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Gasification Combined Cycle: Carbon Dioxide Recovery, Transport, and Disposal

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Notation

The following is a list of the acronyms, initialisms, and abbreviations (including units of measure) used in this document. Some acronyms used only in tables are defined in those tables.

Acronyms, Initialisms, and Abbreviations

| | |
|-------|--|
| B&W | Babcock & Wilcox Co. |
| BFW | boiler feedwater |
| Cp/Cv | ratio of heat capacities |
| DOE | U.S. Department of Energy |
| H | enthalpy |
| HETP | height equivalent to a theoretical plate |
| HHV | higher heating value |
| HRSG | heat recovery steam generator |
| IGCC | integrated gasification combined cycle |
| K | equilibrium constant |
| k | reaction rate constant |
| KRW | Kellogg-Rust-Westinghouse |
| LSU | Louisiana State University |

Chemicals

| | |
|-------------------------------------|---------------------|
| Ar | argon |
| C | carbon |
| CaCO ₃ | calcium carbonate |
| CaMg(CO ₃) ₂ | dolomite limestone |
| CaO | calcium oxide |
| CaSO ₄ | calcium sulfate |
| CH ₄ | methane |
| CO | carbon monoxide |
| CO ₂ | carbon dioxide |
| COS | carbonyl sulfide |
| H ₂ | hydrogen |
| HCl | hydrogen chloride |
| H ₂ O | water |
| H ₂ S | hydrogen sulfide |
| K ₂ CO ₃ | potassium carbonate |
| MEA | monoethanolamine |
| N ₂ | nitrogen |
| NH ₃ | ammonia |
| NO _x | nitrogen oxides |

| | |
|-----------------|----------------|
| O ₂ | oxygen |
| SO ₂ | sulfur dioxide |
| SO _x | sulfur oxides |

Units of Measure

| | |
|------------------|--|
| atm | atmosphere(s) |
| Btu | British thermal unit(s) |
| °C | degree(s) Celsius |
| cm | centimeter(s) |
| cp | centipoise |
| d | day(s) |
| °F | degree(s) Fahrenheit |
| ft | foot (feet) |
| ft ² | square foot (feet) |
| ft ³ | cubic foot (feet) |
| g | gram(s) |
| gal | gallon(s) |
| Gt | gigatonne(s) |
| h | hour(s) |
| ha | hectare(s) |
| hp | horsepower |
| in. | inch(es) |
| in. ² | square inch(es) |
| J | joule(s) |
| K | degree(s) Kelvin |
| kg | kilogram(s) |
| km | kilometer(s) |
| kWh | kilowatt-hour(s) |
| kWhe | kilowatt-hour(s) (electric) |
| L | liter(s) |
| lb | pound(s) |
| m | meter(s) |
| mi | mile(s) |
| mi ² | square mile(s) |
| min | minute(s) |
| MJ | megajoule(s) |
| MMscfd | million standard cubic foot (feet) per day |
| mol | mole(s) |
| MPa | megapascal(s) |
| MW | megawatt(s) |
| MWh | megawatt-hour(s) |
| Nm ³ | normal cubic meter(s) |
| psi | pound(s) per square inch |
| psia | pound(s) per square inch absolute |

| | |
|------|------------------------------------|
| psig | pound(s) per square inch gauge |
| °R | degree(s) Rankine |
| s | second(s) |
| scf | standard cubic foot (feet) |
| scfd | standard cubic foot (feet) per day |
| t | tonne(s) (metric ton[s]) |
| W | watt(s) |
| yr | year(s) |

