | 64,966.57 | 64,966.57 | 29,592.43 |

| Stream no. | 360 | 361 | 362 | 363 |
|---|---|---|---|---|
| Temp., °F | 284.3000 | 284.3000 | 284.3000 | 241.3838 |
| Pres., psia | 741.0000 | 741.0000 | 741.0000 | 417.0000 |
| Enth., mm BTU/hr | –195.81 | –85.902 | –109.91 | –1,075.3 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 29,592.43 | 12,982.20 | 16,610.23 | 161,444.00 |

| Stream no. | 364 | 365 | 366 | 367 |
|---|---|---|---|---|
| Temp., °F | 241.3838 | 880.3931 | 241.3838 | 331.4878 |
| Pres., psia | 417.0000 | 105.0000 | 417.0000 | 105.0000 |
| Enth., mm BTU/hr | –406.30 | –329.50 | –257.35 | –226.02 |
| Vapor, mass fraction | 0.00000 | 1.0000 | 0.00000 | 0.81147 |
| Total, lb/hr | 60,998.75 | 60,998.75 | 38,636.50 | 38,636.50 |

| Stream no. | 368 | 369 | 370 | 371 |
|---|---|---|---|---|
| Temp., °F | 241.3838 | 331.4878 | 363.3692 | 241.3838 |
| Pres., psia | 417.0000 | 105.0000 | 105.0000 | 417.0000 |
| Enth., mm BTU/hr | –319.24 | –280.36 | –835.88 | –92.449 |
| Vapor, mass fraction | 0.00000 | 0.81194 | 1.0000 | 0.00000 |
| Total, lb/hr | 47,928.90 | 47,928.90 | 147,564.16 | 13,879.80 |

| Stream no. | 372 | 373 | 374 | 375 |
|---|---|---|---|---|
| Temp., °F | 951.8254 | 951.8254 | 407.9199 | 951.8254 |
| Pres., psia | 870.0000 | 870.0000 | 270.0000 | 870.0000 |
| Enth., mm BTU/hr | –567.52 | –283.76 | –325.51 | –283.76 |
| Vapor, mass fraction | 1.0000 | 1.0000 | 0.37199 | 1.0000 |
| Total, lb/hr | 105,308.85 | 52,654.43 | 52,654.43 | 52,654.43 |

| Stream no. | 376 | 377 | 378 | 379 |
|---|---|---|---|---|
| Temp., °F | 407.9199 | 406.2503 | 404.5561 | 404.5561 |
| Pres., psia | 270.0000 | 265.0000 | 260.0000 | 260.0000 |
| Enth., mm BTU/hr | –327.78 | –653.29 | –445.59 | –207.76 |
| Vapor, mass fraction | 0.31916 | 0.34713 | 0.00000 | 1.0000 |
| Total, lb/hr | 52,654.43 | 105,308.84 | 68,664.27 | 36,644.57 |

| Stream no. | 380 | 381 | 382 | 383 |
|---|---|---|---|---|
| Temp., °F | 404.5561 | 404.5561 | 404.5561 | 300.0000 |
| Pres., psia | 260.0000 | 260.0000 | 260.0000 | 67.0000 |
| Enth., mm BTU/hr | –682.13 | –36.835 | –645.29 | 0.00000 |
| Vapor, mass fraction | 0.87761 | 0.87761 | 0.87761 | 1.0000 |
| Total, lb/hr | 118,221.38 | 6383.95 | 111,837.41 | 0.00 |

| Stream no. | 384 | 385 | 387 | 388 |
|---|---|---|---|---|
| Temp., °F | 383.4183 | 300.0000 | 300.0000 | 497.4000 |
| Pres., psia | 67.0000 | 67.0000 | 67.0000 | 665.0000 |
| Enth., mm BTU/hr | –2,850.8 | –1,191.4 | –725.97 | –106.02 |
| Vapor, mass fraction | 1.0000 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 504,838.19 | 180,501.69 | 109,983.84 | 16,610.23 |

| Stream no. | 389 | 390 | 391 | 392 |
|---|---|---|---|---|
| Temp., °F | 404.5561 | 300.0000 | 267.0836 | 242.0000 |
| Pres., psia | 260.0000 | 67.0000 | 855.0000 | 67.0000 |
| Enth., mm BTU/hr | –106.02 | –465.47 | –1,213.4 | –748.37 |
| Vapor, mass fraction | 0.12990 | 0.00000 | 0.00000 | 0.00000 |
| Total, lb/hr | 16,610.23 | 70,517.86 | 182,884.25 | 112,366.38 |

| Stream no. | 393 | 394 | 395 | 396 |
|---|---|---|---|---|
| Temp., °F | 242.0000 | 242.0000 | 264.4907 | 467.1291 |
| Pres., psia | 67.0000 | 67.0000 | 67.0000 | 35.0000 |
| Enth., mm BTU/hr | –3,362.3 | –2,613.9 | –1,213.8 | –36.584 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.00000 | 1.0000 |
| Total, lb/hr | 504,838.19 | 392,472.00 | 182,884.25 | 6,530.84 |

| Stream no. | 397 | 398 | 399 |
|---|---|---|---|
| Temp., °F | 80.0000 | 240.0000 | 240.1082 |
| Pres., psia | 14.0000 | 35.0000 | 25.0000 |
| Enth., mm BTU/hr | –4,411.3 | –732.73 | –8,145.6 |
| Vapor, mass fraction | 0.00000 | 0.00000 | 0.046730 |
| Total, lb/hr | 646,554.00 | 109,983.84 | 1,230,902.88 |

# Appendix B1. Methodology of Cost Analysis for the PC-Based Power Plant

## 1. Capital Cost

In the capital cost analyses, a PC power plant was divided into the following process sections: (1) the basic plant, (2) the flue gas cleanup units including an FGD and an SCR, (3) the MEA unit, and (4) the auxiliaries in the oxy-combustion process.

### 1.1. Basic PC Plant

There are thirteen basic process areas involved in a sub-critical PC plant. These processes are listed in Table B1.1.

**Table B1.1. Classification of process areas in the sub-critical power plant**

| | Process area | Operating parameter used for cost scaling |
|---|---|---|
| 1 | Coal handling | Coal feed rate |
| 2 | Coal preparation & feed | Coal feed rate |
| 3 | Feed water & miscellaneous | Water feed rate |
| 4 | PC boiler & accessories | Unit gross capacity |
| 5 | Electrostatic precipitator | Flue gas flow rate |
| 6 | HRSG,[1] ducting, and stack | Flue gas flow rate |
| 7 | Steam turbine generator | Superheat and reheat steam flow rates |
| 8 | Cooling water system | Heat duty of cooling water tower |
| 9 | Ash/spent sorbent handling system | Ash in coal feed |
| 10 | Accessory electric plant | Unit gross capacity |
| 11 | Instrumentation & control | Unit gross capacity |
| 12 | Improvements to site | Unit gross capacity |
| 13 | Buildings and structures | Unit gross capacity |

**[1]HRSG, heat recovery steam generator.**

The capital cost of each listed process area is available for a 401-MWe (net) reference PC plant developed in a U.S. DOE-funded study (Parsons Infrastructure & Technology Group, Inc., 2000, 2002). The values reported in the U.S. DOE study were used to scale-up or scale-down the plant sizes investigated in this study. A power rule was used in the cost-scaling on the basis of operating parameters (see Table B1.1) specific to individual process areas. In this study, a uniform scaling exponent of 0.72 was employed for each process area (East Harbor Management Services Ltd., 2002; see also: http://www.med.govt.nz/ers/electric/fossilfuel/fossilfuel.pdf). The formula for cost-scaling is expressed as follows:

$$BEC_{i1} = BEC_{i0} \times (P_{i0} / P_{i1})^{0.72}$$

where

BEC = bare erected cost;
P = specific operating parameter;
i = process area, 1 to 13; and

1 and 0 = plant under study and U.S. DOE reference plant, respectively.

The total plant cost (TPC) includes the following cost items: bare erected cost, engineering and home office, process contingency, and project contingency.

The calculation of the TPC can be expressed according to

$$TPC = \sum_i BEC_i \times (1 + A_i\% + B_i\%) \times (1 + C_i\%)$$

where

A = cost percentage of engineering and home office;
B = cost percentage of process contingency;
C = cost percentage of project contingency; and
i = type of process area.

For each process area of the basic plant, the engineering and home office fees and contingencies are available for the U.S. DOE reference plant.

### 1.2. Flue Gas Cleanup Units

#### *1.2.1. FGD Unit*

*1.2.1.1. LSFO Process.* The limestone slurry forced oxidation (LSFO) process without dibasic acid (DBA) addition was employed for the power plant burning Illinois high-sulfur coal with an $SO_2$ removal efficient of 95%. The LSFO consists of five process areas, and their bare erected costs are described as follows according to the EPA cost modeling approach (Srivastava, 2000).

- Reagent feed area

$$BM_F = -0.0034 \times (FR_L/1{,}000)^4 + 2.1128 \times (FR_L/1{,}000)^3 - 494.55 \times (FR_L/1{,}000)^2 \\ + (68{,}164.7 \times (FR_L/1{,}000) + 7{,}118{,}470 + C_{B\&H}$$

$BM_F$ = bare erected cost of reagent feed area, \$
$FR_L$ = reagent feed rate, lb/hr;
$C_{B\&H}$ = cost of the ball mill and hydroclones, \$, estimated according to

$$C_{B\&H} = 32.9 \times (FR_L / 2{,}000)^2 + 22{,}412 \times (FR_L / 2{,}000) + 1{,}854{,}902$$

The $CaCO_3$ content of limestone was assumed at 95.3%, and the molar ratio of $CaCO_3$ to $SO_2$ was assumed as 1.05.

- $SO_2$ removal area

The bare erected cost is composed of the cost of absorbers, pumps, and auxiliary facilities.

$$BM_R = Absorbers + Pumps + Auxiliary$$

$$Absorbers = 173{,}978 \times (ACFM / 1{,}000 / N_a)^{0.5575} \times N_a$$

$$Pumps = 910.85 \times (F_{GPM} / N_p)^{0.5954} \times N_p$$

$$Auxiliary = 0.8701 \times (FR_{SO_2} / 1{,}000)^3 \text{ - } 188.2 \times (FR_{SO_2} / 1{,}000)^2 + 34{,}809 \times (FR_{SO_2} / 1{,}000) + 1{,}905{,}302$$

where

$BM_R$ = bare erected cost of $SO_2$ removal area, \$;
$FR_{SO_2}$ = $SO_2$ feed rate, lb/hr;
$F_{GPM}$ = slurry flow rate, gpm; based on L/G = 125 gal/1,000 $ft^{3;}$
ACFM = flue gas flow rate into the absorber, acfm;
$N_a$ = number of absorbers; maximum absorber size was limited to treat 700 MWe; larger size required multiple, equal size absorbers; and
$N_p$ = number of pumps; a maximum single pump was limited to treat 43,000 gpm.

- Flue gas handling area

$$BM_G = ID \quad Fans + Auxiliary$$

$$ID \quad Fans = 91.24(ACFM / N_f)^{0.6842} \times N_f$$

$$Auxiliary = -0.1195 \times (\frac{ACFM}{1{,}000})^2 + 777.76 \times (\frac{ACFM}{1{,}000}) + 238{,}203 + 0.000012 \times (\frac{ACFM}{1{,}000})^3 - 0.1651 \times (\frac{ACFM}{1{,}000})^2 + 1{,}288.82 \times (\frac{ACFM}{1{,}000}) + 559{,}693 - 0.2009 \times (\frac{ACFM1}{1{,}000 N_a})^2 + 1{,}266.4 \times (\frac{ACFM1}{1{,}000 N_a}) + 420{,}141$$

where

$BM_G$ = bare erected cost of flue gas handling area, \$;
ACFM1 = flue gas flow rate out of absorber, acfm; and
$N_a$ = number of ID fans (assumed to be in groups of 2,4, or 8 with a maximum individual fan capacity of 1.6 million cfm).

- Waste/by-product handling area

$$BM_W = Thickener + disposal$$

$$Thickener = 9{,}018\,.7 \times FR_{SO_2} \times 0.95 \times \frac{172}{64 \times 2{,}000} + 114{,}562$$

$$Disposal = 0.325 \times (FR_{SO_2} / 1{,}000)^3 - 168.77 \times (FR_{SO_2} / 1{,}000)^2 + 29{,}091 \times (FR_{SO_2} / 1{,}000) + 773{,}243$$

where

$BM_W$ = bare erected cost of waste/by-product handling area, \$.

- Support equipment area

$$BM_E = Chimney + auxiliary$$
$$Chimney = 23{,}370 \times ACFM1^{0.3339}$$
$$Auxiliary = 0.0003 \times MW_e^{\,3} - 1.0677 \times MW_e^{\,2} + 1{,}993.8 \times MW_e + 1{,}177{,}674$$

where

$BM_E$ = bare erected cost of support equipment area, \$.

The cost of chimney was based on no flue gas reheat.

*1.2.1.2. LSD Process* The lime spray dryer (LSD) process was employed for the power plant burning the PRB low-sulfur coal with an $SO_2$ removal efficiency of 90%. The cost estimation was also referred to the EPA approach (Srivastava, 2000). The bare erected costs were based on five process areas and described as follows.

- Reagent feed area (BMF)

$$BM_F = (170{,}023 \times \frac{FR_L}{1{,}000} + 376{,}411) + (72{,}338 \times GPM^{0.3195})$$

where

$FR_L$ = reagent feed rate, lb/hr; lime purity was assumed as 90%; and molar ratio of CaO to $SO_2$ was assumed as 1.75; and
GPM = slurry flow rate, gal/min; lime content in slurry was assumed as 30%.

- $SO_2$ removal area ($BM_R$)

$$BM_R = Spray \quad Dryers + Auxiliary$$
$$Spray \quad Dryers = N_a \times \left( -3.57 \times (\frac{ACFM}{N_a \cdot 1{,}000})^2 + 9{,}246 \times (\frac{ACFM}{N_a \cdot 1{,}000}) + 791{,}896 \right)$$
$$Auxiliary = N_a \times \left( \begin{array}{l} 581{,}877{,}809 \times Wt\%S^3 - 3{,}653{,}117 \times Wt\%S^2 \\ + 693{,}335 \times Wt\%S + 214{,}198 \end{array} \right) + 677{,}421 \times Wt\%S^{-0.0966}$$

where

$N_a$ = number of absorbers; single absorber was limited to a maximum 300 MWe; larger unit used multiple, equal-sized absorbers;
ACFM = flue gas flow rate into the absorber, cfm; and
Wt%S = coal sulfur content.

- Flue gas handling area ($BM_G$)

$$BM_G = Auxiliary + ID \quad Fans$$

$$Auxuliary = N_a \times \begin{pmatrix} 1{,}721.8 \times (\frac{ACFM}{1{,}000})^{0.683} + \\ 1{,}326.2 \times (\frac{ACFM1}{1{,}000})^{0.7131} \end{pmatrix} + \begin{pmatrix} 15{,}338 \times (\frac{ACFM}{1{,}000})^{0.5} + \\ 47{,}680 \times (\frac{ACFM1}{1{,}000})^{0.5576} \end{pmatrix} + \begin{pmatrix} 4{,}840.4 \times (\frac{ACFM2}{1{,}000})^{0.5} + \\ 2{,}695.9 \times (\frac{ACFM3}{1{,}000})^{0.5} \end{pmatrix}$$

$$ID \quad Fans = 91.24 \times (\frac{ACFM}{N_f})^{0.6842} \times N_f$$

where

ACFM = flue gas rate at the absorber inlet, cfm;
ACFM1 = flue gas rate at the absorber exit, cfm;
ACFM2 = flue gas rate at the particulate control device inlet, cfm;
ACFM3 = flue gas rate at the ID fan exit, cfm; and
$N_f$ = number of fans; the maximum flow rate of a single fan was assumed as 1.6 million cfm.

- Waste/by-product handling area ($BM_W$)

$$BM_W = 2{,}051{,}841{,}884 \times Wt\%S^2 - 1{,}443{,}163 \times Wt\%S + 1{,}026{,}479$$

- Support equipment area (BME)

$$BM_E = -1.211 \times MWe^2 + 2{,}704.2 \times MWe + 1{,}354{,}716.2 + Chimney$$
$$Chimney = 23{,}370 \times ACFM3^{0.3908}$$

According to the EPA approach, both for the LSFO process and the LSD process, 5% for general facilities, 10% for engineering and home office, 5% for process contingency, 15% for project contingency, and 3% for contractor's fee were applied to estimate the total plant cost.

$$TPC_{FGD} = \sum_i BM_i \times (1 + 5\% + 10\% + 5\%) \times (1 + 15\%) \times (1 + 3\%)$$

where

$BM_i$ = bare erected cost of process area i.

### *1.2.2. SCR Unit*

The cost model for SCR retrofit developed in an EPA-funded project was used in this study (Forter and Jozewicz, 2001). The model could be applied to coal-fired boilers with capacity of 100 MW to 850 MW and with $NO_x$ inlet level of 0.15 to 2.5 lb/mm BTU. The cost model is expressed as follows:

$$D = 75 \times \left( \frac{300{,}000}{A} \left[ (\frac{B}{1.5})^{0.05} (\frac{C}{100})^{0.4} \right] \right)^{0.35}$$

where

D = total capital cost of SCR, \$/kW;
A = plant size, kW;
B = $NO_x$ level at SCR inlet, lb/mm BTU;
C = $NO_x$ removal efficiency (assumed to be 90% in this study).

This cost accounts for the total capital requirement (TCR) except for interest and escalation during construction. The total plant cost (TPC) was assumed 10% less than TCR for the SCR unit.

The retrofit model could also be used for SCR installation in a new coal-fired boiler. A new SCR application was reported as 20 to 50% less than a retrofit, and 30% was adopted in this study.

### 1.3. MEA Unit

The capital cost of the MEA unit was estimated by a scaling approach. The reference is based on an MEA unit installed for a 400 MW (net) PC plant in a U.S. DOE study (Parsons Infrastructure & Technology Group, Inc., 2002).

Several important design parameters determine the economics of process scaling, including flue gas conditions, $CO_2$ removal efficiency, solvent flow conditions, and heat duty for solvent regeneration. Based on the design of the Fluor Daniel Econamine FG process, a $CO_2$ removal efficiency of 90% was assumed. In addition, typical lean and rich solvent conditions were chosen similar to the reference MEA unit.

Because flue gas conditions such as the $CO_2$ concentration and temperature are comparable for different PC plants burning either high-sulfur coal or PRB coal, the flow rate of the flue gas was chosen as the only basis for cost-scaling in this study. A power rule was assumed for cost-scaling with an exponent of 0.65. The cost-scaling was expressed as

$$TPC_{MEA} = TPC_{ref} \times \left( \frac{GF}{GF_{ref}} \right)^{0.65}$$

where

$TPC_{ref}$ = total plant cost of MEA unit in the reference plant, \$;
GF = gas flow rate into MEA in the studied plant, lb/hr; and
$GF_{ref}$ = gas flow rate into MEA in the reference plant, lb/hr.

### 1.4. Oxy-Combustion Plant

In addition to the common components described in sections 1.1 and 1.2, additional process areas specific to the oxy-combustion process are described as follows. These include the ASU, flue gas cooling, water vapor condenser, $CO_2$ compression, cryogenic unit, and low temperature flash.

#### *1.4.1. ASU*

The capital cost for the ASU was provided by the American Air Liquide. The bare erected costs for different scale and $O_2$ purity are listed in Table B1.2. The unit cost in terms of \$1,000/(t/day) varies with the scale of oxygen production, but is almost comparable for 95% and 99% $O_2$ purity.

## Table B1.2. Capital cost of oxygen generation

| Oxygen production | | 95% $O_2$ Purity | | 99 % $O_2$ Purity | |
|---|---|---|---|---|---|
| lb/hr | tonnes /day | $1,000 | $1,000/ (tonnes/day) | $1,000 | $1,000/ (tonnes/day) |
| 399,248 | 4,346 | 44,424 | 10.22 | 45,535 | 10.48 |
| 415,450 | 4,524 | 45,745 | 10.11 | 46,888 | 10.36 |
| 851,275 | 9,267 | 80,904 | 8.73 | 82,918 | 8.95 |
| 885,740 | 9,642 | 83,716 | 8.68 | 85,767 | 8.9 |
| 1,597,138 | 17,386 | 141,171 | 8.12 | 144,596 | 8.32 |
| 1,661,801 | 18,090 | 146,395 | 8.09 | 149,944 | 8.29 |

The total plant cost was estimated based on the assumptions of 5% engineering and home office fees and 10% project contingency:

$$TPC_{O_2} = BM_{O_2} \times (1+5\%) \times (1+10\%)$$

### *1.4.2. Flue Gas Cooling*

A water spray cooling system is employed for flue gas cooling prior to the FGD process. The capital cost was estimated as follows (USEPA, 2002):

$$BM_{COOL} = 6{,}025 \times 1{,}000 \times \left(\frac{GPM}{215}\right)^{0.65}$$

$$TPC_{COOL} = 1.3725 \times BM_{COOL}$$

where

$BM_{cool}$= bare erected cost of the cooling system, $;
TPC = total plant cost, $; and
GPM = water consumption rate, gal/min.

### *1.4.3. Water Vapor Condenser*

The capital cost of the flue gas condenser was based on that reported for an 865-MW plant (Anderson and Maksinen, 2002). The capital cost is scaled-up according to the following power rule.

$$TPC_{conds} = C_{ref} \times \left(\frac{FG}{FG_{ref}}\right)^{0.8} \left(\frac{m_W}{m_{Wref}}\right)^{0.8}$$

where

$C_{ref}$ = capital cost of the condenser in literature, $;
FG = flue gas flow rate, cfm; and
$m_W$ = cooling water consumption, lb/h.

### *1.4.4. Cryogenic Unit*

The cryogenic unit cools the compressed $CO_2$ stream to –33°C, and the cooled stream is flashed to remove the gas components, such as oxygen, argon, and nitrogen. The final product is a liquefied $CO_2$ (>99% purity). The process consists of a compressor and an expander for the ammonia refrigeration cycle, heat exchangers, and a low-temperature flash.

Equipment costs were calculated using individual cost models built in CHEMCAD. An installation factor of 1.8 was assumed to cover the costs related to installation materials and labors. Engineering and home office was assumed at 5% of the bare erected cost. Project contingency was assumed at 10%; no process contingency was considered for these mature technologies. Total plant cost can be expressed as follows:

$$TPC_{cryogenic} = \sum_{i} \sum EC_i \times (1 + 80\%) \times (1 + 5\%) \times (1 + 10\%)$$

where

EC = equipment cost; and
i = type of equipment.