Gasification Combined Cycle: Carbon Dioxide Recovery, Transport, and Disposal

by

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Abstract

The objective of the project is to develop engineering evaluations of technologies for the capture, use, and disposal of carbon dioxide (CO₂). This project emphasizes CO₂-capture technologies combined with integrated gasification combined-cycle (IGCC) power systems. Complementary evaluations address CO₂ transportation, CO₂ use, and options for the long-term sequestering of unused CO₂. Commercially available CO₂-capture technology is providing a performance and economic baseline against which to compare innovative technologies. The intent is to provide the CO₂ budget, or an "equivalent CO₂" budget, associated with each of the individual energy-cycle steps, in addition to process design capital and operating costs. The value used for the "equivalent CO₂" budget is 1 kg of CO₂ per kilowatt-hour (electric). The base case is a 458-MW IGCC system that uses an air-blown Kellogg-Rust-Westinghouse agglomerating fluidized-bed gasifier, Illinois No. 6 bituminous coal feed, and in-bed sulfur removal. Mining, feed preparation, and conversion result in a net electric power production of 454 MW, with a CO₂ release rate of 0.835 kg/kWhe. For comparison, the gasifier output was taken through water-gas shift and then to amine, low-pressure glycol, chilled methanol, or hot potassium carbonate CO₂ recovery prior to the combustion turbine. Recovery of CO₂ was set at 90%, and the combustion turbine now was fed a fuel with high hydrogen content. From the IGCC plant, a 500-km pipeline took the CO₂ to geologic sequestering. For these cases, the net electric power production was reduced by 73.6-185.1 MW, with a CO₂ release rate of 0.29-0.53 kg/kWhe (when makeup power was considered). Life-cycle CO₂-sequestering costs ranged from \$113 to \$201/ton of CO₂. Two additional life-cycle energy balances for emerging technologies were considered: (1) high-temperature CO₂ separation with calcium- or magnesium-based sorbents, and (2) ambient-temperature facilitated-transport polymer membranes for acid-gas removal.

Summary

The possibility of global climate change resulting from increasing levels of greenhouse gases is the subject of considerable debate and uncertainty. Because of these concerns, the United Nations Framework Convention on Climate Change (in effect as of March 21, 1994) calls for the

United States and other industrialized nations that are signatories to submit action plans for stabilizing greenhouse gases by September 21, 1994. The most significant releases of carbon dioxide (CO_2) come from the commercial energy sector (Figure S.1); hence, options for greenhouse gas stabilization that are under consideration include strong energy-conservation measures, the capture and sequestering of CO_2 , and the substitution of nonfossil energy sources for fossil-fuel combustion.

Initiatives to limit CO_2 emissions have drawn considerable interest to integrated gasification combined-cycle (IGCC) power generation, a process that reduces CO_2 production through efficient fuel use, is amenable to CO_2 capture, and combines several desirable attributes. First, IGCC systems provide high energy-conversion efficiency, with the prospect of even higher efficiencies if higher-temperature turbines and hot-gas cleanup systems are developed. Second, very low emission levels for sulfur and nitrogen species have been demonstrated at such facilities as the Cool Water IGCC Plant in California. Third, IGCC plants produce gas streams with concentrated CO_2 and high levels of carbon monoxide (CO), which can be easily converted to CO_2 . Capture of this CO_2 prior to combustion requires the treatment of substantially smaller gas volumes than capture after combustion. As a consequence, the recovery of CO_2 in IGCC systems is potentially less expensive than in conventional combustion systems. In addition, CO_2 recovery may be accomplished in conjunction with hydrogen sulfide removal by using several commercially available technologies. Advanced process concepts now under development offer the prospect for further significant improvements.

This optimism must be tempered with the reality that the capture and disposal of CO_2 will have significant impacts on the efficiency and economics of the entire energy cycle. This report

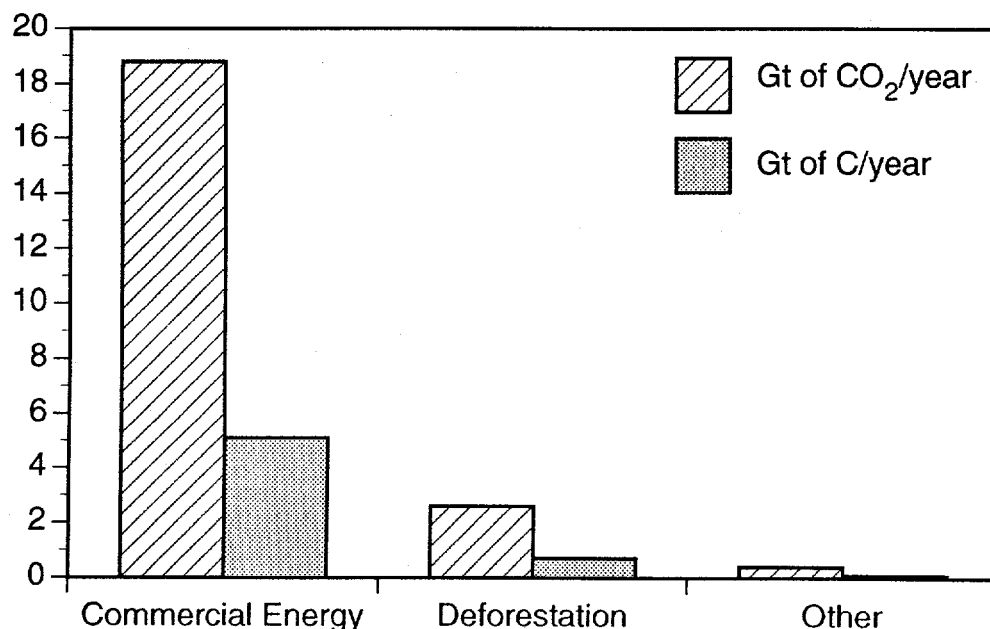


FIGURE S.1 Annual Atmospheric Release of CO_2 and Carbon

presents the results of a study of such energy-cycle impacts that is being conducted for the Morgantown Energy Technology Center, Morgantown, West Virginia. This study compares energy systems that encompass coal mining, preparation, and transport; an IGCC system; CO₂ recovery using commercially available technologies; CO₂ transport by pipeline; and land-based sequestering in geologic reservoirs. The intent is to evaluate the energy-efficiency impacts of recovering CO₂ and to provide the CO₂ budget, or an "equivalent CO₂" budget, associated with each of the individual energy-cycle steps. The CO₂ value used for the "equivalent CO₂" budget is 1 kg/kWh, which represents a typical value for current power generation.

The Argonne National Laboratory project also includes experimental work to investigate novel techniques for using CO₂, but that effort is outside the scope of this report. The research reported here has emphasized commercially available technologies for capturing CO₂, as well as ongoing work to evaluate advanced technologies under development and novel power-system configurations that may enhance system efficiency.

The base case for the comparison is a 494-MW (gross) IGCC system that uses an air-blown Kellogg-Rust-Westinghouse (KRW) agglomerating fluidized-bed gasifier, Illinois No. 6 bituminous coal, and in-bed sulfur removal. Mining, preparation, and transportation of the coal and limestone result in a net electric power production of 454 MW, with a CO₂ release rate of 0.835 kg/kWh. This figure by itself represents nearly a 17% decrease in the CO₂ release rate from the national average CO₂ release rate of 1 kg/kWh. For comparison, the output of the gasifier is taken through a water-gas shift to convert CO to CO₂, and then the shift product gas is separated to recover CO₂ prior to the combustion turbine. A 500-km pipeline then takes the dry 95%-CO₂ gas to geologic sequestering.

Carbon dioxide recovery systems considered in detail are amine CO₂ recovery, glycol solvent CO₂ recovery, chilled methanol CO₂ recovery, and hot potassium carbonate CO₂ recovery. Also reviewed, but not reduced to an expanded material-and-energy balance, are the calcium oxide CO₂ recovery process currently undergoing investigation at Louisiana State University (Section 9) and polymer membrane CO₂ recovery (Section 10).

S.1 Overview of Energy Cycle

The definition of the energy system for the study extends from the coal mine to the final geologic repository for the CO₂, as shown in Figure S.2 (Section 1). The location of the IGCC plant is specified as the midwestern United States. This report adopts a location 161 km by rail from Sesser, Illinois, where the coal is mined. Details of the IGCC portion of the system are taken from Gallaspy et al. (1990), who describe a plant that uses an air-blown KRW gasifier with in-bed sulfur removal. In each case studied, the CO₂ recovery technologies have been integrated into that plant design as much as possible to limit losses of efficiency. For each part of the energy system, CO₂ emissions have either been computed directly from process stream compositions or have been calculated from the energy consumption by using a "CO₂ equivalence" of 1 kg/kWh.