This cost estimation was conducted for the total plant cost (TPC), which consisted of bare erected cost, engineering and home office overheads and fee, and contingencies. Total plant investment (TPI) is calculated by adding the escalation of construction costs and interest during construction (Table 3) to the TPC:

$$TPI = TPC \cdot A \cdot \frac{\left[ [(1+i)/(1+ea)]^n - 1 \right]}{(1+i)/(1+ea)}$$

where

A = cost expended per year, 1/n;
n = construction years;
i = weighted cost of capital (discount rate); and
ea = inflation rate.

**Table B1.3. Items of capital cost estimation**

| | |
|---|---|
| (a) | Bare erected cost |
| (b) | Engineering cost |
| (c) | Process contingency |
| (d) | Project contingency |
| | Total plant cost (TPC) = a + b + c +d |
| (e) | Total cash expended |
| (f) | AFDC (escalation and interest in construction period) |
| | Total plant investment (TPI)= TPC + e + f |
| (g) | Royal allowance |
| (h) | Pre-production costs |
| (i) | Inventory capital |
| (j) | Initial catalyst & chemicals |
| (k) | Land cost |
| | Total capital requirement (TCR) = TPI + g+ h + i + j + k |

The total capital requirement (TCR) covers all other expenditure to complete an entire plant. These include the pre-production, spare parts, inventory capital, initial chemicals, and land costs. The assumptions used for calculating the TCR are listed in Table B1.4.

**Table B1.4. Assumptions for estimating the total capital requirement (TCR)**

| | |
|---|---|
| Prepaid royalties | 0% |
| Pre-production | |
| O&M cost (except for fuel) | 30 days |
| Fuel | 7.5 days |
| Plant modifications | 2% of total plant investment (TPI) |
| Inventory capital | |
| Coal | 60 days |
| Consumables (excluding water) | 60 days |
| Spare parts | 0.5% of total plant cost (TPC) |
| Initial catalyst and chemical inventory | 30 days |
| Land | $6,500/acre |

# 2. O&M Cost

## 2.1. Fixed O&M Cost

The fixed O&M cost includes the cost of operating labor, maintenance material and labor, and administrative and support labor (Table B1.5). The cost of operating labor (OL) was estimated based on the number of operating jobs (OJ) and the labor rate.

$$OL = Labor\ rate \times OJ \times 40\ hrs/week \times 52\ weeks/yr$$

Annual maintenance labor and material costs (ML&M) were estimated as a percentage of the total plant cost (TPC). The percentage varies with individual process areas. It was assumed that 40% of the ML&M is shared by the maintenance labor.

$$ML\,\&\,M = \sum_i F_i \cdot BM_i$$

where

F = maintenance factor;
BM = bare erected cost; and
i = process area or equipment.

Administrative and support labor cost (A&S) was assumed to be 30% of the sum of operating labor cost (OL) and maintenance labor (ML&M) cost:

$$A\,\&\,S = 30\% \times (OL + 40\% \times ML\,\&\,M)$$

### Table B1.5. Estimation of the fixed O&M cost[1]

| Fixed O&M cost | Assumptions |
|---|---|
| Operating labor | Labor rate: $40/hr per person<br>Operating jobs:<br>1. Basic plant: 25 persons/shift[3]<br>2. Flue gas cleanup: FGD[1] and SCR[2]<br>3. MEA unit: 2 persons/shift<br>4. Oxy-combustion: 2 persons/shift for ASU, 2 persons/shift for others |
| Maintenance | % of TPC:<br>1. Basic plant[3]<br>2. Flue gas cleanup: FGD[1] and SCR[2]<br>3. MEA unit: 2.5%<br>4. Oxy-combustion: 2.5% both for ASU and others |
| Administrative & support cost | 30% of O&M labor |

[1] Srivastava, 2000.
[2] Foerter and Jozewicz, 2001.
[3] Gilbert/Commonwealth Inc., 1995.

## 2.2. Variable O&M Cost

The variable O&M cost includes the costs of consumables (chemicals, water, and waste disposal) and fuel. The unit price used in this study is listed in Table B1.6. A 70% loading factor of the power plant is assumed for variable O&M cost.

### Table B1.6. Unit prices of consumables and coals

| Category | Unit price |
|---|---|
| Coal | 20 $/t for PRB coal, 30 $/t for Illinois coal |
| Water | 0.06 $/t |
| Waste waster treatment chemicals | 0.13 $/lb |
| Ash disposal | 10 $/t |
| Limestone | 15 $/t |
| Lime | 65 $/t |
| Steam | 7 $/t |
| Ammonia | 225 $/t |
| SCR catalyst | Published models[1] |
| MEA | $1250/tonne |
| MEA inhibitor | N/A (20% of the total MEA cost assumed) |
| Waste disposal from MEA unit | $175/tonne |
| Caustic sodium | $2,000/tonne |
| Activated carbon | $2,000/tonne |

[1] Foerter and Jozewicz, 2001.

## 3. Cost of Electricity

The cost of electricity (COE) is the levelized (over the plant life) coal pile-to-busbar cost of power expressed in mills per kilowatt-hour.

The financial criteria used in this study (Table B1.7) are from the literature (Parsons Infrastructure & Technology Group Inc., 2002). Based on these assumptions, a levelized carrying charge (applied to TCR) is 0.138. No inflation was considered for categories in the O& M costs.

**Table B1.7. Financial assumptions for the cost estimation**

| | | |
|---|---|---|
| **General characteristics** | | |
| Plant startup date (year) | 2000 | |
| Capital cost year dollars | 2000 | |
| Construction period | 3 years | |
| Capacity factor | 70% | |
| **Financial assumptions** | | |
| Project book life | 20 years | |
| Book salvage value | 0% | |
| Project tax life | 20 years | |
| Tax depreciation method | ACRS class | |
| Property tax rate | 1% per year | |
| Insurance tax rate | 1% per year | |
| Federal income tax rate | 34% | |
| State income tax rate | 4.2% | |
| Investment tax credit% eligible | 0% | |
| **Capital structure** | **% of total** | **Cost, %** |
| Common equity | 42 | 12 |
| Preferred stock | 10 | 8.5 |
| Debt | 45 | 9.0 |
| Weighted cost of capital (after tax) | 8.76% | |
| **Escalation rates (apparent)** | | |
| General escalation | 0% | |
| Primary/secondary fuel price escalation | 0% | |

A levelized busbar cost of electricity for the 70% design capacity factor is calculated using the following relationship:

$$COE = \frac{13.8\% \times TCR + 1.541 \times O\,\&\,M\ cost}{MW_{net} \times 0.7 \times 8{,}760}$$

where

COE = cost of electricity, mills/kWh;
$MW_{net}$ = new electricity output, MW;
TCR = total capital cost, $; and
O&M cost = operating and maintenance cost, $/yr.

## 4. Costs of $CO_2$ Avoidance and $CO_2$ Capture

$CO_2$ costs can be expressed in terms of either the cost per ton $CO_2$ removed or the cost per ton $CO_2$ avoided. The cost of $CO_2$ capture is defined as the increase in cost of electricity per captured $CO_2$ emissions due to installation of a $CO_2$ capture process. It can be expressed as follows

$$\text{Cost of CO}_2\text{ capture (\$/ton)} = \frac{(\$/kWh)_{capture} - (\$/kWh)_{reference}}{(ton\ CO_2/kWh)_{capture}}$$

where

capture = the power plant with $CO_2$ capture facility;
reference = air-blown PC plant without $CO_2$ capture;
\$/kWh = levelized COE; and
ton $CO_2$/kWh = $CO_2$ emissions per kWh generation.

Because a $CO_2$ capture unit consumes considerable electricity and thus reduces the plant output, the $CO_2$ emissions per net kilowatt-hour generation increase correspondingly. The actual avoided emissions are the difference between $CO_2$ emissions before and after the installation of the $CO_2$ capture unit. This relationship can be expressed as

$$\text{Cost of CO}_2\text{ avoidance (\$/ton)} = \frac{(\$/kWh)_{capture} - (\$/kWh)_{reference}}{(ton\ CO_2/kWh)_{reference} - (ton\ CO_2/kWh)_{capture}}$$

# Appendix B2. Methodology of Cost Analysis for the IGCC Plant with Selexol Unit

## 1. Capital Cost

The IGCC plant was categorized into sixteen process areas for the cost estimation (Table B2.1). The bare erected cost for each process area in this study was scaled from an IGCC case study developed in a U.S. DOE/EPRI project (Parsons Infrastucture & Technology Group Inc., 2002). The cost scaling was based on key parameters specific to individual process areas as listed in the last column of Table B2.1. A power law with an exponent of 0.72 was also assumed for the IGCC plant.

**Table B2.1. Classification of process areas in the IGCC plant with the Selexol unit**

| | Process area | Operating parameter for cost scaling |
|---|---|---|
| 1 | Coal handling | Coal feed rate |
| 2 | Coal preparation & feed | Coal feed rate |
| 3 | Feed water & miscellaneous Bop systems | Water feed rate |
| 4 | Gasifier & accessories | Coal slurry feed rate |
| 5 | ASU and accessories | Oxygen feed rate |
| 6 | Gas cleanup and piping | Fuel gas flow rate |
| 7 | CO2 compression | $CO_2$ captured |
| 8 | Gas turbine and accessories | Gross capacity |
| 9 | HRSG, ducting and stack | Flue gas flow rate |
| 10 | Steam turbine generator | Steam flow rates |
| 11 | Cooling water system | Heat duty of cooling water tower |
| 12 | Ash /spent sorbent handling system | Ash in coal feed |
| 13 | Accessory electric plant | Gross capacity |
| 14 | Instrumentation &control | Gross capacity |
| 15 | Improvements to site | Gross capacity |
| 16 | Buildings and structures | Gross capacity |

Based on the bare erected costs of individual process areas, the total plant cost (TPC), the total plant investment (TPI), and the total capital requirement (TCR) are estimated according to the assumptions listed in Table B2.2.

The construction period adopted for the IGCC plant with Selexol unit is assumed to be 4 years. The royalty allowance is considered as 0.5% of the BEC, given the fact that the IGCC process is relatively less mature than the conventional PC plant.

## Table B2.2. Estimation of TPC, TPI, and TCR

| | Items | Assumptions |
|---|---|---|
| (a) | Bare erected cost (BEC) | $\sum BEC_i$, i = 1,2…,16 |
| (b) | Engineering cost | 10% of BEC |
| (c) | Process contingency | 5% of BEC for gasifier, gas cleanup and gas turbine; 0% for others |
| (d) | Project contingency: | 15% of BEC |
| | Total plant cost (TPC) | = a + b + c +d |
| (e) | Total cash expended | $TPC/n*[1+1/(1+ea)^1+1/(1+ea)^2+1/(1+ea)^3]$ |
| (f) | AFDC (escalation and interest in construction period) | $TPC/n*\{i/(1+ea)+[(1+i)^2-1]/(1+ea)^2+ [(1+i)^3-1]/(1+ea)^3\}$<br>(n:construction years; i:discount rate; ea:inflation rate.) |
| | Total Plant Investment (TPI) | = TPC + e + f |
| (g) | Prepaid royalties | 0.5% of bare erected cost (BEC) |
| (h) | Initial catalyst &chemical inventory | 30 days |
| (i) | Startup costs | |
| | Plant modifications | 2% of TPI |
| | O&M costs | 30 days |
| | Fuel costs | 7.5 days |
| (j) | Working capital | |
| | Coal | 60 days |
| | Consumables (excluding water) | 60 days |
| | By-product inventory | 30 days |
| | Spare parts | 0.5% of TPC |
| (k) | Land | $6,500/acre |
| | Total capital requirement (TCR) | = TPI + g + h + i + j + k |

## 2. O&M Cost

An approach similar to that used in Appendix B1 was employed to estimate the fixed O&M costs for the IGCC plant. The total operating jobs are assumed as 17 persons per shift in this study; the maintenance cost is assumed 2.2% of the TPC (Table B2.3).

The variable O&M costs consist of the costs of consumables and fuel. Water and coal consumptions are obtained from simulation, and the others are linearly scaled from literature (EG&G, 2000; Akunuri, 2000). A loading factor of 70% is assumed for electricity generation. The unit purchase prices of consumables and fuels are listed in Table B2.3.

The credit of sulfur by-product from Claus/Scott process is included in the cost estimation. The price of sulfur product is assumed to be $75/t. In addition, operating royalties are also incorporated in this study as 1% of the fuel cost.

## Table B2.3. Assumptions used for estimating the O&M cost of the IGCC plant

| | |
|---|---|
| **Fixed O&M costs** | |
| Operating labor | 17 persons per shift, $40/hr |
| Maintenance costs | 2.2% of total plant cost (40% labor; 60% materials) |
| Administrative & support labor | 30% of total labor (operating + maintenance) |
| Miscellaneous operating costs | 10% of total labor |
| | |
| **Variable O&M costs** | |
| Coal | $20/t for PRB coal, $30.t for Illinois coal |
| Water | $0.06/t |
| Selexol solvent | $1.45/lb |
| Claus catalyst | $470/t |
| SCOT activated alumina | $0.067/lb |
| SCOT cobalt catalyst | $6,000/t |
| SCOT Chemicals | $275/t |
| | |
| By-product (sulfur) credits | $75/t |
| Royalties | 1% of fuel cost |

## 3. Cost of Electricity

The financial criteria used for estimating the cost of electricity (Table B2.4) were referred to in the literature (Parsons Infrastructure & Technology Group Inc., 2002). Based on these assumptions, a levelized carrying charge (applied to TCR) is 0.138. No inflation is considered for categories involved in the O& M costs.

**Table B2.4. Financial assumptions for the cost estimation**

| **General characteristics** | | |
|---|---|---|
| Plant startup date (year) | 2000 | |
| Capital cost year dollars | 2000 | |
| Construction period, yr | 4 | |
| Capacity factor, % | 70 | |
| **Financial assumptions** | | |
| Project book life, yr | 20 | |
| Book salvage value, % | 0 | |
| Project tax life, yr | 20 | |
| Tax depreciation method | ACRS class | |
| Property tax rate, %/yr | 1 | |
| Insurance tax rate, %/yr | 1 | |
| Federal income tax rate, % | 34 | |
| State income tax rate, % | 4.2 | |
| Investment tax credit% eligible, % | 0 | |
| **Capital structure** | **% of total** | **Cost, %** |
| Common equity | 42 | 12 |
| Preferred stock | 10 | 8.5 |
| Debt | 45 | 9.0 |
| Weighted cost of capital (after tax), % | 8.76 | |
| Escalation rates (apparent) | | |
| General escalation, % | 0 | |
| Primary/secondary fuel price escalation, % | 0 | |

A levelized busbar COE for the 70% design capacity factor is calculated according to the following relationship:

$$COE = \frac{13.8\% \times TCR + O\ \&\ M\ cost}{MW_{net} \times 0.7 \times 8{,}760}$$

where

COE = cost of electricity, mills/kWh;
MW$_{net}$ = new electricity output, MW;
TCR = total capital cost, \$; and
O&M cost = operating and maintenance cost, \$/year.

## 4. Costs of $CO_2$ Avoidance and $CO_2$ Capture

The cost of $CO_2$ avoidance and the cost of $CO_2$ capture are as described in Appendix B1. They are expressed as

$$\text{Cost of CO}_2\text{ capture (\$/ton)} = \frac{(\$/kWh)_{capture} - (\$/kWh)_{reference}}{(ton\ CO_2/kWh)_{capture}}$$

$$\text{Cost of CO}_2\text{ avoidance (\$/ton)} = \frac{(\$/kWh)_{capture} - (\$/kWh)_{reference}}{(ton\ CO_2/kWh)_{reference} - (ton\ CO_2/kWh)_{capture}}$$

where

capture = the IGCC plant with Selexol unit;
reference = reference PC plant without $CO_2$ capture;
\$/kWh = levelized cost of electricity; and
ton $CO_2$/kWh = $CO_2$ emissions per kWh generation.

Note that the PC plant without $CO_2$ capture is used as the reference plant for $CO_2$ cost calculation.

# Part 2 Truck and Rail Transportation Options

**Rajani Varagani**

## 1. Introduction

Sequestration of $CO_2$ consists of three independent steps: capture, transportation, and storage. For transport, the captured $CO_2$ must be suitable in terms of pressure, temperature, and composition because $CO_2$ is usually transported either in the supercritical/dense phase or in the liquid phase for maximum throughput and ease of loading and unloading. One of the main concerns for $CO_2$ transportation is inadvertent contact with water, which can cause problems such as hydrate formation and acidity increase. Hydrates are crystals that can plug pipes or valves or other equipment; acid formation is a corrosion issue.

## 2. Transportation Options

$CO_2$ can be transported by the following ways (Odenberger and Svensson, 2003):

1. Truck/motor carriers
2. Rail cars
3. Pipeline transportation
4. Water carriers (ships)

In order to understand the relative magnitude of the $CO_2$ transportation required for sequestration purposes, the total $CO_2$ emissions from coal burning power plants in the United States are approximately 4.9 million t/day (U.S. Department of Energy, 2000), but the nation's current $CO_2$ merchant market truck and rail infrastructure capacity is 30,000 t/day. Assuming mass is proportional to infrastructure, the current truck and rail infrastructure for handling $CO_2$ would have to increase 163 times to sequester and store 100% of the emitted $CO_2$. Hence, transportation by truck and rail is very likely not an option for permanent, long-term sequestration projects. However, it is very important to understand $CO_2$ transportation by truck and rail because that is the most probable scenario for transporting $CO_2$ to relatively short-term, sequestration test sites. Consequently, this report mainly discusses transportation of $CO_2$ intended for short-term demonstration tests of the second phase of the regional partnership program.

## 3. Truck Transportation

Motor carriers or trucks have been used for a long time for small-scale transportation because of their flexibility, adaptability, and reliability. $CO_2$ is transported at vapor pressure as a liquid with a pressure range of 275 to 300 psi and temperature of 0 to 10°F. A typical thermally insulated tank trailer (Figure 3.1) carries up to 22 t of liquid $CO_2$. A truck trailer costs about $110,000, and a tractor that pulls the trailer costs about $30,000. The cost of transportation by truck ranges from $1.75 to $2.00/mile per truck. These cost estimates assume that there is no wait to load or unload the truck and that roundtrip mileage is counted. This transportation technology is readily available but requires a logistic handling system that includes optimizing the number of trucks, scheduling loading and unloading, and choosing the route. Truck transportation is usually more expensive than rail or pipeline transportation.

Figure 3.1. Typical $CO_2$ truck trailer

## 4. Rail Transportation

Historically, railway is one of the main transportation options for bulk and less expensive materials. Railway can be very competitive to truck transportation because railways have large volume handling capacity over long distances. Railway infrastructure, however, plays a key role. The cost of transportation by rail depends mainly on the route from source to sink. If the route is direct, the transportation cost is less expensive, even for longer distances, than that for a shorter, indirect route with multiple transfers. For every rail-to-rail transfer, there is an additional fee that increases the transportation cost. Because the railway industry has high fixed costs and low operational costs, rail cars are often leased by companies rather than purchased.

A typical rail car (Figure 4.1) can hold up to 80 tons of liquid $CO_2$. The pressure inside the car is maintained at a constant 350 psi, the vapor pressure of the $CO_2$. If the temperature increases during transportation, the pressure is relieved by venting the $CO_2$ from the car. Hence, rail transportation is not recommended for long periods due to the associated loss of $CO_2$. The temperature inside the railcar is around 10°F. In the United States, it can take from 1 to 10 days for transportation between any two rail stations.

**Figure 4.1. Typical $CO_2$ railcar**

The typical cost of $CO_2$ transportation by rail is about $25/t and up to $5/t additional charges (e.g., transfer fees and yard usage fees). Distance traveled has minimal effect on transportation cost. Also, the cost to lease an 80-t rail car is approximately $900/month for a minimum of 12 months and can be lower for longer lease durations. As railway tracks may not be present at the wellhead (the point of injection), trucks are likely coupled with rail transportation. Figure 4.2 shows the unloading of $CO_2$ from a railcar to a truck trailer.

**Figure 4.2. Unloading $CO_2$ from railcar to truck**

## 5. Secondary Storage Requirements

After the $CO_2$ is transported by rail and/or truck to the $CO_2$ injection site, secondary storage may be necessary for the following reasons:

- well's injection capacity is relatively low and requires that the truck or railcar remains at the injection well location for unreasonable (uneconomical) durations;
- intermittent injection is unacceptable; and
- higher injection pressure is required that necessitates the use of compressors and/or pumps

Different types of storage tanks are available in the market. Portable tanks are mounted on a truck carrier base and can be moved from one place to another very easily. Portable tanks have a maximum capacity of 60 tons, but availability is limited. Customer tanks are installed permanently on a concrete floor and are intended for long-term usage. Typical customer tanks can hold up to 50 t of $CO_2$ with a rent of $1,500/month for at least 1 year. Storage tanks more than 50 t are available but are assembled at the site. Figure 5.1 shows customer storage tanks with capacities of 250 t (left) and 45 t (right two tanks). The number of storage tanks required depends on such things as injection rate and delivery intervals.

**Figure 5.1. Secondary storage tanks**

## 6. Unloading and Injection Equipment

Unloading $CO_2$ from railcars requires special equipment. Rail towers are built to unload $CO_2$ from rail cars because $CO_2$ is always unloaded from the top of the railcar (in contrast, $CO_2$ is unloaded from the bottom of truck trailers) (Figure 6.1). The number of towers required depends on the unloading rate or injection rate desired. Each tower can cost up to $20,000.

**Figure 6.1. Rail towers for unloading $CO_2$ from railcars**

Compressors and/or pumps are needed for unloading and injecting $CO_2$. The unloading equipment package (includes compressor, pumps, valves, and piping) costs about \$20,000 to \$25,000, and injection equipment package costs about \$50,000 for 200 to 300 tons/day injection rates. Figure 6.2 shows a typical compressor that can be used for $CO_2$ unloading/injection, and Figure 6.3 shows a typical injection skid.