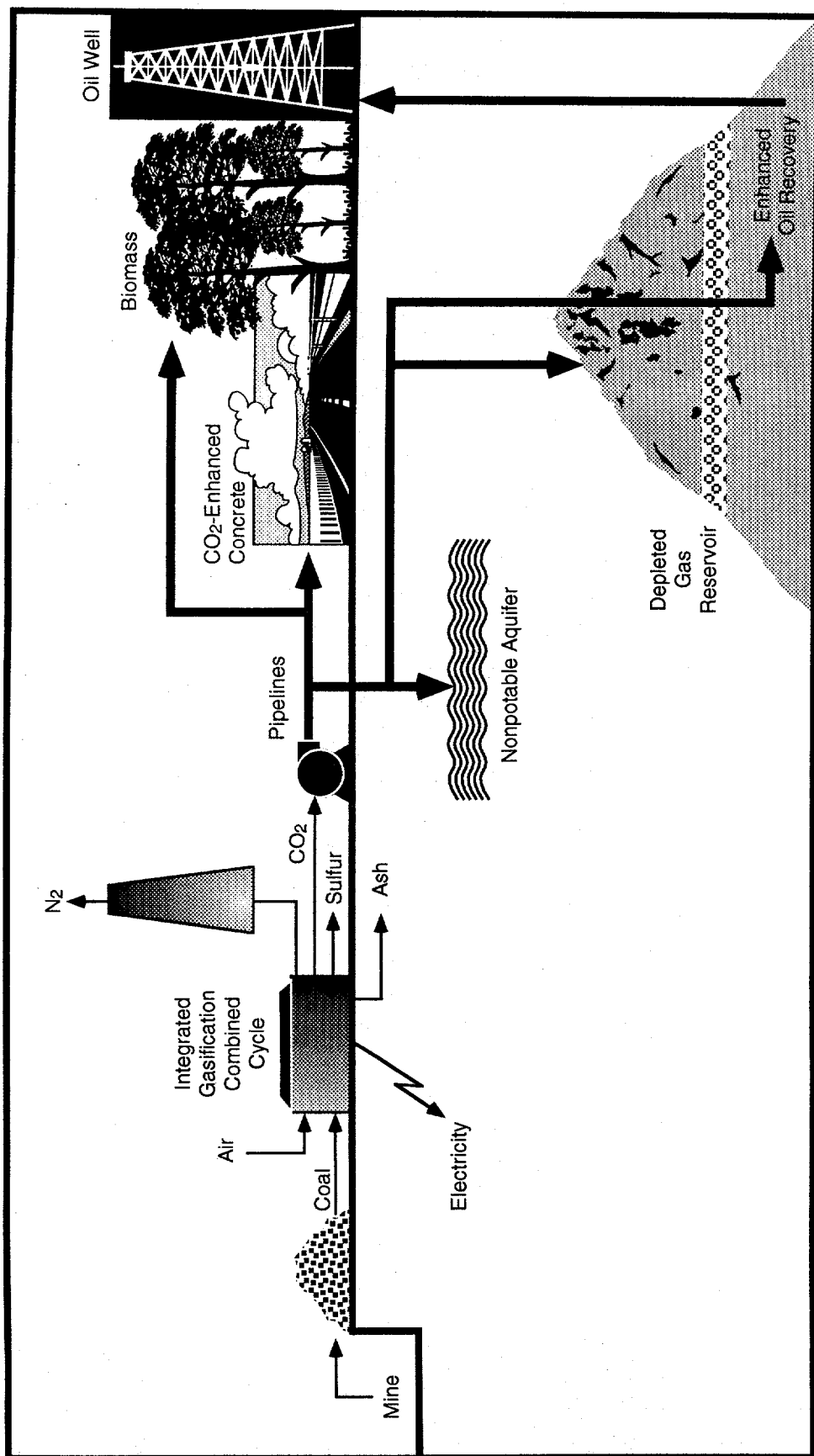


FIGURE S.2 Energy System Components for CO₂ Recovery

In this way, a total CO₂ budget for the system can be derived and compared with those for other options, thereby taking into account effects outside the immediate boundary of the plant.

S.2 Mining, Preparation, and Transportation of Raw Materials

Illinois No. 6 coal from the Old Ben No. 26 Mine is the fuel used in all of the cases studied (Section 2). The ultimate analysis appears in Table S.1. This mine is an underground mine with an associated coal-preparation plant. The power plant is assumed to be 161 km from the mine, with rail shipment of the coal by using a unit train. In order to supply coal at the 152,667 kg/h required by the IGCC plant, the mine and preparation-plant combination requires 2.42 MW of power and produces CO₂ at 2,955 kg/h. The rail transportation produces CO₂ at 637 kg/h.

TABLE S.1 Base Coal Analysis:
Illinois No. 6 (Old Ben No. 26 Mine)

| Variable | Data |
|--|--------|
| Component, as received (weight %) | |
| Moisture | 11.12 |
| Carbon | 63.75 |
| Hydrogen | 4.50 |
| Nitrogen | 1.25 |
| Chlorine | 0.29 |
| Sulfur | 2.51 |
| Ash | 9.70 |
| Oxygen (by difference) | 6.88 |
| Total | 100.00 |
| Ash fusion temperature (reducing conditions) (°C) | |
| Initial deformation | 1,201 |
| Softening (H=W) | 1,238 |
| Softening (H=1/2 W) | 1,285 |
| Fluid | 1,324 |
| Higher heating value, as received | |
| MJ/kg | 27.11 |
| Btu/lb | 11,666 |

Source: Gallaspy et al. (1990).

Limestone is used for in-bed sulfur capture in the gasifier. The assumption is that the limestone is extracted from a surface mine (quarry) about 161 km from the plant and is transported by rail to the plant site. Limestone is used at a rate of 39,795 kg/h by the process, giving an energy consumption for extraction of 0.257 MW, with an associated CO₂ production of 257 kg/h. Shipment of the limestone produces CO₂ at 160 kg/h.

S.3 Handling of Coal and Limestone

The coal-handling system at the plant includes equipment for unloading the coal from the unit train, passing the coal through magnetic separators, and then conveying it to 14-h storage silos (Section 3). The coal is crushed and dried in a series of three fluidized-bed roller mills. The heat for drying is provided by the hot (760°C) flue gas from the IGCC sulfator process. This drying results in a significant CO₂ emission from the energy cycle that is not reclaimed, but which presents a possible opportunity for further reductions. The coal is then held in a 2-h bunker, from which the material is pneumatically conveyed to surge bins ahead of the gasifier lock hoppers. The CO₂ emissions from the sulfator are 13,099 kg/h. Limestone is also prepared by crushing in two pulverizers and is then pneumatically conveyed to a 24-h storage silo and a 2-h storage bunker before being mixed with the coal in the gasifier surge bins. Energy consumption in this part of the plant is 6.18 MW.

S.4 Base Case for Integrated Gasification Combined Cycle

The base case for the comparisons uses air-blown KRW fluidized-bed gasifiers and in-bed sulfur removal (Section 4) (Gallaspay et al. 1990). A simplified schematic diagram for this process appears in Figure S.3. The system includes two heavy-duty industrial gas turbines (1,260°C firing temperature) coupled with a reheat steam-turbine bottoming cycle. Spent limestone and ash from the gasifier are oxidized in an external sulfator before disposal. The sulfator flue gas is taken to the coal-preparation operation for drying coal and is not integrated into the later CO₂ recovery operation.

The hot-gas cleanup system for particulate matter consists of a cyclone followed by a ceramic-candle-type filter. Solids collected are sent to the external sulfator before disposal. Gas temperatures are maintained at approximately 540°C. Supplemental hot-gas desulfurization is accomplished in a fixed-bed zinc ferrite system. Off-gas from the regeneration of this polishing step is recycled to the gasifier for in-bed sulfur capture.

The balance of the in-plant consumption of electricity is 36.6 MW, with a main-stack CO₂ emission of 362,820 kg/h. The gross IGCC electric power output is 493.8 MW, and the total in-plant consumption is subtracted to yield 454.4 MW of net power.

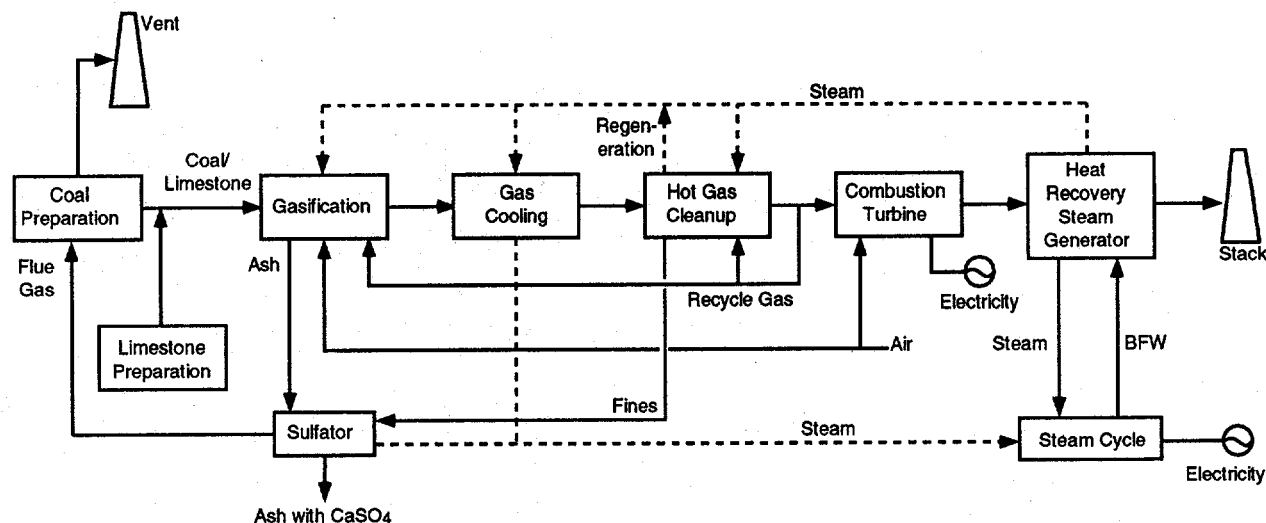


FIGURE S.3 Block Diagram of the Base-Case IGCC System (BFW, boiler feedwater)