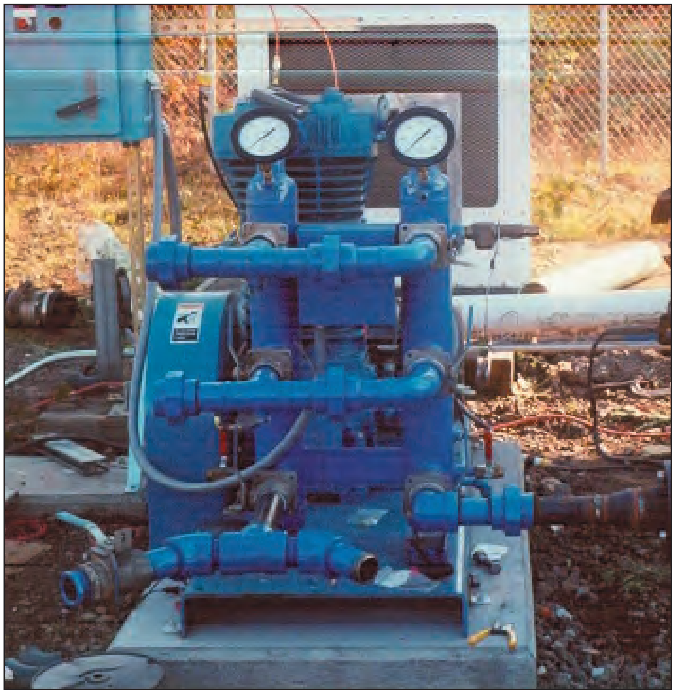


**Figure 6.2. Typical $CO_2$ compressor**

Figure 6.3. Typical $CO_2$ injection skid

## 7. Regulatory Information

In the United States, $CO_2$ transportation by rail, highway, air, and water is governed by federal authority under regulations promoted by the U.S. Department of Transportation (U.S. DOT) (www.access.gpo.gov). All of the equipment used must be in accord with the U.S. DOT standards. Additional information can be obtained from Compressed Gas Association, Inc. (2003).

## 8. References

Compressed Gas Association, Inc., 2003, CGA G6 2003: Chantilly, Virginia.

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U.S. Department of Energy, 2000, Carbon Dioxide Emissions from the Generation of Electric Power in the United States: Washington, D.C.

# Part 3 Assessing $CO_2$ Pipeline Transportation Options in the Illinois Basin

**Douglas J. Nyman**
**J. Steve Dracos**

## 1. Introduction

### 1.1. Scope

The team of D. J. Nyman & Associates and Universal Ensco, Inc. was commissioned to prepare this report on $CO_2$ pipeline transportation options in support of the Midwest Geological Sequestration Consortium's assessment of opportunities for geological carbon sequestration in the Illinois Basin. Our report is intended to provide a synopsis of the design and construction of a pipeline for the transportation of $CO_2$, including permitting and land acquisition at the front end as well as considerations for operation of the completed project. The synopsis is more than hypothetical; it has been prepared by a team that has designed pipelines and, more specifically, one that has designed $CO_2$ pipelines. Although in some respects the document takes the form of a tutorial, it is thought to be realistic with respect to the various activities that are paramount in designing a $CO_2$ pipeline.

Specifically, the team has addressed the following aspects of $CO_2$ pipeline design, construction, and operation:

1. Identify and describe the elements of a CO2 pipeline transportation system, the general properties of CO2, and the impact of these properties on pipeline design, operations, and related safety issues.

2. Describe the design, materials, and construction practices that are common to CO2 pipelines including corrosion allowances, pipe and valve standards, type and spacing of mainline valves, common pumping and metering equipment, and comments based on practices specifically suitable to the Illinois Basin.

3. Identify right-of-way considerations and permitting requirements for pipelines in the basin, including co-location with existing pipelines and power lines, as well as agricultural mitigation (e.g., topsoil handling and drain tile avoidance).

4. Describe operating practices such as pipeline blow-down considerations, marking of the pipeline for third-party damage prevention, and community awareness programs necessary as part of sequestration outreach activities.

5. Prepare a cost matrix of approximate cost per mile based on pipeline diameter and construction settings (e.g., as urban and rural), including adjustment factors as appropriate for farmland, timber, terrain, and soil and rock conditions, which includes power costs based on pipeline diameter and transport volume.

The data presented in this report will be used to provide pipeline transmission cost information into a $CO_2$ capture and sequestration macro-economic model, which will be used to analyze numerous capture, transport, and injection cases. The cost information provided in this report will be used to establish a primary or base transportation case with capture and sequestration locations for more detailed assessment. Detailed cost estimates for the final selected pipeline transportation case will be developed in a future project task (Task 9).

A preliminary conceptual route was selected to give the study a practical basis to facilitate the acquisition of meaningful data from industry sources. The selected conceptual route originates in the Peoria area, in the vicinity of several large coal-fired power plants, goes southward to a point slightly east of Springfield, and then extends south-southeast toward the Illinois, Indiana, and Kentucky border. The southern terminus is within a likely storage fairway emerging from the geological sink characterization studies. For planning purposes the pipeline was assumed to be approximately 200 miles in length (Figure 1.1). Data were gathered from the following sources:

- The authors' experience and accumulated body of knowledge
- Projects recently completed by Universal Ensco in Illinois
- A $CO_2$ pipeline recently completed by Universal Ensco
- Miscellaneous industry publications and industry Web sites
- Vendors (unit costs for material)
- Local union contractors (construction costs)
- County agencies along the route (land costs)

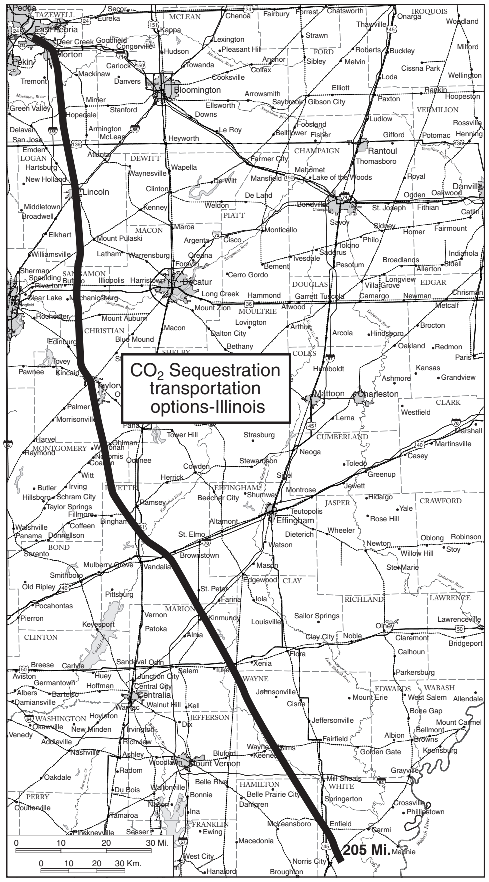

**Figure 1.1. Conceptual route**

### 1.2. Background

$CO_2$ is used extensively by the oil industry to enhance oil production. Currently, there are 70 active $CO_2$ projects of this type in the United States. Most of these involve a large $CO_2$ pipeline as part of the project operation, and almost all are supplied from natural underground sources of $CO_2$. There are currently about 2,400 km of large $CO_2$ pipelines in operation. A representative listing of some of the major $CO_2$ pipelines is given in Table 1.1.

**Table 1.1. Major $CO_2$ pipelines in operation in the United States**

| Pipeline | Date | Diameter (inches) | Length (miles) | Location |
|---|---|---|---|---|
| Bravo | 1984 | 20 | 218 | Texas |
| Transpecto/Bravo | 1996 | 12 | 120 | Oklahoma |
| Cortez | 1984 | 30 | 502 | Colorado-Texas |
| Sheep Mountain | Approx. 1984 | 20/24 | 408 | Colorado-Texas |
| Central Basin | Not available | 26/16 | 26 | Texas |
| Este | Not available | 12/14 | 119 | Texas |
| Trinity | Not available | 12 | 180 | W. Texas |
| Caprock | 1972 | 16 | 140 | Texas |
| Centerline | 2003 | 16 | 120 | Texas |
| Salt Creek | 2003 | 16 | 125 | Wyoming |

$CO_2$ pipelines have compiled an excellent safety record during the 32 years that they have been used in the industry. The design, material, and construction issues related to the safe operation of $CO_2$ pipelines are regulated under the *2001 Code of Federal Regulations, Parts 190–199* (International Code Council, 2002). A point to note is that under federal regulations in the United States, $CO_2$ pipelines are classified as high volatile/low hazard and low risk (due to the product being nonflammable). As a matter of public record, during the period 1991 to 2001, there were no $CO_2$ pipeline-related injuries or deaths. A more detailed discussion of operations is included in Section 7.

Readers who are unfamiliar with pipelines should find it helpful to review Section 6, Construction, and the construction photo album (Appendix C1). A typical $CO_2$ pipeline consists of a receipt meter and pressure regulating facilities, receipt lateral pipelines, a main pipeline, a booster facility (if any), delivery lateral pipelines and delivery meter, and pressure/flow regulating facilities. These items are supported by auxiliary facilities such as a control center and operations and maintenance facilities.

## 2. Route Selection and Design

The selection of a good route is key to the success of any pipeline project. All pipeline projects of considerable length will encounter numerous objections by the stakeholders— landowners, tenants, environmental and other government agencies, and nongovernmental agencies. Although the stakeholders are motivated by a variety of issues and concerns, almost all of their objections will be focused on the route. During any proceedings related to condemnation, the proposed route requires strong justification.

In general, a properly selected, good route will withstand objections, and a poor route will not. As a minimum consequence, poor route selection results in rework and delays, but more serious ramifications are possible. Routes are selected in two stages: preliminary route selection and detailed route selection. The preliminary route is used as a basis for the planning and budgeting phases of the project. Once the project receives funding for permitting and right-of-way acquisition, detailed route selection and design are initiated. In all phases, the co-location of the pipeline route in an existing right-of-way corridor is a very important consideration.

### 2.1. Preliminary Route Selection

The objective of preliminary route selection is to identify several potential pipeline routes that minimize the impact to the public, landowners, and environment. This activity is typically undertaken by a relatively small group of specialists with local knowledge of right-of-way, environmental, construction, and operations. This selection process is essentially defined by the beginning, end, and key crossing points along the route, such as crossings of major rivers or other terrain features and areas of environmental impact, urban centers, parks, or environmental preserves that must be avoided. Available maps and published environmental data for the region to be traversed can be reviewed to determine the optimum route, which may include several route variations. By using GIS (Geographic Information System) resources, the route selection process can be expedited, and the quality of the final product can be markedly improved.

Often an existing corridor (e.g., electric transmission line or pipeline right-of-way) can be identified as a prospective routing for the pipeline under consideration. Even in the open agricultural land of Illinois it is generally best to follow a corridor if one is positioned in the general direction of the pipeline. However, if the corridor leads away from the next key pipeline point, thereby adding distance, the proposed route should move away from the corridor toward the next key point. It is important to recognize that increasing the length of pipeline will impact more landowners at higher overall cost. It is also important to recognize the age of the corridor because an older corridor that was established prior to contemporary environmental regulations might be located in areas that have an unacceptable impact under today's standards. Also, if the corridor has restricted construction room or other issues, the cost of working in the corridor could be higher than the cost associated with a longer route.

As a general guideline, the preliminary route selection for a 200-mile pipeline route in Illinois could be completed in approximately one to two months. Once selected, the preliminary route will be confirmed by ground and/or aerial reconnaissance in the field. Changes will be incorporated into the route as applicable.

### 2.2. Detailed Route Selection

After the preliminary route has been selected and approved, detailed route selection will commence. The key points and route variations in the preliminary route are analyzed, and a procedure for assessing route variations is established. The reasoning for the routing is documented so that consistency can be demonstrated during discussions with agencies and landowners as well as during potential eminent domain proceedings. The detailed route is scrutinized carefully through use of high-resolution, project-specific aerial photography and numerous field trips until a final preferred route is determined. Specific route information is entered into a GIS database.

Mainline and lateral valve sites will be located near roads to maximize ease of access; however, locations near populated areas will be avoided if possible due to the potential for vandalism or similar disruptions. Mainline and lateral valve settings will be located to avoid overhead obstructions, power lines, and floodplains.

Detailed cost estimates are normally prepared at the conclusion of selecting a final preferred route, although a limited number of route variations may remain under consideration pending land acquisition and permitting.

### 2.3. Detailed Route Design

Once a preferred route is identified, the pipeline company will contact landowners to discuss the project and seek permission to conduct civil and environmental surveys. These surveys are required for use in the detailed pipeline design and for preparing local, state, and federal permit applications. It is important to note that, even though pipeline officials may begin discussions with landowners at this point, the project is still being analyzed for feasibility, and neither the project nor the pipeline route is finalized at this time. Selecting a pipeline route generally involves discussing and evaluating options with landowners, environmental agencies, and regulatory officials. Once alternatives have been analyzed, field studies are initiated to confirm the environmental assumptions. These environmental studies generally follow procedures set out by federal and state law, sometimes resulting in environmental impact statements or environmental assessments that are published in draft form for public comment. During this process, landowner agreements must be obtained, and minor route variations are made to accommodate individual landowner's physical needs. The necessary permissions to locate within a corridor also must be obtained during this phase of the project. The process of obtaining this permission is discussed further in Section 5, Right-of-Way.

### 2.4. Corridors

During all phases of pipeline routing, the use of existing corridors for electric power transmission, other pipelines, or utilities is a major consideration. In urban or industrial areas, routing in existing corridors is required. Because the $CO_2$ needed for the sequestration will be coming from existing power plants that are probably located in urban areas, it is necessary to follow one of several corridors, such as power lines, utility lines, and railroads, that connect to the plant. Of these corridors, the power line corridor is probably the preferred pipeline route for two reasons. First, the power company or its affiliate generally owns it in fee, and, second, power line corridors generally have adequate workroom for installation of a pipeline without the use of adjacent property. Utilities generally do not own the corridor in fee, but rather have an easement, which makes it necessary to contact the underlying landowner to obtain a separate easement, which is already encumbered by the existing easement. Railroad rights-of-way are owned in fee but are so narrow that additional workspace from adjacent landowners is usually needed.

When a pipeline is routed in corridors containing high-voltage power lines, consideration must be given to areas where interference from electrical currents may occur. Interference can also occur in places where the pipeline crosses, is in close proximity to, or parallels high-voltage power lines. Special grounding procedures, such as installing zinc ribbon anodes along the pipeline and gradient ground mats at valves/test stations will ensure personal protection and allow for safe operation of the pipeline.

## 2.5. Geohazards

A $CO_2$ pipeline in central and southeastern Illinois may be subject to a several types of geohazards, including landslides in steeply sloped ground, seismic hazards such as liquefaction, seismic-induced landsliding, seismic wave propagation, ground shaking, and coal-mining subsidence. These potential geohazards can be avoided or mitigated through careful route selection and reconnaissance, state-of-the-art engineering methodology, and innovative design practice.

### *2.5.1. Landslides*

Landslides are downslope movements of soil masses, most often triggered under static conditions by changes in moisture conditions and water table or by changes in slope geometry, such as undercutting of the toe of the slope. Such changes result in a redistribution of shear stresses along the slope, with a concurrent reduction in soil shear strength along the slide plane or a reduction in the resistance against sliding provided at the toe of the slope.

When a pipeline is contained within a sliding soil mass, it must deform to accommodate the differential displacement between the sliding mass and the adjacent ground outside the zone of sliding. Figure 2.1 is a schematic illustration of the effect of a landslide on a pipeline crossing a slide zone.

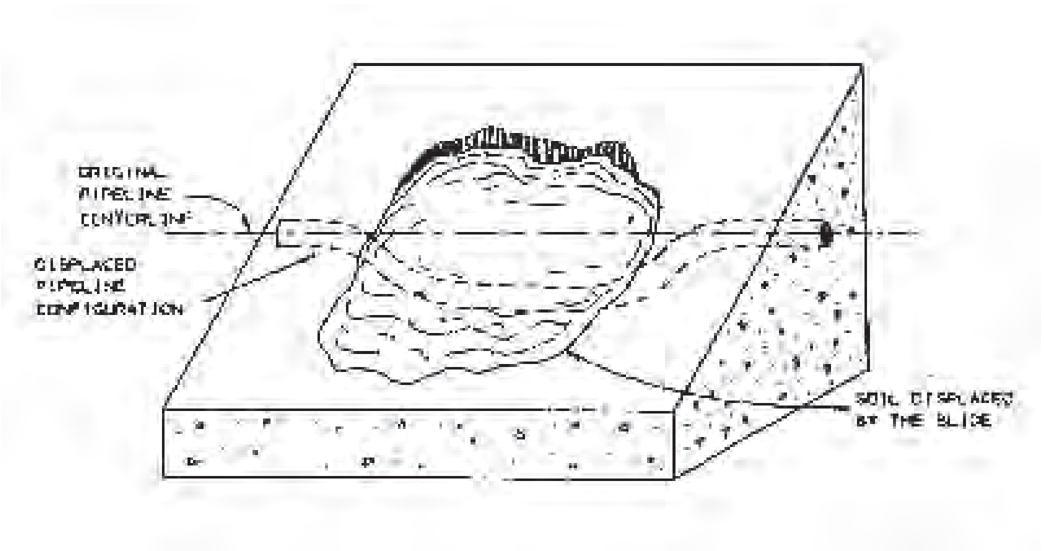

**Figure 2.1. Buried pipeline subjected to landslide**

The most significant landslide hazards that can affect buried pipelines are slumps, shallow slides, deep rotational slides, and translational slides. Slumps and shallow slides are caused primarily by inertial forces, but they are often assisted by densification of loose soil or liquefaction of underlying sediments. These movements occur mostly along the margins of embankments, cut-and-fill slopes, and slopes with relatively shallow cover in hilly or mountainous terrain. Deep slides involving significant components of translation and rotation of a soil mass can develop catastrophically and affect large areas. Translational slides are more likely to occur on natural slopes, whereas rotational slides often occur in homogenous materials such as highway fill embankments. A landslide frequently causes underthrusting in soils near the base of its slope so that substantial compression and bending may be transferred to pipelines located there.

The landslide threat to the proposed $CO_2$ pipeline in Illinois is thought to be minimal because most, if not all, of the route is through relatively flat or gently rolling terrain that is not particularly susceptible to landslides. In steeper areas where landslide risk may be present, the preferred approach is simply to avoid such hazards by careful routing away from them. Engineered solutions to stabilize slopes (e.g., drainage, buttresses, and tie-backs) are normally required only in mountainous terrain with limited routing options.

### *2.5.2. Seismic Hazards*

Considerable attention has been focused in recent years on the potential consequences of future earthquakes in the central United States. The New Madrid Seismic Zone (NMSZ) in the central Mississippi River Valley is the primary seismic source for this area. The NMSZ is known mainly for the succession of four shocks, known collectively as the New Madrid earthquake sequence, which occurred on December 16, 1811 (two shocks), January 23, 1812, and February 7, 1812. The epicenter locations for these four earthquakes, which were felt over a major portion of the eastern United States, are shown on Figure 2.2 (Hopper, 1985). Street and Nuttli (1984) estimated that three of these four earthquakes would have measured between 8.4 and 8.7 surface wave magnitude, $M_s$. In addition to the four main shocks, there were numerous aftershocks.

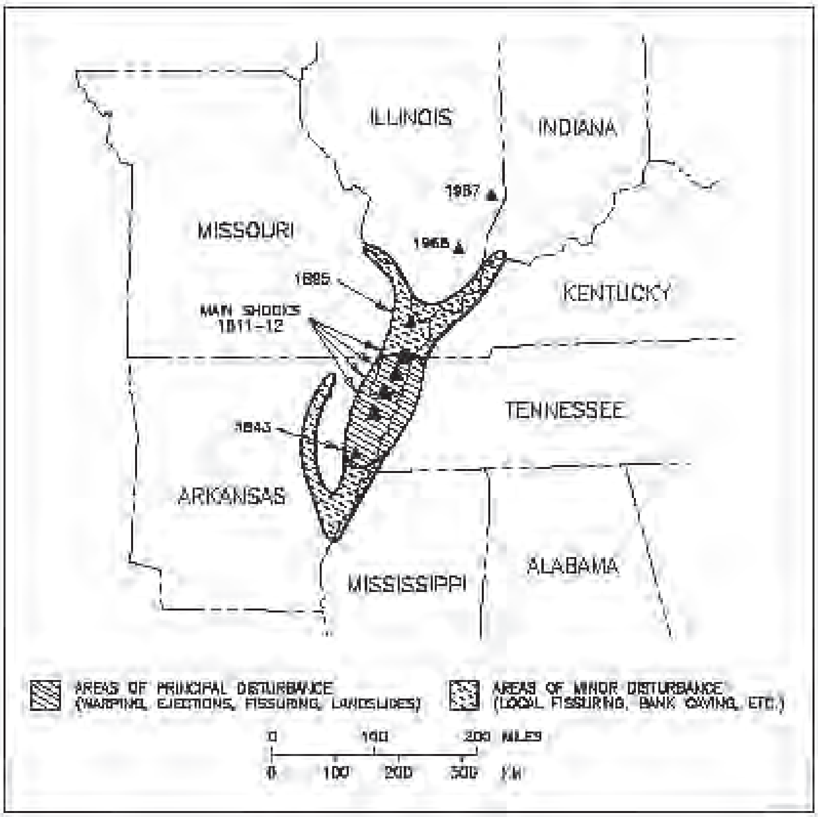

**Figure 2.2. Epicenters of New Madrid earthquakes and areas of soil disturbance in 1811–1812 (Hopper, 1985; Fuller, 1912)**

Based on a field study 100 years after the New Madrid earthquakes, Fuller (1912) characterized ground disturbance in the area affected by the earthquakes (Figure 2.2). Fuller noted numerous observations of ground fissuring, some of which were several hundred feet or more in length and 30 ft (9 m) wide. It is believed that much of the fissuring described by Fuller was lateral spreading or flow slides associated with liquefaction of a sand layer typically 6 to 15 ft (2 to 5 m) deep (Obermeier, 1985). Fuller also called attention to numerous landslides along river banks and in the high bluffs bordering the lowlands on the east side of the Mississippi River in southwestern Kentucky and western Tennessee and upstream along the Ohio River as far as Indiana. Owing to the general absence of documented accounts, it is believed that ground failure in Illinois was relatively insignificant, except, of course, for areas along the Mississippi and Ohio Rivers.

Since 1812, two significant shocks have occurred in the NMSZ, one in 1843 in northeastern Arkansas and the other in 1895 in southeast Missouri. These earthquakes were estimated to have had body wave magnitudes ($m_b$) of 6.0 and 6.2, respectively (Nuttli, 1974). Other large earthquakes in the region include two earthquakes in southern Illinois, a 5.5-$m_b$ event near Broughton in 1968, and a 5.1-$m_b$ event near Olney in 1987, and a magnitude 5.0 event near Evansville, Indiana, in 2002. The Evansville event occurred in the Wabash Valley Seismic Zone (WVSZ), which is located in southeastern Illinois and southwestern Indiana. The WVSZ is thought by some investigators to be capable of producing "New Madrid"-sized earthquake events. Field investigations of prehistoric earthquakes in the region indicate that an event of estimated magnitude 7.0 occurred in the WVSZ approximately 6,100 years ago (http://www.cusec.org). Thus, it would be prudent to consider this proximate seismic source in the design of a $CO_2$ pipeline route that passes into southeastern Illinois.