S.5 Integrated Gasification Combined Cycle with CO₂ Recovery

Several changes were made to the base-case IGCC plant to incorporate CO₂ recovery (Figure S.4). These changes entailed processing the cleaned fuel gas through a "shift" reaction to convert the CO to CO₂, recovering the CO₂, and then combusting the low-CO₂ fuel gas in a modified turbine/steam cycle to produce electricity. Gas cleaning and sulfator performance were considered to be unaffected by these changes.

The fuel gas from the KRW process is high in CO. Conversion of the CO to CO₂ in the combustion process would result in substantial dilution of the resulting CO₂ with nitrogen from the combustion air and with water (H₂O) from the combustion reaction. If the CO₂ is removed prior to combustion, a substantial savings in the cost of the CO₂ recovery system is possible because of reduced vessel size and reduced flow rate of solvent. The CO in the fuel gas must first be converted to CO₂ by the shift reaction:



The resulting CO₂ can then be recovered, leaving a fuel rich in hydrogen (H₂) for use in the gas turbine.

The shift reaction is commonly accomplished in a catalyst-packed tubular reactor. A relatively low-cost iron oxide catalyst is effective in the temperature range of 340-590°C. Below that range, a more expensive copper oxide catalyst is required. While the equilibrium concentrations of products are favorable at lower temperatures, this consideration must be balanced against the need for larger reactors. In view of these conflicting considerations, high CO₂ recovery

is best achieved by staged reactors that allow for cooling between stages. For this study, a two-stage system was chosen, configured to achieve 95% conversion of CO to CO₂.

A number of CO₂ removal technologies are commercially available for application to IGCC systems; however, all of these options involve cooling or refrigerating the gas stream, with an attendant loss of thermal efficiency. To minimize the loss, the heat removed during cooling must be recovered and integrated into the system. Several options for this integration were evaluated, including steam generation alone, fuel gas preheating with supplemental steam generation, and fuel gas saturation and preheating. In the latter case, moisture condensed from the fuel gas prior to CO₂ recovery is injected into the clean fuel gas stream as it is heated by recovered heat following CO₂ removal. This process allows additional heat to be absorbed before combustion and increases the mass flow rate through the gas turbine. The balance of the thermal energy is used in the heat recovery steam generator for feedwater heating and steam generation.

The commercially available processes for CO₂ recovery operate by absorption of the CO₂ in a liquid solvent and subsequent regeneration of the solvent to release the CO₂. The temperature of absorption is solvent specific. In general, however, the solvents have low boiling points so that substantial cooling of the synthesis gas is required, as noted previously. Furthermore, lower temperatures favor absorption, thereby reducing the necessary solvent flow rate. This process implies a need for further cooling or refrigeration of the solvent, with additional energy losses. The regeneration of the solvent is also energy intensive for most processes because it is usually accomplished by flashing (pressure reduction) or heating (or both). If flashing is employed, repressurization of the solvent is required. Heating is generally accomplished by the extraction of steam from the steam cycle.

The alternative CO₂ recovery processes that have been evaluated to date include the following:

1. Absorption by monoethanolamine (MEA), with regeneration by throttling and heating via a reboiler (Section 5);
2. Two variants on the glycol CO₂ recovery system:
 - a. High-pressure (1,000-psi) absorption by a glycol-type solvent, with regeneration by flashing; and
 - b. Low-pressure (250-psi) absorption by a glycol-type solvent, with regeneration by flashing (Section 6);
3. Absorption by chilled methanol, with regeneration by thermal stripping and distillation (Section 7); and
4. Absorption by hot potassium carbonate (K₂CO₃), with regeneration by throttling and reboiler heating (Section 8).

From an energy perspective, the glycol options have the advantage, particularly for the low-pressure system, which does not require compression of the synthesis gas prior to absorption.