The principal earthquake hazards for pipeline systems in southern Illinois include ground failure due to liquefaction or landslides and ground shaking effects on aboveground facilities and equipment. Ground settlement and seismic wave propagation are less important, but could possibly affect buried pipelines in certain special circumstances. Surface fault rupture is also potentially hazardous to pipelines, in general, but there is no evidence of active faults in southern Illinois that might affect the planned route of the $CO_2$ pipeline. The potential effects of the various earthquake hazards on the $CO_2$ pipeline is described in this report.

Transmission pipelines are typically buried under a soil cover of 3 to 4 ft or deeper in agricultural areas. Burial tends to make a pipeline more susceptible to large permanent ground distortions such as landslide or liquefaction-induced ground displacement. The effects of seismic ground movement along a pipeline route are generally sporadic, depending on local soil, groundwater, and topographic conditions. The nature of the ground displacements are not well defined and generally can occur at any point within an area of potential movement. The amount of ground displacement depends largely on the intensity and duration of earthquake ground shaking.

*2.5.2.1. Liquefaction.* Soil liquefaction is defined as "the transformation of a granular soil from a solid state to liquefied state as a consequence of increased pore water pressure and reduced effective stress." Liquefaction occurs as seismic waves propagate through saturated granular sediment layers, which induce cyclic shear deformation and collapse of loose particulate structures. As collapse occurs, contacts between grains are disrupted, and loads previously carried through those particle-to-particle contacts are transferred to the interstitial pore water. This load increases pore water pressure and concomitantly decreases intergranular or effective stress. As pore water pressures increase, the sediment layer softens, allowing greater deformation and an accelerated rate of collapse of the particulate structures. When the pore pressure reaches a certain critical level, the effective stress approaches zero, and the granular sediment begins to behave as a viscous liquid rather than a solid. Liquefaction has occurred.

Liquefaction occurs in natural deposits in a rather narrow range of geologic and soil environments. Sediments that are most susceptible to liquefaction are granular soils that remain loose and uncemented after deposition during recent geologic time (modern or late Quaternary eras). Liquefaction occurs only in saturated sediments beneath a shallow groundwater table. Strong ground shaking can also trigger the transformation of deposits into a liquefied state, as could occur for a strong event occurring in the NMSZ or the WVSZ.

Liquefaction by itself poses little hazard to pipelines. Damage generally occurs when liquefaction leads to ground deformation or ground failure. Liquefaction may lead to flow failure, lateral spread, ground oscillation, buoyant rise of buried pipelines, or ground settlement. The type and extent of ground failure depends on site geometry and the depth, thickness, and extent of the liquefied layer.

1. Flow failure is the most catastrophic type of permanent ground deformation caused by liquefaction. Flow failure occurs on steeper slopes (greater than 6% or 3.5 degrees) underlain by loose liquefiable soils. Flow failures are characterized by large lateral displacements (several meters or more) and severe internal disruption of the failure mass. The preferred mitigation strategy is to avoid potential flow slide areas, as is done for potential landslide areas.

2. Lateral spread occurs on slopes that are too gentle to develop flow failure or in areas near a free face, such as an incised river channel. Lateral spreads involve sideways movement of surficial soil layers down the gentle slopes or toward the free face. Lateral spread displacements may be as large as several meters accompanied by shear failure zones as deep as several meters, depending on the depth of the liquefied soil and the geometry of the site. Displacements commonly occur as far as a few hundred meters from incised river channels. Displacement occurs in response to a combination of gravitational and earthquake-generated inertial forces acting on sediments within and above the liquefied zone. During displacement, the soil layers commonly break into large blocks, which transiently jostle back and forth and up and down in the form of ground waves (ground oscillation) as they migrate laterally. Displacements usually range up to 20 ft, but where ground conditions are particularly vulnerable and shaking is intense, larger displacements have occurred.

   Lateral spreads create extension or tensional features such as open fissures at the head (upslope) of the failure, shear deformation along the margins, and compressional features such as buckling at the toe. These movements have pulled apart, sheared, and/or compressed pipelines during past earthquakes. For example, lateral spreads fractured water, gas, and oil pipelines during earthquakes in Niigata, Japan (1964), San Fernando, California (1971), Northridge, California (1994), and Kobe, Japan (1995), and elsewhere (O'Rourke and Tawfik, 1983; Oka, 1996; O'Rourke, 1995).

3. Ground oscillation occurs on flat terrain in response to inertial forces acting on decoupled soil materials above or within the liquefied zone. This decoupling allows large transient ground motions or ground waves to develop, but permanent displacements are usually small

and chaotic. Observers of ground oscillation commonly note slow-moving ground waves, up to a meter high, with wavelengths of tens of meters, accompanied by opening and closing of fissures. Generally, welded steel pipelines are capable of withstanding the flexural strains associated with ground oscillation.

4. Buoyant rise of pipelines may occur when the surrounding soil liquefies. For example, buried oil and gasoline tanks buoyantly rose during the 1993 Hokkaido Nansei Oki earthquake in Japan (Youd et al., 1995). Special measures to prevent buoyant rise are a design consideration for some pipelines in liquefiable areas. In most instances, well-fabricated steel pipelines can accommodate some rise, especially if distributed over a large length of the line, without exceeding the flexural strength of the pipe. Where mitigation is necessary, one effective measure is to surround the pipe with well-compacted backfill. Another measure is to embed the pipeline beneath the liquefiable layer to avoid the hazards associated with liquefaction. Often concrete coating, as used in standard floodplain and marsh construction, is sufficient to prevent buoyant rise. Buoyant rise is unlikely to be an issue for a small-to-medium diameter $CO_2$ pipeline, because the high operating pressure requires thick wall pipe, which, in combination with the weight of the liquid $CO_2$, results in a negatively buoyant pipeline system.

5. Ground settlement may occur due to the compaction of liquefiable deposits during earthquake shaking. Settlements may range from a few percent of the thickness of loose liquefiable layers to a fraction of a percent for denser sediments. Uniformly thick layers of homogeneous sediment usually compact and settle rather evenly with little consequent damage. Where granular layers vary in thickness, however, differential settlements may develop that could damage a buried pipeline. However, because of the generally ductile nature of steel pipelines and the generally small and widely distributed strains induced by all but very large and very concentrated ground settlements, very little damage has occurred to well-constructed steel pipelines from liquefaction-induced ground settlement.

Lateral spread is the most serious liquefaction-related hazard for the $CO_2$ pipeline system, because a pipeline crossing a zone of lateral spread displacement must deform longitudinally and in flexure to accommodate ground displacement. The areas most susceptible to lateral spread are at river and stream crossings at the southeastern end of the proposed pipeline route; these areas are closest to the NMSZ and WVSZ. In such areas, the design approach is to provide for a gently sloping (approximate 10%) transition from normal burial depth in the floodplain to the maximum depth beneath the channel and to use pipe and welds capable of mobilizing high strain without loss of pressure integrity. Some damage requiring repair may occur for large lateral spread displacement, but the overarching objective is to maintain pressure integrity in the deformed pipeline segment.

*2.5.2.2. Earthquake-induced Landslides.* As already described, most landslides occur under natural conditions related to increased moisture and/or changes in slope geometry. However, landslides can also be triggered by the inertial forces associated with seismic ground shaking, which increase the driving force for slope movement. Seismically induced landslides typically occur in slopes that have only marginal stability under static moisture, water table, and steepness conditions. Earthquakes exacerbate the climatic hazard of these marginally stable slopes (Transportation Research Board, 1996).

*2.5.2.3. Seismic Wave Propagation.* Body waves, including compression waves and shear waves, propagate radially from the source of earthquake energy release (hypocenter) into the surrounding rock and soil medium. Compression waves cause axial compressive and tensile strains in the ground in a radial direction away from the hypocenter. Shear waves cause shear strains in the ground perpendicular to these radial lines. When the compression waves and shear waves are reflected by interaction with the ground surface, surface waves (Love waves and Rayleigh waves) are generated. Except at very large distances from the epicenter, the magnitude of surface waves is much less than body waves.

A pipeline buried in soil that is subject to the passage of these ground waves will incur longitudinal and bending strains as it conforms to the associated ground strains. In most cases, these strains are relatively small, and welded pipelines in good condition typically do not incur damage. Propagating seismic waves also give rise to hoop membrane strains and shearing strains in buried pipelines, but these strains are small and may be neglected.

Well-constructed, buried oil and gas pipelines in good condition generally have not been affected by seismic wave propagation. There is no reported case of failure of a ductile, full penetration welded oil or gas pipeline attributable to wave propagation alone. Recent earthquake experience (Honegger, 1999) has indicated that wave propagation is a credible earthquake hazard for pipelines only in cases of extremely poor-quality girth welds or corrosion defects subjected to very high levels of seismic ground motion.

*2.5.2.4. Ground Shaking.* Earthquake ground shaking will cause seismic dynamic loading of above-ground pipeline facilities. Typical facilities include buildings, structures, vessels, liquid storage tanks, piping, mechanical and electrical equipment, control systems, instrumentation, and communications. The seismic design of pipeline facilities follows typical building code approaches, such as the the 2003 edition of the International Building Code (ICC, 2002) and ASCE Standard 7-02 (ASCE, 2002). Special attention should be given to ensuring the operational integrity of systems that provide essential monitoring, control, safety, and emergency functions. Examples of critical components include monitoring instrumentation, communications equipment, computer hardware, remote valve auxiliary equipment, emergency power systems, and uninterruptible power supplies.

### *2.5.3. Coal Mining Subsidence*

According to the National Mining Association (www.nma.org), Illinois ranks seventh in the U.S. coal production and produces about 3% of the nation’s coal. Illinois coal is mined by both underground and surface mining methods. The extraction of underground-mined coal causes displacement of the overlying strata that may reach the ground surface through subsidence. Underground mining in Illinois uses (1) room-and-pillar mining and (2) longwall mining methods. Each method is uniquely associated with subsidence of the ground surface.

Room-and-pillar mining drives entries, hallways, panels, and rooms in the coal seam, leaving 50 to 70% of the coal in place to prevent collapse of the cap rock into the mined cavities. The configuration of mined panels is intended to prevent or limit surface subsidence, and such may be the case for many years. However, slow deterioration of underground pillars long after mining has ceased may manifest itself as subsidence years or decades later.

Longwall mining provides for the full extraction of large panels of coal perhaps 500 to 1,200 ft in width and one to several miles long. Longwall mining is accomplished with a longwall mining machine that extends the full width of panel and roof shields to prevent the collapse of the roof of the mined panel onto the equipment. As longwall mining progresses along the length of the panel, the roof collapses behind the advancing equipment and shields. Subsidence due to long wall mining operations occurs quite rapidly as mining progresses, usually within days and typically complete within about 4 to 6 weeks.

For room-and-pillar mining, surface subsidence may be "pit" or "sag." Pit subsidence causes an abrupt drop in the surface and nearly vertical or belled outward walls (Bauer and Hunt, 1982). Pit subsidence is caused by collapse of a mine void and is most often associated with abandoned, shallow mines in areas with weak soil and rock overburden. Sag subsidence is usually associated with high extraction room-and-pillar mining or longwall mining. Sag depressions at the surface are typically several 100 ft wide with gentle slopes.

Regardless of the mining method used, the subsidence associated with underground mining can present a serious threat to the integrity of a buried pipeline. The pipeline must deform (bend, elongate, or shorten) to accommodate the vertical and horizontal displacement of the ground surface. Pit subsidence along a pipeline right-of-way could result in an abrupt loss of support for the pipeline while possibly leaving some of the overburden material on top of the pipeline as an external gravity load effect. Sag subsidence would cause the pipeline to deform in a sag-like shape to assume the new profile of the ground. Deformation at the margins of the sag depression could overstress the pipeline. If the pipeline bend exists within the subsidence depression, horizontal ground strains could possibly overstress the pipe bend as an applied thrust acts outward in the bend.

The prudent course of action for a pipeline project is simply to avoid routing the pipeline through areas of Illinois with potentially unstable abandoned mines or where high extraction mining could be reasonably anticipated. Nevertheless, if the $CO_2$ pipeline is located in a coal mining area, potential ground subsidence needs to be monitored over the course of its operating life. If high-extraction mining, particularly longwall mining, is to be conducted along the right-of-way, it is probably necessary to uncover (excavate) the pipeline prior to mining to allow the pipeline freedom of movement and to temporarily support the pipeline on sand bags or cribbing to restrict actual pipeline displacement to acceptable amounts. Additional safeguards in areas subjected to possible mining subsidence include the use of highly ductile pipe with overmatching girth welds (i.e., welds stronger than the pipe to mobilize ductile pipe behavior without weld fracture). This combination of ductile pipe and overmatching welds is often referred to as "high-strain" pipe.

## 3. Design Elements

### 3.1. $CO_2$ Properties

It is important to understand the properties of $CO_2$ in the range of usual pipeline operating temperatures and pressures. Depending on the temperature and pressure, $CO_2$ exists as a solid, liquid, or gas. The phase diagram shown in Figure 3.1 describes the relationship of these variables for $CO_2$ over a range of pressures and temperatures.

As noted on the phase diagram, the triple point (pressure, 5.1 atm/75.1 psia; temperature, 56.7°C/133°F) is the pressure and temperature at which the three phases (gas, liquid, and solid) exist simultaneously in thermodynamic equilibrium. The critical point (pressure, 72.8 atm/1,067 psia; temperature 31.1°C/88°F) is the point above which the liquid and gas phase become indistinguishable. This region of the phase diagram—known as the dense, superfluid, or supercritical phase—is the region in which $CO_2$ pipelines operate.

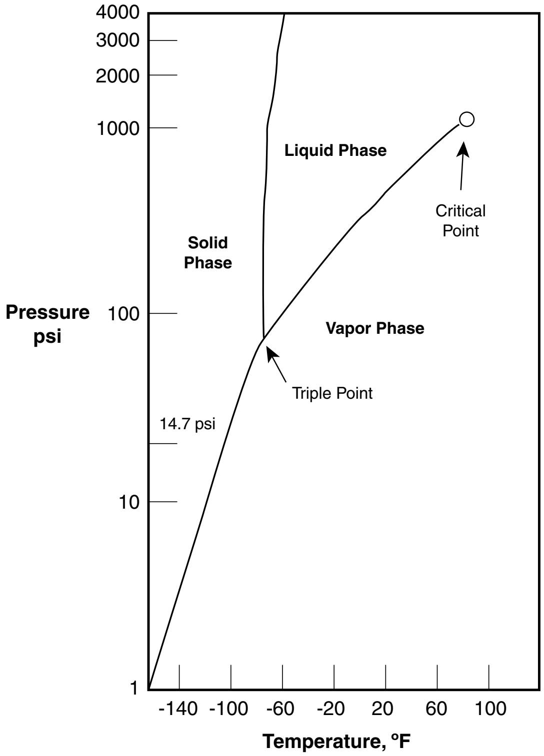

**Figure 3.1. Properties of $CO_2$ over a range of pressures and temperatures**

Another important property of $CO_2$ is the solid-gas phase boundary. Physically, this boundary implies that the gas and solid can coexist and transform back and forth without the presence of liquid as an intermediate phase. A solid evaporating or changing phase directly into the gas is called sublimation. Solid $CO_2$ is called dry ice because it does not go through a liquid state in its phase transition at ambient conditions.

It is not economical to operate a $CO_2$ pipeline in a gaseous region because of the larger-diameter pipeline needed to transport the product in its less dense, gaseous state. Instead, a $CO_2$ pipeline system normally operates in the superfluid region. Pressures are normally 2,500 to 2,700 psig at the inlet (nominal maximum operating pressure) and 1,400 to 1,600 psig at the outlet (nominal minimum operating pressure). These pressure ranges, although higher than most pipelines, are similar to those used on liquified petroleum gas (LPG) pipelines.

The ambient temperature for an underground pipeline in an Illinois agricultural area varies from 45°F in the winter to 70°F in the summer.

$CO_2$ is considered as a "natural refrigerant," a category that includes ammonia and hydrocarbons such as ethane, propane, and butane. As with any refrigerant, when $CO_2$ is under high pressure and is flashed across an orifice to a lower pressure, it attempts to absorb heat from the surrounding environment. If heat cannot be absorbed in the process, the refrigerant will drop to a lower temperature. Materials for the

$CO_2$ pipeline need to be specified to withstand the refrigeration temperature that can be expected during a pressure reduction. This refrigeration temperature usually calculated from the expected maximum operating pressure to atmospheric pressure. If not specified properly, the pipeline materials could become brittle under certain conditions and fail in a catastrophic manner.

Natural gas lines that have combinations of contaminants, which include $CO_2$, $H_2O$, and $H_2S$, can experience major internal corrosion. In these contaminated natural gas pipelines, water reacts with $CO_2$ and/or $H_2S$ to form carbonic or sulfuric acid, and corrosion of the pipe steel occurs. Similarly, corrosion problems will occur in a pure $CO_2$ pipeline if water is present as a contaminant. In all $CO_2$ pipelines, the process plant limits the water content of $CO_2$ before it is input into the pipeline meter and regulator facilities. A number of pipelines have operated successfully while requiring that water contact be limited to 30 lb of water per million standard cubic feet. An example of the composition of the $CO_2$ provided to the pipeline for a recent project is listed in Table 3.1. The water content in this example corresponds to approximately 8 lb/million standard cubic feet.

**Table 3.1. $CO_2$ composition (example) showing a typical component supply limit**

| Component | mol% |
|---|---:|
| Nitrogen | 0.041670 |
| Methane | 1.116670 |
| Hydrogen sulfide | 0.000330 |
| Carbonyl sulfide | 0.026790 |
| Ethane | 0.014900 |
| Water | 0.016670 |
| Carbon dioxide | 98.782990 |
| Total | 100.000000 |

### 3.2. Flow Rate Design

Flow rate is the single most critical design input for a $CO_2$ pipeline. A set length, design flow rate, and a set of pressures and pipe diameters can be analyzed to reach the most economical and practical design. Most $CO_2$ pipelines operate with an upstream pressure of 2,500 psig or higher and have a delivery pressure near 1,500 psig. The meter and regulator stations at each end require a pressure drop of approximately 50 psi each, giving a pipeline differential of 1,000 psi to achieve the flow rate.

For preliminary sizing, the use of these inlet and outlet pressures and approximate physical properties allow a suitable hydraulic analysis to be performed. The physical properties used in the analysis are average specific gravity and viscosity calculated based on the average pipeline pressure and ground temperature. A brief narrative of the assumptions and methodology follows:

1. $CO_2$ = average pressure of 1,900 psig

2. $CO_2$ density = 48.49 lb/cu. ft. (calculated using FlashCalc for pure $CO_2$ at 1,900 psig and 80°F, assumed average temperature)

3. $CO_2$ viscosity = 0.06 centipoise (cP) for $CO_2$ at 80°F.

4. Pipe wall thickness (WT) is standard wall as defined by the manufacturing codes.

Liquid density = 48.49 lb/ft$^3$ (water = 62.4) at 1,900 psig and 80°F
Vapor density at STP = 0.1152 lb/ft$^3$ (air = 0.076)

5. Pressure drop was calculated using the following transmission factor (T):

$T_{(oil)} = 3.6 \log (Re/8).$

Smooth pipe function,

$$(F_{smooth}) = T_{oil} - \frac{4\log\left(\frac{R_e}{T_{oil}}\right) - 0.6 - T_{oil}}{4\log\left(\frac{4}{T_{oil}\ \ln(10)}\right) - 1} - 0.$$

Three iterations were used for this function.

Darcy-Weisbach friction factor:

$$f_{oil} = \frac{4}{T_{oil}^2}.$$

Darcy's formula for head loss (Crane Technical Paper #410):

$$h_L = \frac{0.1863\, f\, L v^2}{D_o}.$$

These properties and hydraulic formula, when adjusted for fluid density, are used to estimate the pressure drops per mile for each of several different diameter pipelines (4 to 24 inches) under consideration as shown in Figure 3.2. For the initial analysis, it is assumed that $CO_2$ transport pipeline flows with only the inlet pressure provided by compression at the power plants and without intermediate pressure boosting. This assumption is a common assumption for $CO_2$ pipeline projects and allows for future expansion to an ultimate rate with the addition of a booster station.

With an assumed pipeline pressure drop of 1,000 psi, the average pressure drop per mile for the conceptual 200-mile Illinois pipeline under review is 5 psi/mile. Referring to the graph in Figure 3.2, the capacity of various pipeline diameters can be determined for a pressure drop of 5 psi/mile. The delivery flow rate can be determined by entering the graph on the pressure drop axis (ordinate) at 5 psi/mile. The intersection of the 5 psi/mile pressure drop value with the curve for a specific pipe diameter gives the flow rate on the flow rate axis (abscissa). Pipeline capacities for the 200-mile Illinois Basin pipeline (without boosting) are in Table 3.2 (column 2) for pipe diameters ranging from 12 to 24 inches.

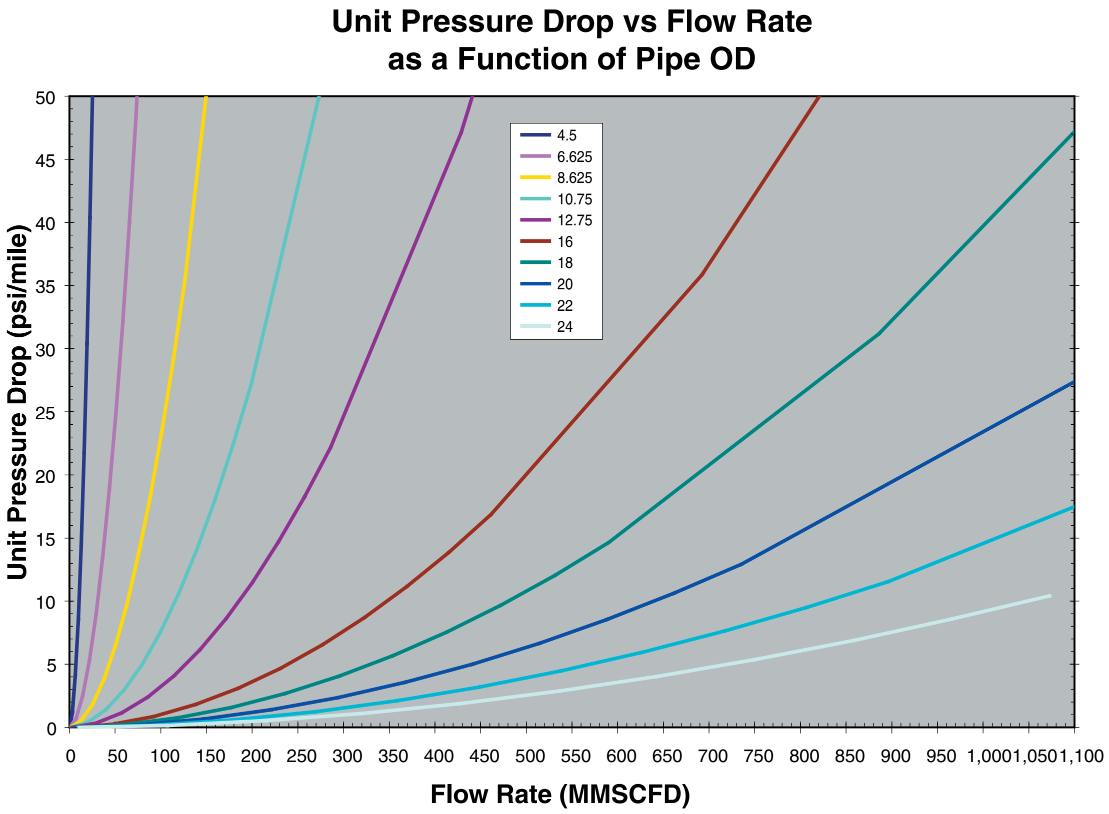

Figure 3.2. Pressure drop per mile for each diameter pipeline

Table 3.2. Flow capacity as a function of pipe diameter and pressure drop

| Pipe diameter (inches) | Flow capacity with inlet pressure only, 1,000 psi pressure drop over 200 miles (MMSCFD)[1] | Flow capacity with 100% boosting at mid-point, 1,000-psi pressure drop over 100 miles | | |
|---|---|---|---|---|
| | | Flow capacity (MMSCFD) | Required BHP for 100% boosting at 100-mile mid-point | |
| | | | BHP/mile | BHP/ 100 miles |
| | | | | |
| 12 | 125 | 190 | | 2,400 |
| 16 | 250 | 350 | | 3,600 |
| 18 | 340 | 490 | | 5,500 |
| 20 | 450 | 650 | | 6,800 |
| 22 | 560 | 840 | | 8,600 |
| 24 | 700 | 1,050 | | 11,000 |

[1] Million cubic feet per day.

If 100% boosting (1,000 psi) were applied at the pipeline midpoint, then the pressure loss would be an average of 10 psi/mile. Starting at the 10 psi/mile value on the ordinate axis, the following flow rate capacities would increase as given in column 3 of Table 3.2. The required brake horsepower (BHP) size for the booster station can be determined by using the graph shown in Figure 3.3. To use the graph, start at the ultimate flow rate for each diameter, read the horsepower per mile where it intersects the diameter, and then multiply by the distance. For the case of a booster station at the midpoint of the hypothetical 200-mile Illinois Basin pipeline, the required BHP per mile and per 100 miles is given in columns 4 and 5, respectively, of Table 3.2 for the pipe diameters under consideration.

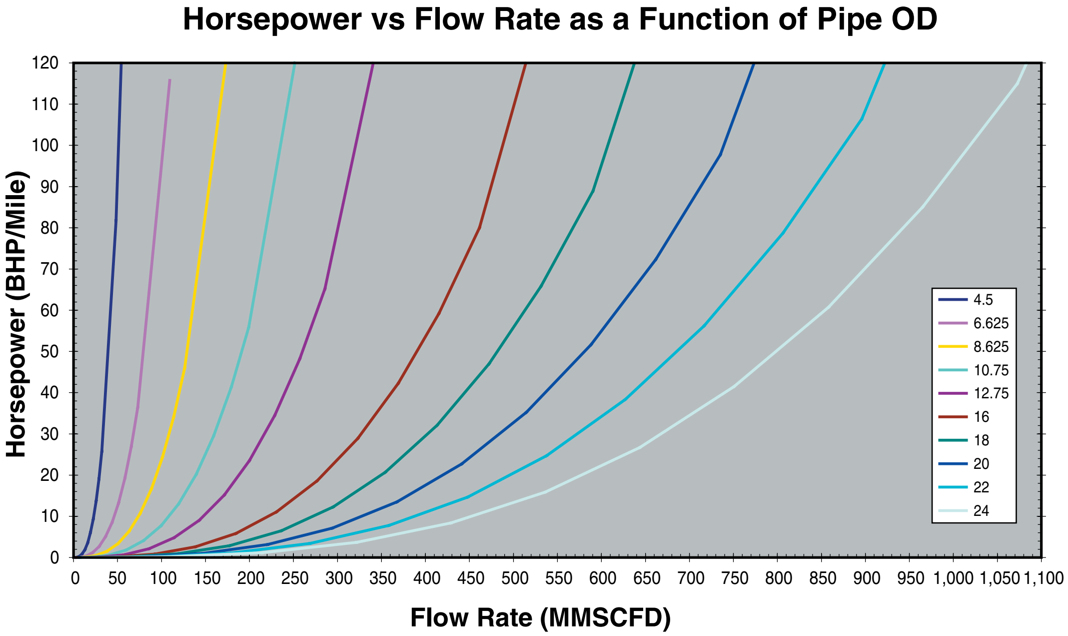

**Figure 3.3. Required brake horsepower size for the booster station**

## 3.3. Design

### *3.3.1. Codes – Federal, State, and Industry*

The design of a $CO_2$ pipeline is very similar to the design of a high-pressure LPG pipeline. The minimum design for a $CO_2$ pipeline is covered by the U.S. Department of Transportation (U.S. DOT), Federal Code 49 CFR, Transportation, Part 195, Transportation of Hazardous Liquids by Pipeline, Section, 195.0, which prescribes safety standards and reporting requirements for pipeline facilities used in the transportation of hazardous liquids or $CO_2$.

Illinois, with the exception of the Illinois Department of Agriculture, does not have regulations pertaining to $CO_2$ pipelines. The Illinois Commerce Commission (ICC) regulates the safety requirement of the gas pipelines in the state but defers the safety regulation of Part 195 liquid jurisdictional pipelines to the U.S. DOT.

The following design elements are presented with a reference to the specific U.S. DOT 195 regulation with a recommended method of applying the regulation to a $CO_2$ pipeline.

### *3.3.2. Pressure Design Formula*

The design parameters for steel pipe are determined in accordance with the following equation (see U.S. DOT, Sec. 195.106, Internal Design Pressure):

$$P = (2St/D)EF$$

where

P = design pressure (psig),
S = specified minimum yield strength (psi),
D = specified outside diameter of the pipe (inches),
t = specified wall thickness of the pipe (inches),
E = seam joint factor, and
F = design factor.

In general, pipeline design pressure and other factors are known, and the wall thickness can be calculated.

*3.3.2.1. Design Factor "F."* This design factor will be determined as a result of conditions or a combination of conditions such as crossings, fabrications, station yards, and special areas (Sec. 195.106, Internal Design Pressure). For the $CO_2$ pipeline under consideration, a design factor of 0.72 would be common for the mainline and lateral line in all areas where normal installation methods and cross-country conditions prevail. A design factor of 0.60 would be used for facilities piping at meter/regulator stations, and a design factor of 0.72 might be used for piping at pipeline launcher or receiver facilities. Section 195.106 allows for a design factor of 0.72 to be used in all cases; however, a more restrictive design factor is common practice for facilities, crossings, and congested areas. Also, the pipe installed at all highway, road, and open cut waterway crossings and in all fabricated assemblies needs to be of sufficient wall thickness and grade to be in compliance with a design factor of 0.60. Directionally drilled waterway crossings and railroad crossings shall be designed using a design factor of 0.50 or 0.60 depending on the length and depth of the crossing.

Based on the conceptual Illinois Basin route of 200 miles, the estimated number of road and river crossings for the pipeline will break down approximately as in Table 3.3.

**Table 3.3. Mileage by design factor to accommodate road and river crossings**

| Route (miles) | Factor | Area |
|---|---|---|
| 180 | 0.72 | Open |
| 15 | 0.60 | Roads and mildly congested urban or commercial |
| 5 | 0.50 | Rivers and congested urban or commercial |

*3.3.2.2. Yield Strength "S."* The specified minimum yield strength for the pipe steel set forth in the manufacturing specifications is used in the design equation that determines the pressure rating of the pipe. The yield strengths measured during mill testing, which will be higher than the rated yield strength, are not used in the calculations. These actual higher yields provide another inherent safety factor in pipeline design. Examples of pipe grades are X42, X52, X60, X65, and X70. The numerical values are the rated yield strength of the pipe steel in pounds per square inch divided by 1,000.

A $CO_2$ pipeline would be constructed with X65 or X70 grade pipe if built in the next 5 years. Over the next several years, higher grade pipe will be introduced into the industry, which could have application in larger-diameter $CO_2$ pipelines.

The seam joint factor E (see Sec. 195.106, Internal Design Pressure) is 1.0 for butt-welding methods under consideration for $CO_2$ pipelines.

### 3.3.3. Pipe Wall Thickness

Design, yield, and seam joint factors, combined with expected pressure, indicate wall thickness for the diameters that are probable for the project. The wall thickness (inches) for a 2,800 psig, X65-grade pipeline is shown in Table 3.4 for the list of outside diameters under consideration and for the common safety factors (0.72, 0.60, and 0.50).

**Table 3.4. $CO_2$ pipeline, 2,800 psig, grade X65 pipe**

| Wall thickness (inch) | | | |
|---|---|---|---|
| Diameter (inch) | Design factor (F) | | |
| | F=0.72 | F=0.6 | F=0.5 |
| 4 | 0.135 | 0.162 | 0.194 |
| 6 | 0.198 | 0.238 | 0.285 |
| 8 | 0.269 | 0.317 | 0.380 |
| 10 | 0.322 | 0.386 | 0.463 |
| 12 | 0.381 | 0.458 | 0.549 |
| 16 | 0.479 | 0.579 | 0.689 |
| 18 | 0.538 | 0.696 | 0.775 |
| 20 | 0.598 | 0.718 | 0.862 |
| 22 | 0.658 | 0.790 | 0.948 |
| 24 | 0.718 | 0.862 | 1.034 |

### 3.3.4. Pipe Wall Thickness Transitions

End preparation for butt-welded sections of pipe having unequal wall thickness varying more than 3/32 inch (0.093 inch) will be taper bored or internally beveled.

### 3.3.5. Pipe Bending Design

The pipeline contractor will perform field bending during the construction of the pipeline system. The maximum allowable bend is up to 1.5 degrees in any length along the pipe axis equal to the diameter of the pipe (e.g., length of 24 inches in a 24-inch diameter pipe). To allow for pipe alignment during fit-up, the minimum straight tangent on a bend should be 4 to 5 ft on each end. For large areas requiring directional changes over a short distance, induction bends can be engineered and purchased in advance. The company will generally provide the contractor with induction bends for side bends greater than a set angle and for areas where major elevation changes are expected.

### 3.3.6. Depth of Cover

The minimum depth of cover required by U.S. DOT 195.24 would need supplementation to meet the the Illinois Department of Agriculture local land use requirements. The required depths will be as per the following:

- Industrial, commercial, and residential areas: 48 inches
- Agricultural and cultivated lands: 60 inches
- Public road surfaces and railroads: 60 inches (or permit requirements)

- Drainage ditches of public roads and railroad crossings: 48 inches (or permit requirements)
- River and major stream crossings: 60 inches (or permit requirements)
- Minor stream crossings, drainage canals, and ditches: 60 inches
- Freshwater supply: 60 inches (uncased)
- Irrigation canals: 60 inches (uncased)

#### *3.3.7. Buoyancy Design*

The pipe should be designed to be negatively buoyant both during and after construction. In general, in locations where the ditch cannot be dewatered during construction, a minimum negative buoyancy of 115% is normally used. Increased negative buoyancy may be required in areas where there is a potential for soil liquefaction or liquefied backfill that has higher than normal specific gravity. Acceptable methods to achieve the required negative buoyancy include ditch dewatering, concrete coating of pipe, or concrete weights (both set-on and bolt-on types).

#### *3.3.8. Horizontal Directional Drill*

A horizontal directional drill (HDD) is used to make river and other crossings where it is impractical or environmentally unacceptable to construct from the surface (see Appendix C1, Figures C1.11 to C1.15). The HDD has a minimum cover below the bottom of a river to prevent breakout of the drilling fluid. Also, the pipe handling equipment at the site limits the exit angle of the crossing, which is especially true for the larger-diameter pipes. For pipe handling, smaller angles are preferred, although a smaller angle increases the possibility of a drilling fluid breakout at the end of the HDD. In addition, the minimum radius of curvature for the pipe must be considered. Larger pipe has a proportionately larger minimum radius of curvature than a smaller pipe. These factors—minimum cover, angle of entry, angle of exit, and minimum radius—combine to give the minimum distance for a crossing. The design of the crossing must take into account geological conditions at the site.

### 3.4. Materials and Equipment

#### *3.4.1. Summary*

By developing a specification with input from the project metallurgist, the engineer, and the purchasing agent and by working only with pre-qualified suppliers, fully adequate equipment and materials can be readily obtained at a reasonable price.

#### *3.4.2. Industry Code*

The U.S DOT CFR Sec. 195 regulates $CO_2$ pipeline safety; this regulation requires that the pipe be provided in accordance with the American Petroleum Institute Code 5L or 5LX. The 5L code has two levels that a pipe can meet. The $CO_2$ pipe should be specified to meet the higher level PSL 2 requirements, as well as additional supplemental requirements, as determined by the engineer and metallurgist.

#### *3.4.3. Line Pipe*

The line pipe is the key material component for any pipeline. For a $CO_2$ pipeline, the line pipe must meet the U.S. DOT regulations just referenced. These regulations allow the pipe to be manufactured by a prescribed manufacturing method with varying grades of steel. In addition, the U.S. DOT regulation requires that the pipe have the proper chemistry and ductility to match the intended services.

### *3.4.4. Method of Manufacture*

Three manufacturing methods are used for $CO_2$ pipelines: seamless, electric resistance welded (ERW), and double submerged arc welded (DSAW).

Each manufacturing method has advantages:

- Seamless pipe, as implied by its name, has no longitudinal seam weld. This pipe is used in plant facilities or fabrication that contains numerous connections. The lack of a seam, which should not align at a weld join, simplifies the layout and welding process.

- ERW pipe is made from flat steel coils. The coils are uncoiled and formed into pipe, and the seam is welded by an electric resistance process. ERW pipe can be made in lengths up to 80 feet, which reduces field handling welding requirements. The maximum diameter of standard wall ERW pipe is 24 inches. The upper limits of wall thickness required for $CO_2$ pipelines are also at the upper limits in the 20- to 24-inch range. Recent 16-inch $CO_2$ pipelines successfully used ERW pipe.
- DSAW pipe can be made in larger diameters than the ERW. The steel is delivered in flat plates, which are then formed into the pipe shape using large hydraulic presses. The seam weld is then made using a submerged arc process. Thick-walled pipe of large diameter is manufactured using this process.

In the past, ERW pipe was not considered to be reliable enough for high-pressure $CO_2$ service because of seam failure issues with the higher grades of steel needed for $CO_2$ pipeline. However, ERW pipe from premium ERW mills has been used on recent $CO_2$ pipeline projects. For diameters in the 24-inch range, DSAW mills should be considered to ensure the best quality, because the price difference may not be significant.

### *3.4.5. Grade of Steel*

Steel pipe can be made of various grades of steel as long as it meets the American Petroleum Institute (API) 5L or 5LX standard. API 5L is for low-grade steels, and APL 5LX is for the higher-grade steels that are now commonly used for $CO_2$ pipelines. The grade defines the tensile strength of the steel (e.g., grade X60 is 60,000 psi) used to form the pipe. The higher the grade, the more stress the steel can safely withstand before it yields. For a specific high pressure, the higher grade requires a thinner wall thickness and therefore has a reduced material, field handling, and welding costs. Over the years, the definition of "high" grade has gradually increased such that X65 or X70 is presently considered to be high grade. Tensile strength should be specified to limit the range of tensile strengths that are acceptable to less than 20,000 psi over the specified grade. In designated geohazard areas, it may be necessary to control the yield-to-tensile strength ratio to ensure ductile behavior.

### *3.4.6. Chemistry of Steel*

Steel should be made using the basic oxygen or electric furnace process. The steel should be fully killed and made with a fine-grain, clean steel process and treated for inclusion shape control. If any $H_2S$ exists in the $CO_2$, then the steel will need to be tested for hydrogen-induced cracking (HIC). This requirement limits the mill sources for the steel supply. Specifying certain maximums and minimums for carbon equivalent, sulfur, and other components is also standard operating procedure.

### *3.4.7. Toughness/Ductility of Steel*

Based on the planned operating pressure, temperatures, and pipeline diameter, a specified toughness of the steel is needed to prevent brittle behavior of the pipeline system. This toughness specification level must take into account the refrigerating effects of the $CO_2$ during operational upsets and possible small leaks. The chemistry and the overall manufacturing process determine the toughness of the steel. Toughness is measured by use of a Charpy impact test and by a drop weight tear test. This latter test can be considered optional but should be conducted on a sampling of the pipes to determine the need for further testing.

### *3.4.8. Inspection and Mill Quality*

The full ERW weld seam needs to be inspected by non-destructive testing that does not include radiographic techniques. All API 5L pipe is hydrotested in the mill. It is recommended that pipe be tested to 95% yield or greater. For large pipe, it is common to increase the hold time to 15 seconds and to require that the first pipes from a coil be measured for excessive deformation. The pipe should be inspected using longitudinal, long-wave transducers to identify pipe body laminations. A final test should be run to ensure that the pipe does not contain any residual magnetism.

### *3.4.9. Valves*

Valves should be full-opening, ball-type, manufactured and tested in accordance with a detailed specification that is specially written for $CO_2$ service. In general, most $CO_2$ pipelines use ANSI 1500-rated valves. Valve trim should be suited for NACE MR-01-75 with elastomers that are suitable for the requirements for $CO_2$ service. Generally, the elastomers are a harder type, such as Viton or Nitrile, rather than the softer rubber type, such as Buna.

The valve used for the mainline service should be a through-conduit valve to ensure passage of inspection and cleaning pigs. The mainline valve should have weld-by-weld end connection. Side tap valves connected to the pipeline should be welded by a ring joint flange end connection, which results in a pipeline that is welded from one end to the other and decreases the chances for nuisance leaks that could result in an unnecessary shutdown.

Unless impractical, mainline and lateral valve sites are located near roads to maximize ease of access; however, locations near populated areas should be avoided if possible due to potential for vandalism. In areas of high population density, the valve spacing along the mainline should be decreased. The design factor in the hoop pressure equation for a valve setting is usually 0.5 or 0.6, but the internal diameter should be checked for passage of internal inspection pigs.

### *3.4.10. Meters*

The metering for $CO_2$ custody transfer is usually by orifice metering set up in the same way as a natural gas or an LPG ethylene metering facility. The metering is based on orifice differential pressure. This differential pressure and temperature are inputs to an onsite computer that calculates the amount of $CO_2$ being delivered. The regulators are usually on the downstream side to protect the facility from over pressure and to ensure adequate backpressure in the pipeline. All of the electronic care is housed in an enclosed building. An example of the metering facility recently designed by Universal Ensco is shown in Figure 3.4.

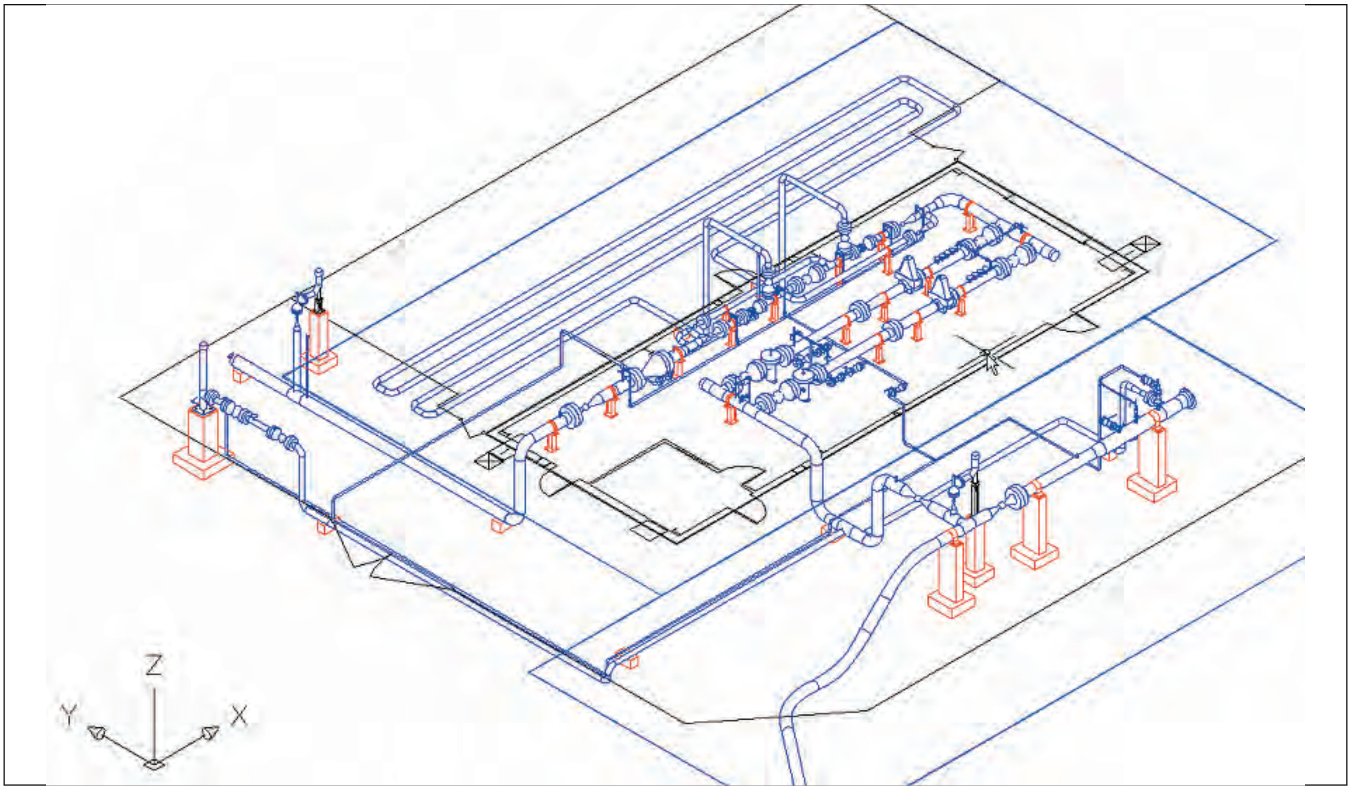

**Figure 3.4. Anadarko $CO_2$ meter-regulator station**

### *3.4.11. Pumps*

If a flow rate is required that is in excess of that available, as free flow from the injection plant to the delivery points, an intermediate pump station can be installed. At the pump station, an electric drive, horizontal split-case multistage centrifugal pump can be installed. The pump can be controlled by a variable speed motor control system or by the installation of a pressure control valve. The pump seals need to be designed to account for the low lubrication provided by the $CO_2$.