S.6 Pipeline Transport of CO₂

Once the CO₂ has been recovered from the fuel gas stream, its transportation and use or disposal remain as significant issues (Section 12). Carbon dioxide represents a large-volume, relatively low-value by-product that cannot be sequestered in the same way as most wastes from coal utilization (i.e., by landfilling). Large volumes of recovered CO₂ are likely to be moved by pipeline. In some cases, existing pipelines could be used, perhaps in a shared mode with other

products. In other cases, new pipeline construction would be required. Costs for pipeline construction and use vary greatly on a regional basis within the United States. The recovered CO₂ (363×10^3 kg/h) represents a gas volume of 4.02 million normal cubic meters (Nm³) per day (142×10^6 standard cubic feet per day [scfd]). The initial compression of the natural gas to pipeline pressures (for these studies, supercritical pressures of 2,100 psia were chosen) constitutes the major source of power consumption that must be borne by every pipeline. The receptor site for sequestering CO₂ from the IGCC plant is assumed to be 500 km from the plant site, requiring three booster stations. Table S.2 summarizes the power requirements for pipeline compression and transport.

S.7 Sequestering of CO₂

Proposals have been made to dispose of CO₂ in the ocean depths; however, many engineering and ecological concerns associated with such options remain unanswered, and the earliest likely reservoir is a land-based geologic repository (Section 12) (Hangebrauck et al. 1992). A portion of the CO₂ can be used for enhanced oil recovery, which sequesters a portion of the CO₂; or it can be completely sequestered in depleted gas/oil reservoirs and nonpotable aquifers. Both the availability of these zones and the technical and economic limits to their use need to be better characterized. Levelized costs were prepared with recognition that the power required for compression will rise throughout the life cycle of these sequestering reservoirs. The first reservoirs that would be used will, in fact, be capable of accepting all IGCC CO₂ gas for a 30-year period without requiring any additional compression costs for operation.

TABLE S.2 Power Requirements for Pipeline Compression and Transport^a

| Requirement | Power Requirement | | | |
|---------------------------------------|-------------------|---------------------|------------------|------------------------------------|
| | Amine | Low-Pressure Glycol | Chilled Methanol | Hot K ₂ CO ₃ |
| CO ₂ recovery (2,100 psia) | 38.63 | 33.04 | 21.96 | 38.38 |
| Pipeline booster stations | 2.07 | 2.07 | 2.07 | 2.07 |
| Total | 40.70 | 35.10 | 24.02 | 40.45 |

^a Basis, CO₂ at 322,000 kg/h (142 million standard cubic feet per day [MMscfd]).

S.8 Comparison of Power and CO₂ Emissions for IGCC/Shift/CO₂ Recovery Systems

The IGCC base case without CO₂ controls yields the energy consumption and CO₂ emissions appearing in Section 13. The IGCC plant delivers 493.8 MW of power (gross). The fuel-cycle power use results in a net power of 454.4 MW at a CO₂ emission rate of 0.835 kg/kWh. This rate is significantly below the typical CO₂ emission rate of 1 kg/kWh for the installed electric power generating capacity used throughout the study. Table S.3 enumerates the power consumption and associated CO₂ emissions throughout the entire power cycle: mining and transport, IGCC plant with and without CO₂ recovery and compression, pipeline booster stations, and sequestering. The values for the power consumption for compression and the values for the associated CO₂ emissions are segregated differently from those in Table S.2 to distinguish between CO₂ emission rates for power used within the IGCC facility and typical emission rates for power purchases from the grid for the pipeline booster stations. Subsequently, the values for gas-turbine and steam-turbine power generation are summed to yield the gross power. The difference between the gross power and the power consumption for the entire power cycle yields the value described as the "net cycle energy."

Lastly, the atmospheric emissions of CO₂ and the change from the KRW base case are considered for the scenario in which no replacement power is used and for the scenario in which replacement power must be purchased from the existing power grid (with its attendant high CO₂ emissions). For each of these scenarios, the CO₂ emission rate per unit of net cycle energy power and the rate of power consumption for sequestering CO₂ are calculated. The data from Table S.3 are summarized graphically in Figure S.5.