## 4. Permits

As with any construction project, permits from a number of jurisdictional agencies are required for the $CO_2$ pipeline. The permits can be classified as business, environmental, and utility permits. It is vital that the permit process be started as early as possible. The process requires the involvement of environmental, right-of-way, and engineering specialists. An overview of the necessary permits and the planning required to obtain them follows. The discussion is based on recent experience in Illinois.

### 4.1. Business Permits

### *4.1.1. Illinois Commerce Commission*

The Illinois Commerce Commission (ICC) is the prime permitting agency for pipelines in Illinois. Unless granted by the federal government, the ICC is the sole authority by which an intrastate pipeline can obtain eminent domain rights. The importance of the power of eminent domain is discussed in Section 5, Right-of-Way.

For a successful application, the ICC grants a "certification in good standing" that allows the pipeline to begin the eminent domain process. To obtain the certification, the ICC requires the application be completed based on Illinois statutes. The application must

1. be properly completed;
2. demonstrate that the applicant is fit, willing, and able to accomplish the project in accordance with Illinois statutes;
3. demonstrate that the project meets the standards for the public's convenience; and
4. demonstrate that the project is a "public necessity."

The ICC works in an independent manner and in the past has exercised its authority in a judicious manner. The application process is expected to take approximately 18 to 24 months.

### 4.2. Environmental Permits

Regardless of the exact route selected, a north-to-south sequestration pipeline in the Illinois Basin will inevitably traverse environmentally sensitive areas. Specifically, the route will cross a number of watercourses that are under the jurisdiction of the U.S. Army Corps of Engineers (U.S. ACOE). The U.S. ACOE will require a permit as part of its procedural review. During the review process, it will ask for comments from a number of other governmental agencies. Those agencies will be looking for specific items of compliance and mitigation in their area of special concern. The U.S. ACOE will review comments and, if reasonable, will include them as stipulations in the permit that it ultimately grants.

The USACOE permits are as follows:

1. Section 404, Dredge and Fill**:** Jurisdictional wetlands crossed along the route. This section is interpreted liberally by the U.S. ACOE and most applications are approved.
2. Section 404, Dredge and Fill, Section 10, Navigable Waters: Crossing Streams: Major streams are generally permitted as horizontal directional drilled crossings unless there are technical reasons that a crossing cannot be made in this manner. Generally the U.S. ACOE will seek input from the other agencies such as the Office of Fish and Wildlife, Illinois Department of Natural Resources.

The following environmental permits would be needed for the $CO_2$ pipeline project:

1. Illinois Department of Natural Resources: Rivers, Lakes and Streams Construction Permit
2. Illinois Department of Natural Resources: License to Cross
3. Illinois Environmental Protection Agency: NPDES (National Pollution Discharge Elimination System): Controls the discharge of silt into the surface water by imposing effluent limitations.
4. Illinois Environmental Protection Agency: NPDES Discharge Permit, discharge of liquids to streams during the construction of the pipeline.
5. Illinois Environmental Protection Agency: Construction Storm Water Permit, runoff from the right-of-way during construction of the pipeline.
6. Illinois Department of Agriculture: Agriculture Mitigation Agreement. Permit is not mandatory but is conventionally agreed upon. Note: The ICC looks upon this as a very important agreement to ensure public convenience.

7. County Land Use Department: Land Use Permit. In rural areas a pipeline does not impact normal usage; however, in an urban area, this could be an issue.
8. County Engineering Department: Site Development (Erosion Control) Permit

#### *4.2.1. Plans for Pollution Prevention*

The following pollution prevention plans must be submitted for the project:

1. Illinois EPA Pollution Prevention Plan Outline
2. Environmental Construction Mitigation Plan
3. Spill Prevention, Containment and Countermeasure Plan
4. Hazardous Waste/Contaminated Soil Plan
5. Inadvertent Release of Drilling Mud Plan
6. Plan for Unanticipated Historic Properties and Human Remains (Unanticipated Discoveries Plan)

### 4.3. State and County Highways and Railroad Crossing Permits

Highway and railroad crossing permits are generally routine and can be handled as a matter of course during the project. The large number of permits needed requires organization and on-site meetings with the agencies. Railroad companies require extra depth for crossing due to impact loading, and impact loading calculations must be submitted with the application.

## 5. Right-of-Way

### 5.1. Pipeline Right-of-Way/Easement Definition

A pipeline right-of-way is a strip of land over and around pipelines where some of the property owner's legal rights have been granted to a pipeline company. A right-of-way agreement between the pipeline company and the property owner is also called an easement and is usually filed in the public records with property deeds. Rights-of-way and easements provide a permanent, limited interest in the land that enables the pipeline company to operate, test, inspect, repair, maintain, replace, and protect one or more pipelines on property owned by others.

### 5.2. Negotiations

A company right-of-way representative will be the first contact between the pipeline company and the landowners. The objective of the first contact is for all landowners to understand all proposed features of the pipeline, including the alignment, underground depth, pipe size, temporary and permanent width of the easement, and aboveground equipment prior to construction. During these contacts, the representative will explain the project and the proposed process between the pipeline company and the landowners. If possible, the company representative will attempt to reach agreement with the landowner for negotiating an easement and will request permission to survey and stake the preliminary route for environmental, engineering, and construction evaluations.

The company will compensate each landowner for these rights:

1. The privilege of establishing a permanent easement across the landowner's property. Payment for the easement will be based on market value principles and the number of acres required. Although the company will obtain a permanent easement, the landowner will retain ownership and use of the land.
2. Damages to crops, grazing lands, timber, or any structures directly caused by the construction and maintenance of the pipeline. Construction damages will be paid on the area affected by the actual construction. The settlement for damages to crops either can be paid in advance, based on records of local yields, or can be paid after construction, based on the actual crop losses.

After the conditions and the amount of compensation for an easement are reached and the easement agreement is executed, payment will be issued to the landowner.

### 5.3. Eminent Domain

The company will negotiate with each landowner under a voluntary "willing buyer-willing seller" process. Should the process break down, the company could have the option of seeking help though the Illinois court system. This option is available only after the ICC has approved the project as one that is necessary and beneficial to the state. If the ICC approves the project and an agreement with the landowner cannot be reached, the easement for the property may be acquired using the state's eminent domain authority. A more complete discussion of this permit is covered in Section 4, Permits.

The procedures within the acquisition process leading up to the use of eminent domain are very structured. The pipeline company must demonstrate that it has negotiated in good faith on the alignment submitted to the court in a request for eminent domain. This demonstration would include being able to document that the route was varied to accommodate the special requirements of the landowner and that a fair offer was submitted. It is common practice to obtain an appraisal and an abstract for each property condemned. A registered surveyor prepares a certified plat detailing the route, and this plat is included in a letter to the landowner as part of the documentation in the final offer.

Filing a complaint in circuit court starts the condemnation process. The complaint must set forth authority, purpose, description, and names in regard to the pipeline and property being condemned. The property owner may raise a challenge as to the necessity of taking the property, whether the taking is for public use, whether the condemner made a bona fide attempt to offer fair compensation, or on any other basis relating to the right of the plaintiff to condemn.

These challenges are raised as a "traverse and motion to dismiss." The traverse denies the allegations of the complaint; the motion to dismiss points out certain specific objections that the property owner has to the condemnation. The judge rules on the issues raised in the traverse and motion to dismiss. There are no procedural rules that establish when a defendant must file a traverse and motion to dismiss or when a court must adjudicate the issues raised by such a pleading. As with other civil litigation in Illinois, both the condemner and the landowner in an eminent domain action are subject to discovery, including interrogatories, document requests, and depositions. Matters typically subject to discovery include the designation of expert witness, information related to witnesses, and the basis for the appraiser's opinions regarding value. Assuming the traverse and motion to dismiss is denied, the case proceeds to a trial to determine just and reasonable compensation. If either side so requests, a jury trial is held; otherwise, the trial is presented to the judge. A trial may take several days. The only issue is the amount of compensation. Upon the award, both the condemner and the landowner may appeal the determination.

Regardless of which party appeals the judgment, the condemner may enter the condemned property as long as a bond is posted in the amount of the compensation.

The estimated time for condemnation proceedings in Illinois varies by county and is influenced by the number of condemnations in the county. Because the proceedings are formal trials, there are multiple opportunities for the defendants to introduce delays through procedural maneuvering. A reasonable duration for planning purposes for the legal portion of the condemnation proceedings is 12 months.

### 5.4. Right-of-Way Widths

The right-of-way agreement or easement defines the width of the permanent easement and any temporary work space above the permanent easement to support the operations necessary to construct the pipeline. The easement will describe additional areas where extra temporary work space is required for construction at crossings and other special areas. The pipeline is not necessarily located in the center of the easement.

Any pipeline can be installed using a limited right-of-way, but the cost of construction and operation increases dramatically in such a case. Prior to starting any permitting and environmental activity, the width of the right-of-way should be designed to take into account such things as landowner operations, environmental conditions, anticipated weather conditions at the time of construction, and the diameter of the pipeline.

In the heavily farmed areas of Illinois, the width of the right-of-way needs to be increased to accommodate the handling of topsoil and the crossing of drain tiles. An approximation for permanent easement and temporary work space needed for a central Illinois pipeline project are shown in Table 5.1. These widths are based on knowledge of the existing land use, environmental requirements, and experience on recent projects in Illinois.

**Table 5.1. Pipeline right-of-way easement and work space in southern Illinois**

**Pipeline Right Of Way Easement And Work Space**
**Southern Illinois**

| Dia (inches) | Permanent Easement (ft) | Subtotal | | | Allowance For Extra Temporary Workspace (%) | Total Acres / Mile |
|---|---|---|---|---|---|---|
| | | Extra Temporary Work Space (ft) | | Acre/mile | | |
| 4 | 25 | 25 | 50 | 6 | 10 | 6.6 |
| 6 | 25 | 25 | 50 | 6 | 10 | 6.6 |
| 8 | 30 | 30 | 60 | 7.3 | 10 | 8 |
| 10 | 30 | 30 | 60 | 7.3 | 10 | 8 |
| 12 | 30 | 40 | 70 | 8.5 | 10 | 9.3 |
| 16 | 40 | 50 | 90 | 10.9 | 10 | 12 |
| 18 | 40 | 50 | 90 | 10.9 | 10 | 12 |
| 20 | 40 | 50 | 90 | 10.9 | 10 | 12 |
| 22 | 40 | 50 | 90 | 10.9 | 10 | 12 |
| 24 | 40 | 50 | 90 | 10.9 | 10 | 12 |

# 6. Construction

## 6.1. General

Installing a long distance pipeline is accomplished in an assembly line process that moves along the pipeline route. The assembly line consists of equipment and personnel organized into what is termed in the industry as a pipeline spread. The spread is staffed and supplied based on the pipeline diameter length, schedule, terrain, permit and land requirements, and the anticipated weather and groundwater conditions. The spread begins at one end of the pipeline and works toward the other end with the specialty crews commencing work chronologically one after another. Once all of the crews are operating smoothly, the spread moves along the pipeline route at a rate of 1 to 2 miles per day. During a peak production period, a spread can install 3 miles per day.

Generally, the pipeline owner will contract with an engineering procurement and construction management company to supervise a pipeline construction contractor during the construction phase of the project. The construction contractor is responsible for organizing the construction spreads and the specialty crews performing work within each spread. The order in which the specialty crews perform work can vary due to terrain, weather considerations, groundwater conditions, and permit stipulations, but they generally follow this procedure:

1. Pre-construction activities and survey re-stake
2. Clearing and grading
3. Crossing, road and stream
4. Stringing and bending
5. Ditching
6. Pipe gang, facing and lineup
7. Firing line, welding and joint coating
8. Lower in and backfill
9. Cleanup
10. Tie-ins and fabrications
11. Testing and drying

These pipeline construction spread activities are illustrated in Appendix C1.

The construction spread will be staffed to support the desired rates of progress. Typically, the smaller-diameter pipelines (4 inches) will have a staff of about 150; a large-diameter pipeline (36 inches) will have about 500. A support staff monitoring the contractor for contract compliance, quality, landowner issues, and environmental will be staffed at about 20% of the construction staff.

After the ditch, pipe gang, firing line, and lower backfill crews reach the end of the spread, some portion of these crews and their equipment are rolled back to support the tie-in and cleanup operations.

## 6.2. $CO_2$ Welding and Refrigeration Issues

The properties and characteristics of $CO_2$ require the modification of standard pipeline construction practices in two areas: weld procedure development and the line fill process.

A welding procedure must be developed that provides adequate toughness to resist brittle failure in a leak event while maintaining the necessary field practicality. The project engineer, metallurgist, and a construction representative can work together to design a qualified procedure. A weld procedure is qualified by testing it on the pipe as shown in Figure 6.1.

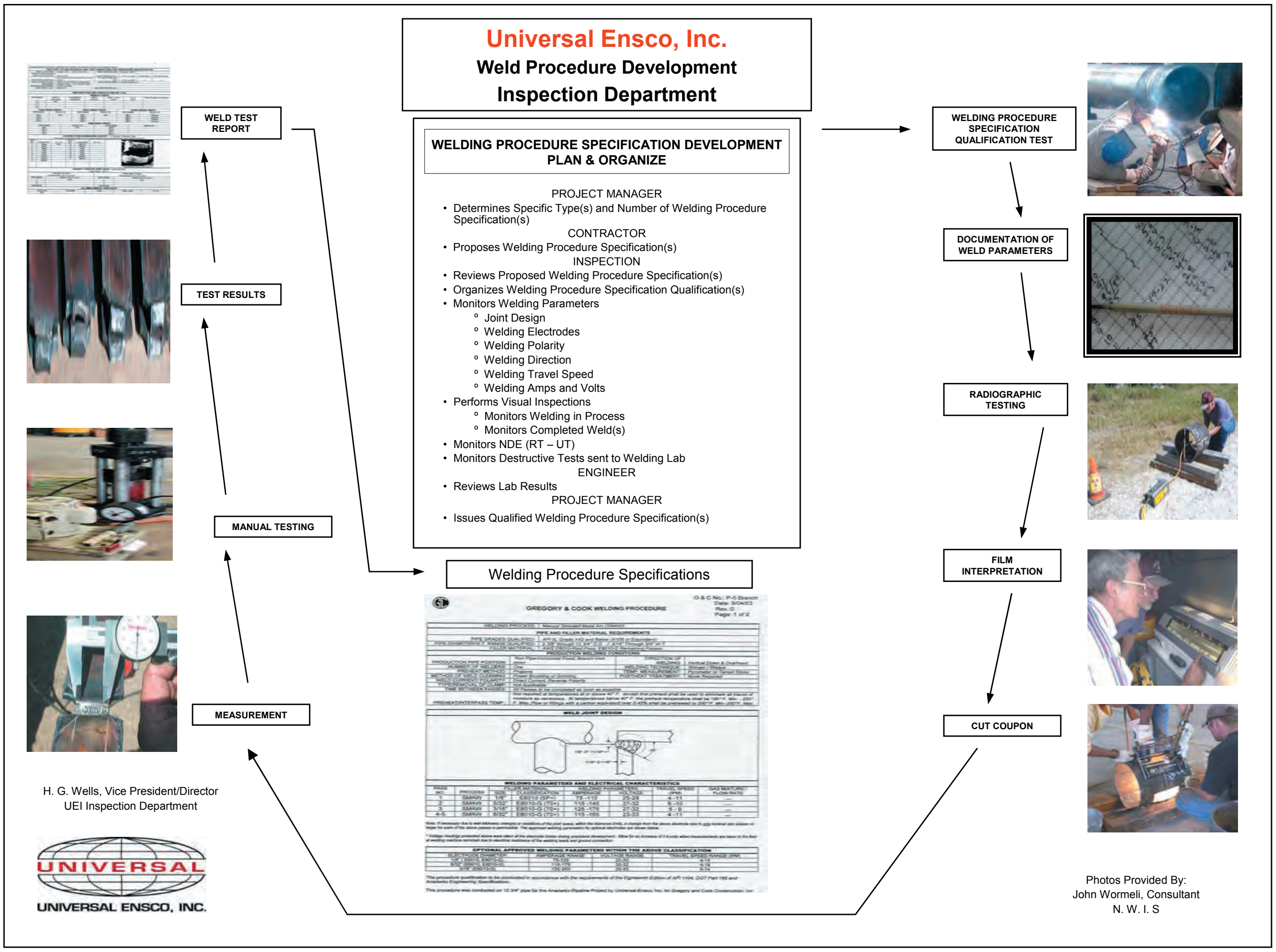

**Figure 6.1. Process for testing the weld procedures**

During the filling process, the refrigeration property of $CO_2$ must be considered to avoid dry ice. The filling process has two purposes. The first purpose is purging the pipeline of dry air by using $CO_2$, and the second is to pack the pipeline with $CO_2$ to the minimum operating pressure. The purging part of the filling process can also be handled in two ways. Either $CO_2$ can be delivered from the process as a gas at ambient temperature, or it can be throttled through a control valve into the pipeline. In the throttling method it is important that the refrigeration effect of $CO_2$ be considered (Figure 6.2).

**Figure 6.2. $CO_2$ venting during pipeline purging**

The refrigeration effect can be mitigated by using a line heater to raise the temperature of the $CO_2$. The alternative to the line heater is to maintain a backpressure above approximately 100 psig. The arrival of the $CO_2$ will be visible at the vent site as the vapor that condenses when the refrigerated $CO_2$ begins to discharge. $CO_2$ could form dry ice (i.e., solid $CO_2$), if the pressure is lowered too fast (Figures 6.3 and 6.4).

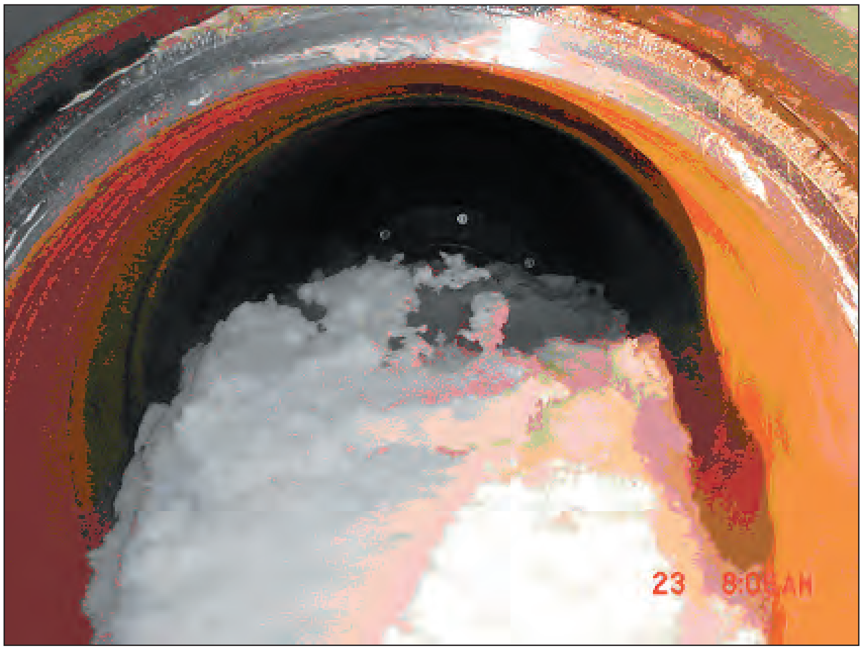

**Figure 6.3. Dry ice formed by depressuring too rapidly**

**Figure 6.4. Pig damage from depressuring too rapidly**

### 6.3. Agricultural Issues in Illinois

The terrain of central and southern Illinois is flat to rolling with a high percentage cleared for farming operations. Unfortunately, the subsurface and these farming operations can cause major difficulties in pipeline construction, including

1. unstable ditch conditions,
2. intermittent high groundwater in certain areas, and
3. drain tiles used for farming low areas.

A thorough investigation into the subsoil conditions should be made during the planning phase to determine the extent of ditch stability issues that may arise. If these ditch issues are expected to be consequential, then the right-of-way should be laid out to allow for the welding to proceed ahead of the ditching so that the ditch is open for the minimum amount of time and is closed up at the end of each day. If the conditions are isolated and dependent on weather conditions, then the contractor can slope the ditch and use a backhoe or crane to clam the ditch in advance of the lower-in (Figure 6.5).

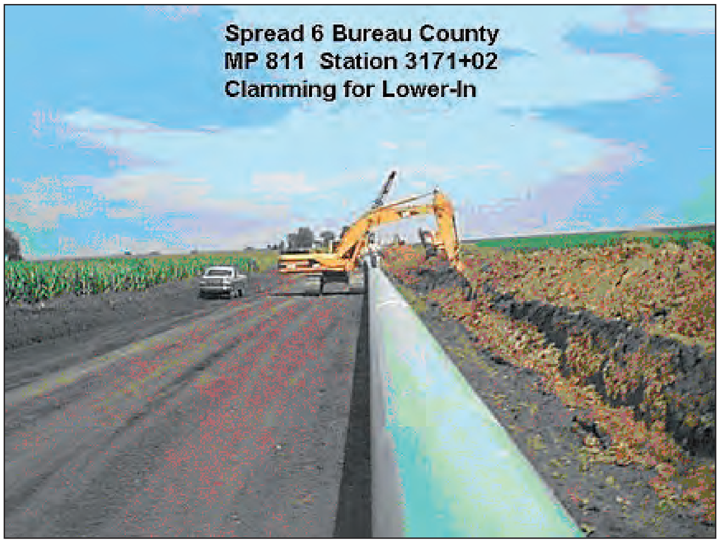

**Figure 6.5. Backhoe reaches over pipe to clam out ditch**

Where the water table is high, water will accumulate in the trench during the weld operation. Prior to lowering, most of this water will be pumped out of the trench, but during lower-in and before backfill, the water will continue to accumulate. To isolate the water and prevent it from being carried forward by the backfill operation, the contractor will plug the pipe in intervals (Figure 6.6). This procedure will also stabilize the pipe during backfill operations.

**Figure 6.6. Crane plugs pipe during backfall operations**

Set-on weights provide a solution in areas that cannot be effectively dewatered (Figure 6.7). The weights are costly and require logistical consideration. First, they must be manufactured in advance and in sufficient quantity to be available when needed. Second, before the pipe is lowered in, an adequate quantity of weights must be delivered to the right-of-way at the necessary locations. Experienced supervisors must oversee the field situation to ensure that the team uses the minimum number of weights necessary to maintain the proper cover.

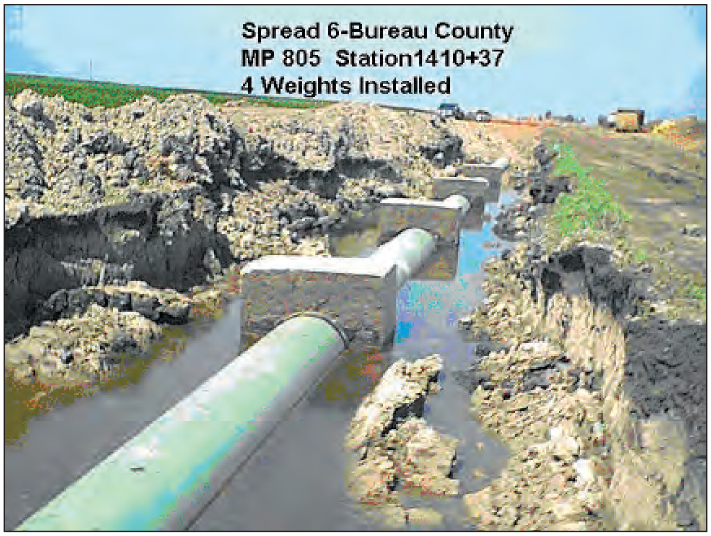

**Figure 6.7. Set-on weights in a low areas where sand subsoil layer is on top of a deep layer of clay subsoil**

The integrity of the existing drain tiles must be maintained across the full right-of-way and during the entire construction period. As shown in Figure 6.8, the string crew is working over the existing drain tiles at a location where temporary drain tile segments are in place behind the ditching operation. These temporary tiles will be removed during lower-in and replaced with permanent repairs after rough backfill is completed. If topsoil on the working side of the easement (Figure 6.8) is removed, the probability is increased that the heavy equipment will damage the drain tiles.

**Figure 6.8. String crew working over existing drain tiles where temporary drain tiles (white segments) are in place**

Areas with poor soil conditions at crossings can be determined in advance by interviews with county engineers, interviews with farmers, and by completing geotechnical studies. The severity of the poor soil situation will depend on the level of groundwater at the time of crossing. Contracts can be written by the engineer to risk-share the impact of the unknown groundwater situation with the contractor.

**Figure 6.9. Road bore in a sandy area with high groundwater showing required sheet pile and ground dewatering systems**

If possible, issues such as the ditch, high groundwater, and drain tile problems shown in Figure 6.5 through Figure 6.9 are best addressed, and difficulties minimized, by selecting a construction period beginning mid-summer and ending in the fall. During this period, rain patterns are favorable. In addition, the warm temperatures and the growing vegetation combine to minimize the impact of rainfall events. This time window of July through October does, however, present several minor issues that need to be considered:

1. The short summer construction season limits each spread length to 100 to 200 miles. Two contractors can mobilize in late June for kickoff after July 4, finish the mainline by September 15, and then test, tie-in, and cleanup by October 31.
2. A large-diameter pipeline requires that pipe be delivered to the pipe yard in April when frost-law road hauling restrictions could be in place. The yards need to be located along major roadways to limit the possibility of being subject to frost regulations and to mitigate exposure to the extra traffic safety issues related to poor visibility associated with the height of the corn along the roadways.

## 7. Operations

$CO_2$ pipelines have operated with very few incidents. Statistics on pipeline incidents for both natural gas and hazardous liquid pipelines, which include $CO_2$, are available from the Office of Pipeline Safety, U.S. DOT. Statistics for the period from 1986 to 2001 on pipeline incidents in the United States are summarized in Table 7.1. The reported safety record of $CO_2$ (zero fatalities and zero injuries) was excellent for this period.

## Table 7.1. Pipeline incidents in the United States, 1986 to 2001

| Pipeline | Natural gas transmission (1986–2001) | Hazardous liquids (1986–2001) | $CO_2$ (1990–2001) |
|---|---|---|---|
| Incidents (no.) | 1287 | 3035 | 10 |
| Fatalities (no.) | 58 | 36 | 0 |
| Injuries (no.) | 217 | 249 | 0 |
| Property damage (US$) | 285,300,000 | 764,200,000 | 469,000 |
| Incidents per 1,000-km pipeline (no./yr) | 0.17 | 0.82 | 0.32 |
| Property damage per 1,000-km pipeline (US$/yr) | 37,000 | 205,400 | 15,200 |

The envisioned $CO_2$ pipeline system will consist of receipt facility lateral pipelines, a main pipeline, booster facility if needed, and delivery lateral pipelines facilities. The operation of these facilities will be accomplished by a properly equipped staff of specialists based at a location removed from the pipeline and its facilities.

Personnel will make regular visits to all parts of the unattended facilities to ensure a safe operation. These unattended facilities are monitored 24 hours per day by a supervisory control system. The pipeline facilities are at one end of the system, and the supervisory control center is at the other end. The pipeline flow conditions and the facilities (valves and meters) are monitored and controlled from the control center. Trained pipeline operators staff the center 24 hours per day. The operators look for upset conditions and data inconsistencies, and they monitor pressure and flow conditions that could indicate leaking.

The public can contact the pipeline control center by telephone if they observe an emergency condition. The control center can shut down the pipeline if it is determined that leaking exists. The control center will be in contact with the receipt and delivery operations to ensure that the incoming and outgoing pipeline flows match those specified. The $CO_2$ line will have a considerable amount of leeway in “input versus output” due to the compressibility of the $CO_2$. This compressibility will allow the pipeline to be “packed” or “unpacked” to support both planned and unplanned operational changes.

To avoid the most common source of leaks—“third-party” damage—and to ensure prompt reporting of leaks, the pipeline will conduct a “public outreach” program. The pipeline emergency phone number is posted and maintained on signs at all pipeline, road, railroad, and water crossings. Public agencies such as fire, police, and civil defense are kept informed of the pipeline location, the telephone numbers of the control room, and the telephone numbers for the local pipeline operating personnel. Residents who live along the pipeline route will periodically be mailed a flyer that identifies the pipeline company; provides facts on the product characteristics; advises residents on how to determine the location of the pipeline; gives information describing the method of identifying a leak; describes procedures in case of a suspected leak; and provides the control room contact information in the event of an emergency. The flyer is also sent to local contractors who are likely to be excavating near the pipeline. A pipeline company Web site also provides all of this information.

In another effort to avoid damage by third parties, the operating company will be enrolled in a “call-before-you-dig” program. In general, these programs require all persons to call a “call-before-you-dig” center at least two working days prior to the start of excavation. They require the pipeline companies to be members as well. The program in Illinois is called JULIE. JULIE (Joint Utility Locating Information for Excavators), also known as the “Illinois One-Call System,” is a not-for-profit corporation that provides professional and non-professional (i.e., homeowners) excavators with a toll-free number (1-800-892-0123) for locating and marking of underground facilities at no charge to the excavator. JULIE serves as a notification service for underground facility owners, taking information about planned excavations and distributing this information to its membership. Following this notification, it is the responsibility of each facility owner to mark the location of the relevant underground facilities at the excavation site. JULIE neither owns nor marks underground facilities.

Pipeline maintenance crews also work to ensure the integrity of pipeline infrastructure. Qualified personnel hired or contracted by the pipeline company ensure that welding operations, valve inspections, pipeline repairs, corrosion prevention system checks, and electronic equipment maintenance (among many other functions) are performed safely and according to pre-established procedures. These people work on the daily maintenance of the system and to assist with major pipeline repair and replacement projects. Pipeline employees are also trained in emergency response procedures and work with local emergency responders if a pipeline accident occurs. Venting of a $CO_2$ pipeline is required during filling and some depressuring operations (e.g., Figure 7.1). Prior to venting, local agencies and landowners should be notified.

**Figure 7.1. $CO_2$ venting from mainline block valve site**

A company or a team that is built from the ground up can accomplish the operation of the pipeline, or it can be subcontracted to a third party. The grassroots operations can be an independent, stand-alone operation, part of the $CO_2$ capture plants, or part of the $CO_2$ geological storage operation.

Because of the complexities of operating a pipeline in a heavily farmed area such as central and southern Illinois, a contractual arrangement for operation (and possibly ownership) with a relevant company is a better option than forming a new pipeline company. Relevant companies would be those companies that are currently operating $CO_2$ and/or LPG pipelines in the Midwest. A partial listing of companies operating $CO_2$ pipelines includes BP, Kinder Morgan, Transpectos, Exxon Mobil, Trinity LLC, and Anadarko. A partial listing of Midwest LPG pipeline operators includes Texas Eastern Products Pipeline, Marathon, Ashland, Exxon Mobil, Enterprise, Koch, and Buckeye. Clearly, a number of companies exist that could be considered as possible operation candidates for the operation of the $CO_2$ pipeline.

The LPG companies that operate in Illinois or nearby areas are the best candidates for this operation for the following reasons. First, companies are accustomed to operating in accordance with the previously discussed U.S. DOT 195 standard. Second, they understand the characteristics of high vapor pressure liquids similar to $CO_2$. The third and most important reason is that they have a base of operations in a heavily farmed area that can be expanded to accommodate the $CO_2$ pipeline operation much more economically than can be done by a newly formed pipeline company.

## 8. Cost Estimate

### 8.1. Introduction

The previous sections present the many elements that impact the cost for the $CO_2$ pipeline system under consideration. Once a proposed route has been selected and the distance for the pipeline lay has been determined, the primary cost consideration is the selection of the pipe diameter. Pipeline diameter is the key variable in establishing the flow capacity of the system. To clearly demonstrate the relationship between pipeline diameter and cost, a unit cost or per mile cost was developed for right-of-way, materials, and construction using varied pipeline diameters. These costs and the basis for their determination are discussed in the sections that follow. Using the total per-mile costs, a percentage factor, based on experience, was applied to account for the additional services necessary to install the pipeline. For pump stations, the cost per brake horsepower (BHP) was developed based on the costs for similar pump stations currently under construction in Pennsylvania. The pipeline operating cost projected for a range of diameters is based on a current estimate for a similar LPG pipeline project in Illinois. Maintenance cost for the pump stations was estimated using data provided by industry contacts, and power costs were calculated using an industry rule-of-thumb.

### 8.2. Pipeline Right-of-Way Cost

The cost to acquire right-of-way is directly related to the value of the property plus the costs associated with the interruption of business currently associated with the property. Although fair price can always be legitimately debated, experience has shown that a reasonable budget for right-of-way easements can be developed by using the actual value of the land that is impacted by the construction process. The cost associated with interruption of business in agricultural areas can be calculated based on an estimate of the years of crop loss. Combining these two cost elements gives a reliable subtotal that should be reviewed and factored for possible condemnation costs. Due diligence should be applied to this aspect of the process since the expense of condemnation proceedings can be considerable.

Based on the conceptual route, a limited analysis of recent land sales and crop production data was performed for the counties involved. The value of farmland is a function of the soils/productivity, demand, type of end user, the investment market, and the interest of the seller. Soil productivity is maintained in the county Supervisor of Assessments Office. Each parcel is evaluated, and a record of the productivity index is established. This step is key in the assessment of the land value and a major element in the appraisal process that is initiated prior to the sale of the property. The productivity index is reflected in the crop yield data shown in Table 8.1. That information shows that the relative quality of soils, production, and the land values are highest at the northern end of the conceptual route and tend to decrease toward the southern end of the route. There are, however, potential exceptions within each county.

**Table 8.1. Land value and crop information, Illinois, as of July 23, 2004**

| County | Crop | Yield (bushel/acre) | Land value ($1,000) |
|---|---|---|---|
| Tazewell | Corn | 151 | 3–5 |
| | Soybeans | 52 | |
| Logan | Corn | 154 | 3–5 |
| | Soybeans | 52 | |
| Sangamon | Corn | 194 | 3–4.5 |
| | Soybeans | 48 | |
| Christian | Corn | 140 | 4.5+ |
| | Soybeans | 49 | |
| Montgomery | Corn | 127 | 3–4 |
| | Soybeans | 45 | |
| Shelby | Corn | 151 | Best, 3.5; lower, 2.5–3 |
| | Soybeans | 52 | |
| Fayette | Corn | 86 | East edge, 2.8–3.2; lesser quality, 2–2.5 |
| | Soybeans | 31 | |
| Marion | Corn | 133 | 1.5–2 |
| | Soybeans | 28 | |
| Clay | Corn | 69 | 1.2–1.8 |
| | Soybeans | 23 | |
| Wayne | Corn | 78 | 1.5; higher for bottomland |
| | Soybeans | 25 | |

### *8.2.1. Land Values*

For the 200-mile conceptual case, the application of the values in Table 8.1 results in the following estimate for land values: $3,750/acre for the first 120 miles and $2,850/acre for the final 80 miles. Overall, the cost for land is approximately $3,150/acre.

### *8.2.2. Crop Loss Payments*

In the area where construction occurs, the crop loss payments for the first year are for a 100% crop loss. There is considerable debate regarding the impact to crops in subsequent years. There is also a lack of unanimity on the issue of determining an appropriate price for the crop in subsequent years. It is safe to budget 100% for damage in Year 1, 75% for residual damage in Years 2 and 3, and 50% for residual damage in Years 4 and 5. This budget results in a projected damage payment of 3.5 times the average crop yield of the acreage occupied during construction. This compensation is often disbursed at the initiation of construction to expedite the right-of-way acquisition process.

Corn is generally a more valuable commodity than soybeans. Therefore, the most conservative method of estimating crop loss values involves using the per-bushel price of corn. Per-bushel corn prices have fluctuated from $1.99 in November 1994, to over $5.00 in July 1996, to under $2.00 in 2000. Prices are currently at $2.50 per bushel. A reasonable budget value for corn is $3.00 per bushel. Based on the data in Table 8.1, the average corn production for the route is about 125 bushels/acre. Applying these guidelines provides a crop loss payment of approximately $1,300/acre.

The easement cost of $3,150/acre and the crop loss payment of $1,300/acre are combined with the acres per mile for the range of pipe diameters shown in Table 5.1. In addition, an experience factor is applied to capture the cost of the probable condemnation proceedings. The total right-of-way costs are shown for the range of diameters are shown in Table 8.2.

**Table 8.2. Pipeline right-of-way cost**

| Diameter (inches) | ($/mile) | ($/diameter inch/mile) | (million $/200 miles) |
|---|---|---|---|
| 4 | 36,713 | 9,178 | 7.3 |
| 6 | 36,713 | 6,119 | 7.3 |
| 8 | 44,500 | 5,563 | 8.9 |
| 10 | 44,500 | 4,450 | 8.9 |
| 12 | 51,731 | 4,311 | 10.3 |
| 16 | 66,750 | 4,172 | 13.4 |
| 18 | 66,750 | 3,708 | 13.4 |
| 20 | 66,750 | 3,338 | 13.4 |
| 22 | 66,750 | 3,034 | 13.4 |
| 24 | 66,750 | 2,781 | 13.4 |

### 8.3. Pipeline Material Cost

Pipe is the primary material cost for the project. Pipe cost can be estimated by determining the weight of the pipe needed, estimating cost per unit weight of the pipe. For the conceptual $CO_2$ project, the proposed wall thickness for the pipe diameter was presented in Section 3 under pipe wall thickness. In that section, the wall thicknesses were calculated to allow for increased wall thickness in areas of congestion or areas of special construction technique. The estimated cost of pipe per ton is assumed to be $950/t. Historically, this price would be considered to be high, since pipe sold for $700 to $750/t for an extended period of time. Over the past year, the price of pipe has rapidly escalated to $1,050 to $1,100/t. There are signs that this price will moderate over the next few years; however, it is unlikely that the cost will be below $900/t in the foreseeable future. The unit price for pipe is adjusted by a 15% factor to allow for $CO_2$ metallurgy requirements, taxes, and delivery. Because the pipeline requires a number of mainline block valves and needs a minimum of two flow-rate meters (inlet and outlet), a cost allowance was determined for each pipe diameter. These costs were allocated into the unit cost per mile and ranged from $5,750/mile for 4-inch pipe and $18,400/mile for 24-inch pipe. The pipe cost and block valve cost are combined and shown in Table 8.3.

Table 8.3. Pipeline material cost

| Diameter (inches) | ($/mile) | ($/diameter inch/mile) | (million $/ 200 miles) |
|---|---|---|---|
| 4 | 24,303 | 6,076 | 5 |
| 6 | 47,630 | 7,938 | 10 |
| 8 | 79,370 | 9,921 | 16 |
| 10 | 115,424 | 11,542 | 23 |
| 12 | 159,084 | 13,257 | 32 |
| 16 | 247,199 | 15,450 | 49 |
| 18 | 310,766 | 17,265 | 62 |
| 20 | 381,893 | 19,095 | 76 |
| 22 | 460,465 | 20, 930 | 92 |
| 24 | 546,136 | 22,756 | 109 |

## 8.4. Pipeline Construction Cost

The construction cost estimate for the pipeline is based on recent experience and discussions with union contractors who are familiar with the area. To obtain an approximate cost, several contractors were given a conceptual route map and were asked to provide an all-inclusive lay price. Discussions with the contractors indicated that it would be better to divide the unit pricing into three different types of construction: agricultural lay, congested lay, and special crossings for roads and rivers/streams. Unit costs were thus developed for these three types of construction. The vast majority of the route will be open agricultural lay. A rough estimate determined that only about 20,000 feet of congested lay and about 6,500 feet of special crossing techniques would be required. These variables were applied to each diameter. A 5% factor has been added to account for the cost associated with the non-destructive testing of the welds. The combined totals are shown in Table 8.4.

Table 8.4. Pipeline construction cost

| Diameter (inches) | ($/mile) | ($/diameter inch/mile) | (million $/ 200 miles) |
|---|---|---|---|
| 4 | 85,071 | 21,268 | 17 |
| 6 | 115,915 | 19,319 | 23 |
| 8 | 141,753 | 17,719 | 28 |
| 10 | 173,476 | 17,348 | 35 |
| 12 | 210,730 | 17,561 | 42 |
| 16 | 275,533 | 17,221 | 55 |
| 18 | 306,206 | 17,011 | 61 |
| 20 | 336,354 | 16,818 | 67 |
| 22 | 365,978 | 16,635 | 73 |
| 24 | 395,601 | 16,483 | 79 |

## 8.5. Pipeline Services Cost

The pipeline will require professional services including but not limited to engineering, survey, mapping, right-of-way acquisition, legal, permitting acquisition, environmental consulting, geotechnical analysis, vendor inspection, and construction inspection. Based on extensive experience in project management, it is recommended practice to base the estimate for these services on the subtotal of the cost for right-of-way, materials, and construction. For the smaller 4-inch line, the percentage required for professional services is approximately 20% and, for larger lines, is 12%. These costs are shown in Table 8.5.

## 8.6. Pipeline Total Cost

Combining the costs for right-of-way (Table 8.2), materials (Table 8.3), construction (Table 8.4), and services (Table 8.5) provides the total projected cost for the pipeline (Table 8.6). Table 8.7 is a cost summary that shows the cost per mile of right-of-way, material, construction, services, and total as they relate to the range of diameters.

**Table 8.5. Support services cost**

| Diameter (inches) | ($/mile) | ($/diameter inch/mile) | (million $/ 200 miles) | Factor (%) |
|---|---|---|---|---|
| 4 | 29, 217 | 7,304 | 6 | 20 |
| 6 | 38,049 | 6,341 | 8 | 19 |
| 8 | 47,812 | 5,977 | 10 | 18 |
| 10 | 56,678 | 5,668 | 11 | 17 |
| 12 | 67,447 | 5,621 | 13 | 16 |
| 16 | 88,422 | 5,526 | 18 | 15 |
| 18 | 95,721 | 5,318 | 19 | 14 |
| 20 | 102,050 | 5,102 | 20 | 13 |
| 22 | 107,183 | 4,872 | 21 | 12 |
| 24 | 121,018 | 5,042 | 24 | 12 |

**Table 8.6. Total cost**

| Diameter (inches) | ($/mile) | ($/diameter inch/mile) | (million $/ 200 miles) |
|---|---|---|---|
| 4 | 175,304 | 43,826 | 35 |
| 6 | 238,307 | 39,718 | 48 |
| 8 | 313,435 | 39,179 | 63 |
| 10 | 390,078 | 39,008 | 78 |
| 12 | 488,992 | 40,749 | 98 |
| 16 | 677,905 | 42,369 | 136 |
| 18 | 779,444 | 43,302 | 156 |
| 20 | 887,047 | 44,352 | 177 |
| 22 | 1,000,375 | 45,472 | 200 |
| 24 | 1,129,505 | 47,063 | 226 |

## Table 8.7. Summary cost

| Diameter (inches) | Right-of-way ($/mile) | Materials ($/mile) | Construction ($/mile) | Services ($/mile) | Total cost($/mile) |
|---|---|---|---|---|---|
| 4 | 36,713 | 24,303 | 85,071 | 29,217 | 175,304 |
| 6 | 36,713 | 47,630 | 115,915 | 38,049 | 238,307 |
| 8 | 44,500 | 79,370 | 141,753 | 47,812 | 313,435 |
| 10 | 44,500 | 115,424 | 173,476 | 56,678 | 390,078 |
| 12 | 51,731 | 159,084 | 210,730 | 67,447 | 488,992 |
| 16 | 66,750 | 247,199 | 275,533 | 88,422 | 677,905 |
| 18 | 66,750 | 310,766 | 306,206 | 95,721 | 779,444 |
| 20 | 66,750 | 381,893 | 336,354 | 102,050 | 887,047 |
| 22 | 66,750 | 460,465 | 365,978 | 107,183 | 1,000,375 |
| 24 | 66,750 | 546,136 | 395,601 | 121,018 | 1,129,505 |

### 8.7. Pump Station Cost

Under certain conditions, it may be preferable to use a smaller pipeline and install a pump station to increase capacity. In this case, the flow rate will require additional pressure boosting due to the increased pressure drop in the smaller pipeline. The cost savings of the smaller pipeline versus the larger pipeline (Table 8.6) will be partially offset by the cost of the pump station(s) and their additional power and operating costs.

The capital cost for a pump station is directly proportional to the installed BHP. Low-pressure stations for crude oil are around $800 to $900/BHP, whereas the stations used for $CO_2$ and refined products cost around $1,000/BHP. This capital cost per BHP has been confirmed on a current, multi-station project in Pennsylvania.

The additional pressure drop (due to friction) of the smaller pipeline determines the additional BHP required. This friction drop and BHP for the additional pump station(s) can be determined by using Figures 3.3 and 3.4. For example, if a flow rate of 450 million cubic feet per day is required through 200 miles, a 16-inch pipeline with a pump station (total cost of $134 million from Figure 3.4) can be used instead of a 20-inch pipeline (total cost $176 million from Figure 3.4). The required BHP of the pump station is required to determine the cost of the station(s) and the true capital savings of the 16-inch pipeline compared with the 20-inch pipeline. Referring to Figure 3.2, for a flow rate of 450 million cubic feet per day, the pressure drop per mile for a 16-inch and 20-inch pipeline would increase to 16 psi/mile from 5 psi/mile. The BHP of the station could be determined by using Figure 3.3 at 450 million cubic feet per day and 16-inch pipe diameter to be 75 BHP/mile.

The receipt facility would still deliver the same flow and pressure to the pipeline. As discussed previously in Section 3, it is reasonable to assume a receipt pressure of 2,500 psig and a minimum pressure of 1,500 psig, resulting in 1,000 psig differential. The friction drop in the 20-inch line will deliver the $CO_2$ the total distance without additional pumping. However, the 16-inch pipeline would consume the 1,000 psig of available pressure at a point 62.5 miles downstream from the receipt point. At that point, pump station(s) must provide the remaining pressure to reach the delivery point, which is an additional 137.5 miles. At 75 BHP/mile, required pumping capacity is approximately 10,000 BHP. At

$1,000/BHP, the pump stations will cost $10 million. The true cost of the 16-inch pipeline will be (from Table 8.6) $144 million versus $176 million for a 20-inch pipeline.

## 8.8. Operating Cost

Pipeline operating costs will be nominal compared with installation cost. An operating company recently prepared an estimate for a 140-mile, 6-inch LPG pipeline expansion that incorporates into an existing operation. Adjusted for escalation and new regulations, this estimate was $480,000/yr or about $1,000/ mile for a 6-inch pipeline. Normalizing this value to unit diameter gives $667/inch diameter/mile per year. Based on this information, operating costs have been summarized in Table 8.8 on a per-mile and a 200-mile basis for pipelines constructed of 4- to 24-inch diameter pipe.

**Table 8.8. Annual pipeline operating costs**

| Diameter (inches) | ($/mile) | ($/200 miles) |
|---|---|---|
| 4 | 2,667 | 533,333 |
| 6 | 4,000 | 800,000 |
| 8 | 5,333 | 1,066,667 |
| 10 | 6,667 | 1,333,333 |
| 12 | 8,000 | 1,600,000 |
| 16 | 10,667 | 2,133,333 |
| 18 | 12,000 | 2,400,000 |
| 20 | 13,333 | 2,666,667 |
| 22 | 14,667 | 2,933,333 |
| 24 | 16,000 | 3,200,000 |

The cost of operating a pump station includes the cost of the maintenance of the facilities and the cost of power to drive the facilities. The maintenance cost is a function of the cost of the initial installation, and the power cost is a function of the actual power consumed. Discussions with operating companies indicate that the yearly cost of personnel and maintenance for an electric station is approximately 5% of the installation cost. Because installation costs are about $1,000/BHP, yearly maintenance costs will be approximately $50/BHP. For the 200-mile, 16-inch pipeline with approximately 10,000 BHP, the yearly maintenance cost is projected to be approximately $500,000/year.

Assuming that the installed stations operate continuously 24 hours per day, 365 days per year, then the power cost is also a function of the cost of the BHP installed and the unit cost for power. A rule of thumb for electric power costs is that 1 BHP operating for 1 yr will consume $240/yr of electricity at a cost of $0.03/kWh. If the cost of electricity is assumed to be $0.045/kWh for a modern large load, then 1 BHP operating for 1 yr will cost $360/yr. Therefore, the electric power cost for the example 16-inch pipeline with 10,000-installed BHP would be $3.6 million/yr.

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# Appendix C1. Contruction Photo Gallery

**Figure C1.1. Pipe yard**

Pipe yards are stockpile locations near the pipeline route. Pipe yards are vital to the logistical support for a pipeline spread, and, ideally, all of the pipe is accumulated before construction begins. As a minimum, the yards must be supplied in a manner that supports construction.

The yards are located along the route at a spacing that is dependent on the diameter of the pipeline. The smaller the diameter is, the further apart the spacing needs to be. On a recent 120-mile, 16-inch pipeline, only two pipe yards were used. On an 880-mile, 36-inch pipeline, the yards were spaced 30 to 40 miles apart. Yards for smaller pipe diameter do not have to be adjacent to the route. Pipelines larger than 16 inches generally need to be close to the route to minimize hauling. If practical, the yard should be located near an existing railroad spur off a major roadway at a site developed for other industrial use.

**Figure C1.2. Spread construction yard**

To mobilize and operate a major spread requires a base of operation. The base is set up by the contractor to serve as a point for marshalling personnel and equipment, a location for delivery of miscellaneous materials, a site for fabrication and testing of welders, and a shop for equipment maintenance.

The yard also serves as the location for the construction managers, quality inspectors, surveyors, and environmental inspectors. The location of this yard is generally on the edge of a town or city near the route and is at the middle of the spread.

**Figure C1.3. Survey re-stake**

The pipeline company—through its engineering, construction management, and procurement company—will stake the pipeline centerline and right-of-way limits well in advance of construction. This preliminary staking permits the environmental inspectors to ensure that the route complies with the permit restrictions.

These re-stake crews proceed along the right-of-way supported by a right-of-way agent who is making advance contact with the tenants and landowners along the route. The right-of-way agents can also address any landowner issue of dispute. The crews use GPS survey equipment to navigate and stake the points based on georeferenced electronic drawings that were prepared in compliance with the permits and right-of-way agreements.

**Figure C1.4. Topsoil stockpiling progress**

The cleaning crew includes a small fence crew that constructs temporary gaps or gates where the pipeline crosses existing fences. These gates are kept closed at all times by crews following behind the fence crew.

The clearing crew removes obstructions from the right-of-way so that it is a suitable worksite for the work crews that follow. Generally, the area is completely cleared of timber, brush, and other vegetation. Timber, if marketable, is generally salvaged for the property owner. The remaining limbs, brush, and debris are burned or chipped and disposed of along the right-of-way in an approved manner. Bulldozers that have special blades are used to expedite the clearing process.

**Figure C1.5. Completed grading. Note that the skids stringing, pipeline centerline stakes, and topsoil are stockpiled at the edge of the right-of-way**

After clearing, the grading crew prepares the route for construction. In agricultural areas, topsoil separation is required in accordance with the easement and permits. In hilly terrain, preparations may require cutting down existing grades on high points and filling in low points.

The clearing and grading crews also install temporary bridges and culverts necessary to make the right-of-way useable as quickly as possible after rainfall events. They install ramps at roadways that allow equipment to approach and cross without impacting travel. These ramps provide access to the right-of-way from public roadways.

**Figure C1.6. Stringing from the pipe yard**

The contractor takes possession of the pipe at the pipe yard, which requires special trailers to distribute the pipe along the right-of-way. This work is generally subcontracted to a specialized hauling contractor who is familiar with the job and the government regulations. The pipe is handled with care to avoid damage to the coating and to the ends of the pipe.

**Figure C1.7. Setting pipe from the string truck on skids**

The pipe is placed along the right-of-way in a predetermined location and is positioned in a manner that facilitates the subsequent work processes. Spacing allowances are made for livestock movement or vehicles. Joints of special wall thickness or coating types must be placed in accordance with the engineering drawings.

Figure C1.8. Pipe bent to accommodate direction change

The pipe must conform to the horizontal and vertical changes in route direction by free stressing, bending, or the use of hot bends. For lines 8 inches and smaller, free stress of the pipe using its natural flexibility is practical. For larger pipelines, the pipe must be field bent to accommodate direction changes.

**Figure C1.9. Pipe lifted into bending machine mandrel**

After the optimum angle for a bend is determined, the total is divided into a series of smaller bends. The pipe is marked at the beginning and end points of the bends. The pipe is then inserted into the bending machine, and the series of smaller bends is completed. If field conditions do not permit field bends, hot bends made offsite are for bends of large deflection.

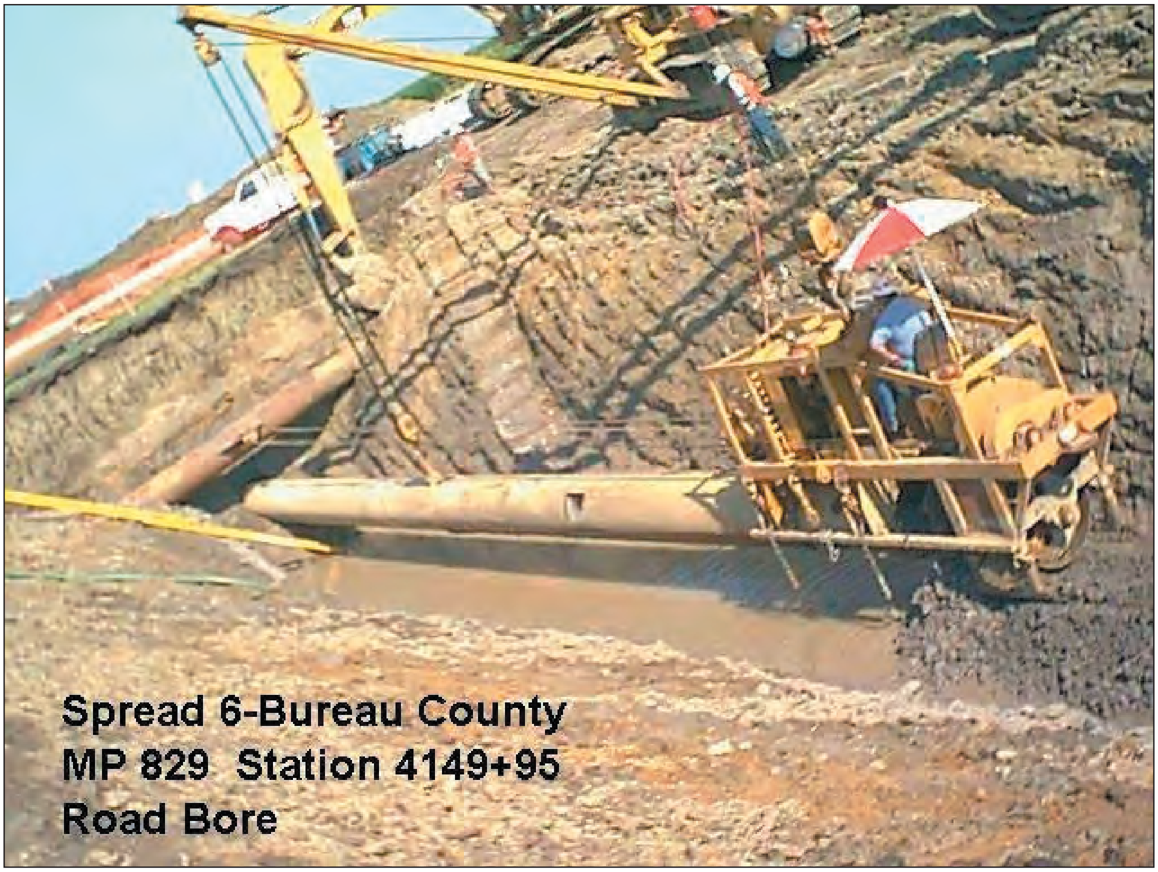

**Figure C1.10. Road boring operation**

At the start of the construction project, the road and stream crossing crews are generally in front of the main spread. During the project, they are sometimes caught and passed by the main spread. The road boring crews leapfrog along the route using public roadways. If the clear and grade crew passes them, they will often move down the right-of-way like a conventional crew.

Crossings of roads and streams generally require thicker pipe and extra depth. These are located by special permit drawings for each crossing. Roads are generally bored using a conventional boring machine that augers out a void under the roadway. Once boring is complete, a welded section of pipe is pulled into place and left to be tied into the mainline pipe to be installed later.

Streams or pipelines are usually crossed by open cutting the stream bed or excavating around the pipeline. Then a segment of pipe that has been built to the engineered profile of the crossing is installed. An example of this process is shown in the description of the tie-in crew.

Figure C1.11. Horizontal directional drill site

Horizontal directional drilling is generally done in the same manner as road crossing (drill and pull pre-welded pipe) but using an initial pilot hole that is directionally steered to match an engineered entry point, profile, and exit point. Small lines 8 inches and under can be economically installed for short distances and can even be used for lines over 5,000 ft.

This method of installation has geological limitations. It is not suitable for areas where soils are sandy and poorly graded, contain gravel or cobble, or are in highly faulted rock. The process is configured in a manner that allows the cuttings from the horizontal directional drilling operation to be returned to the drilling unit for handling and disposal.

**Figure C1.12. Horizontal directional drilling pilot coming out**

The horizontal directional drilling pilot is steered using a variable angle kicker tool similar to the techniques used in oil and gas directional drilling. The tool is steered using sensors at the tool and an electronic grid on entry and exit banks. The straightness of the pilot hole becomes critical for longer drills.

**Figure C1.13. Horizontal directional drilling barrel reamer coming out**

After the pilot hole has been properly established, a reamer is used to increase the size of the hole to accommodate the pipe.

**Figure C1.14. Horizontal directional drilling—ready to pull the prewelded pipe**

The pipe for the horizontal directional drilling is laid out, welded, inspected, and pre-hydrotested on the opposite bank from the HDD drill unit. After the hole has been completely reamed and debris removed using special tools, the pipe is ready to be pulled into the hole. The use of stabilizers and barrel reamers in front of the swivel is determined by the site conditions.

**Figure C1.15. Pulling operations under way**

The exit angle of the pilot hole determines the equipment considered necessary to assist in the pulling operation. The pipe needs to be lifted to the height where it “breaks over” and goes into the drilled hole at the proper angle to minimize resistance. The larger the exit angle is, the larger the equipment required to support the operation. The pipe is cradled on rollers to allow it to be pulled into the bore by the horizontal directional drilling unit.

**Figure C1.16. Pipe going in under levee canal**

The horizontal directional drilling unit is sized for the expected resistance during the pulling phase of the process. A wide range of unit sizes is available. Once the unit is pulled into place, a tie-in crew connects the pipe to the mainline. The pipe is then hydrotested with the rest of the pipeline.

**Figure C1.17. Wheel trencher**

The ditching operation excavates a stable trench so that the pipeline can be installed at the proper depth of cover. Prior to ditching, it is mandatory to call the appropriate agencies and notify the local utilities. It also is a common requirement for an electronic pipe locator to be used immediately in front of the ditching operation.

There is no one piece of equipment that will perform equally well under all conditions. The topography, geology, and weather on each project are unique. In general, a wheel trencher works well in stable soils because it is fast, makes a smooth trench, and stacks the spoil neatly in a pile. The hydraulic track backhoe is the alternate piece of equipment and is used when conditions are less than ideal or for smaller projects.

**Figure C1.18. Pipe is ready for pipe gang—the pipe is strung on skids, the ditch is complete, and the ditch spoil is segregated from topsoil**

Rock ditches can sometimes be accommodated by a wheel, saw trencher, or even backhoes. Some rock may require ripping by a bulldozer or blasting in advance of trenching. In an area of a rock ditch, the trench is padded with screened backfill before and after the pipe is lowered in. In all cases, the slope of the sides of the trench must be sufficient to prevent caving before the pipe can be welded and lowered.

In areas of high groundwater or for small pipelines, the ditch follows the welding. The trench is then dug to match the pipe. The pipe is lowered in the trench, and the trench is completely backfilled by the end of the day.

**Figure C1.19. End facing to ensure the highest weld quality**

The pipe is provided from the pipe mill with an end bevel for welding. For larger high-pressure lines that may have rigid welding requirements, end facing is sometimes required. End facing ensures that the bevels are correct and that the ends of the pipe are square for precise fit-up.

**Figure C1.20. Lineup clamp shoes for pipe joint**

Internal line clamps are not needed for small pipelines because their dimensional tolerances are within the tolerances allowed by the welding specifications. External lineup clamps are used to ensure proper welding set up for smaller pipelines. For larger lines (16 inches or greater), the internal line clamp is used as shown.

**Figure C1.21. Pipe joint being stabbed over internal lineup**

The pipe gang, working just ahead of the welding gang, sets the pipe in place so that welding can begin. The internal welding lineup clamp is moved forward to the end of the joint of pipe while welding is being completed. Once it is positioned on the end of the pipe being welded, the next joint is stabbed over the internal lineup clamp. That joint is positioned, spaced, and aligned with the previous joint, and the clamp is activated. The stringer pass (first pass) and some portions of the hot pass (second pass) are completed prior to release of the internal lineup clamp.

**Figure C1.22. Manual welding from both sides (12 o'clock to 6 o'clock)**

The firing line welders follow the pipe gang and complete the welding passes necessary for a complete weld. The number of passes depends on the thickness of the pipe. Generally, the welding operation is organized with a series of stations, each accomplishing a specific portion or pass of each weld. These stations start, finish, and move at the same time—like a wagon train.

Each weld is inspected visually, and a certain percentage of the welds are inspected by nondestructive testing. For manual welds, nondestructive testing is generally accomplished using external X-ray sources for small pipelines and internal X-ray machines for larger pipelines.

**Figure C1.23. Mechanized welding, large-diameter high-pressure pipeline**

Recent advances in the application of mechanized welding technology make it suitable for medium-diameter pipelines (20 to 24 inches) that operate at very high pressure and use high-grade (X80) steel pipe. Mechanized welding is best inspected by ultrasonic inspection because of its steep bevel angle.

**Figure C1.24. Coating using fusion bond epoxy**

There are many coating systems available to coat the weld joints, including conventional prime and tap, prime and heat shrink sleeves, fusion bond epoxy, and cold applied epoxy/urethane systems. Generally larger pipelines do not use tap or shrink sleeves. In all cases, cleaning the weld joint prior to application is critical.

**Figure C1.25. Jeeping for pipe coating defects**

When the pipe is welded and the joints are coated, the pipe is then ready to be lowered into the trench. To ensure that the coating has not been damaged, it is jeeped with a high-voltage coil. The voltage is set based on the coating thickness and properties so defects can be detected without damaging the coating. Just prior to lowering in, the pipe is lifted off the skids, and the coil is moved along the pipeline. Defects are marked and repaired.

**Figure C1.26. Pipe welded, coated, and ready to be lowered in**

The lower-in process is generally required to stay close to the welding process. In an area where wet subsoil conditions exist, it is common for the ditching and lower-in to be done on the same day so that no areas of the ditch are open at the end of the day.

**Figure C1.27. Lower-in tractors in various boom positions**

The lower-in operation uses side boom tractors, which are specially designed for this operation. The side booms have large counter weights on the side opposite the boom. These weights can be extended to give the tractors very stable lifting capacity. The number, size, and spacing of the side booms vary with pipe diameter, wall thickness, and soil conditions. The rollers that cradle the pipe are shown in a subsequent photo.

Figure C1.28. Lower-in crew making a road crossing

The lower-in crews, like the other crews, must take special precautions to prevent damage to roadways as they cross. Generally, traffic is stopped, and two rows of old tires are placed across the road. The side booms are guided to keep their tracks on the tires. Once the crew is across the roadway, the tires are removed, any debris is cleaned up, and traffic flow is restored. Note that the cradles on the end of the cables are roller types that allow the lower-in tractors to move in a continuous manner.

**Figure C1.29. Surveying weld location and cover**

As part of the as-built record, the locations of welds and pipe cover are recorded by survey or by inspection. An open ditch as-built is usually done for larger pipelines, and a closed ditch as-built is done for smaller pipelines.

**Figure C1.30. Lower-in backhoes (foreground) and dozer (background)**

After pipe has been installed in the trench, the backfill operation can commence.  Backfill is often required to be done on all pipe installed each day. For large-diameter pipe, backhoes generally plug the pipe by placing soil directly on the pipe to keep it stable for the subsequent backfilling operation. Then backhoes with a special wide buck pull the ditch spoil pile into the trench. A dozer follows along, providing further cleanup if necessary. On smaller pipelines, a dozer is often used to do rough backfill.

**Figure C1.31. Cleanup (aerial view)**

The cleanup operation is critical to comply with permit and landowner restrictions. Once the rough backfill operation has been completed, the cleanup operation can begin. Construction debris and surface rocks are removed to approved disposal locations. Subsoil is de-compacted as appropriate, and the surface is restored to the original profile. The topsoil is then returned to the original location.

This operation is very weather sensitive because the soil must be dry enough so that it can be worked without leaving ruts. Generally low ground pressure tractors are used for this operation.

Landowner releases are obtained, if possible, to provide documentation that the cleanup operation has been satisfactorily completed.

**Figure C1.32. Prefabricated mainline valve being installed in rock ditch**

Fabrication crews prepare and install valve sites, side taps, and scraper traps in the pipeline spread. The crews must ensure that foundations are at proper elevations, pipe is square, and that pipe is level. Generally, a tie-in crew will connect these to the mainline pipe.

**Figure C1.33. Optimal welding conditions are essential**

The pipe must be properly aligned and prepared so that optimal welding conditions are present, mitigating the need for repairs. Once the weld is made (as quickly as possible), it must be inspected, coated, and surveyed so that it can be backfilled.

**Figure C1.34. Tie-ins: Sag and overbend section set in place**

Tie-ins are made to connect areas of special construction to the mainline pipe. The tie-in crew is organized, with very experienced staff, and is equipped to handle multiple situations.

**Figure C1.35. Hydrotest pressures provide accurate measurement using dead weight and testers**

Once the pipeline has been completed for a minimum distance, that segment can be water filled and tested. The pipe segment is generally cleaned with air pigs or with a slug of water in front of a fill pig when the segment is being filled with hydrotest water. The fill is accomplished using a low-head, high-volume pump that takes water from an approved water source. After the segment is full of water, a high-pressure, low-volume pump is used to raise the pressure to the required test pressure. Pressure and temperature are then recorded. Pipelines are tested for a minimum time period (generally 8 hours) at a pressure that exceeds the maximum operating pressure by a specific safety factor (generally 25%).

**Figure C1.36. Dewatering pigs**

Dewatering and fill pigs are installed in the pipeline segments before the tests are started so that the water can be moved to the proper location at the end of the test.

**Figure C1.37. Bank of air compressors set up for dewatering of the test segment**

Air compressors are used to push test water to the next test segment or to a disposal point. Water for hydrotesting of a new pipeline can be discharged in most locations as long as it has been discharged through a hay bale structure.

**Figure C1.38. Drying pig deflector being installed after test/dewater header has been removed**

Once the test segment has been dewatered, several segments are connected into a larger segment for drying and cleaning. The air compressors used in dewatering are used to compress air through special dryers that remove the moisture from the air. This dry air then pushes a series of foam pigs and brush pigs through the cleaning segment. This process is repeated until the line is cleaned and the air being discharged has a dewpoint pressure less than in the specification.