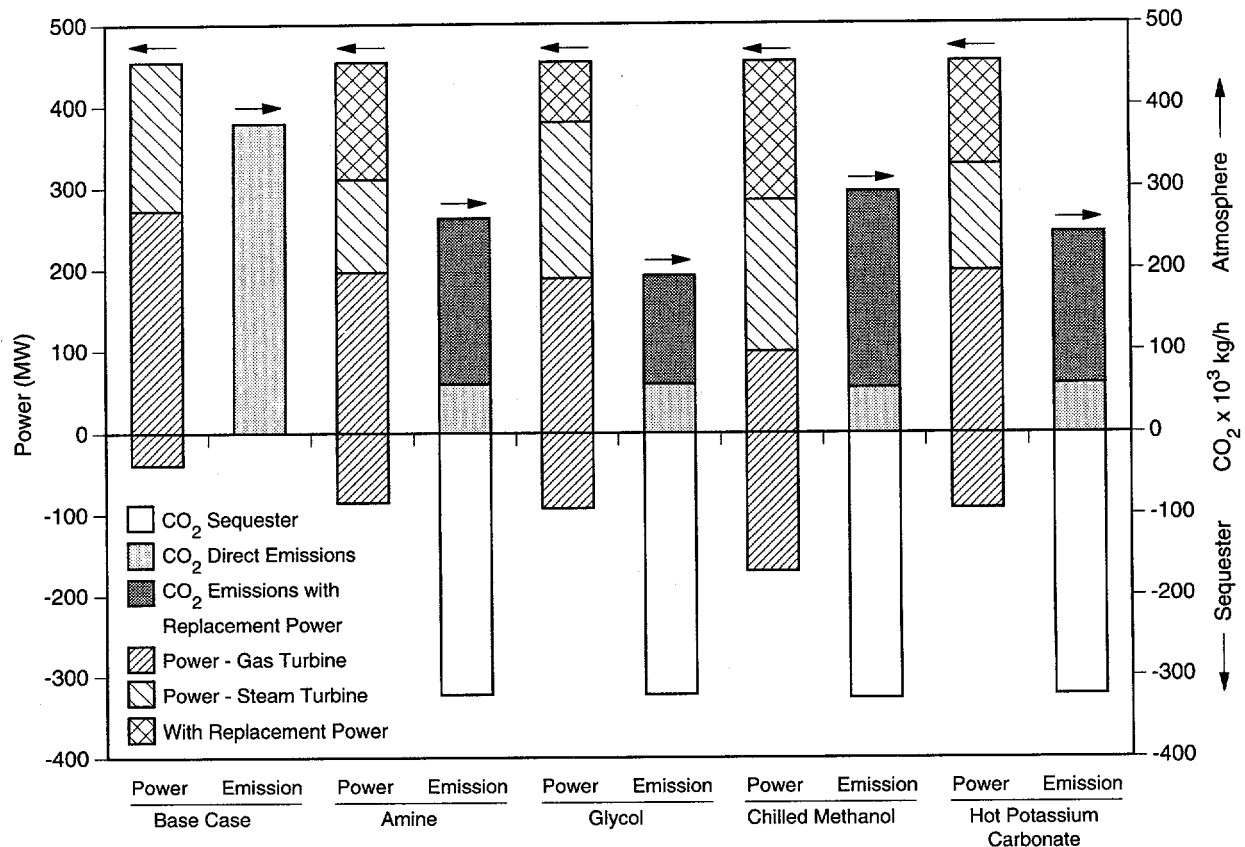
The number of greatest significance is the value for the rate of power consumption for sequestering CO₂. On the basis of this value, the ranking of the CO₂ recovery systems is as follows: low-pressure glycol > hot K₂CO₃ > amine > chilled methanol.

For the low-pressure glycol CO₂ recovery cases (the most favorable cases), the system efficiency is reduced from the base-case value of 39.6% to 34.5%, with a reduction of about 74 MW in the net system output. If the energy used for compression of the CO₂ at the IGCC plant is combined with the booster station requirements, nearly one-half of that reduction is seen to be caused by transportation of the CO₂ to the disposal site. Nevertheless, the total CO₂ emissions are reduced to only 16% of those in the base case, and the emission rate is reduced from 0.835 kg/kWh to only 0.156 kg/kWh.

TABLE S.3 Comparison of Power and CO₂ Emissions for IGCC/Shift/CO₂ Recovery Systems

| Factor | Base Case | | | Amine | | | Low-Pressure Glycol | | | Chilled Methanol | | | Hot K ₂ CO ₃ | | |
|---|------------|----------------------------------|--|------------|----------------------------------|--|---------------------|----------------------------------|--|------------------|----------------------------------|--|------------------------------------|----------------------------------|--|
| | Power (MW) | CO ₂ Emissions (kg/h) | | Power (MW) | CO ₂ Emissions (kg/h) | | Power (MW) | CO ₂ Emissions (kg/h) | | Power (MW) | CO ₂ Emissions (kg/h) | | Power (MW) | CO ₂ Emissions (kg/h) | |
| Mining and transport | 2.80 | 3,452 | | 2.80 | 3,452 | | 2.80 | 3,452 | | 2.80 | 3,452 | | 2.80 | 3,452 | |
| IGCC plant/compression | 36.63 | 376,077 | | 80.13 | 47,158 | | 87.35 | 47,267 | | 164.98 | 42,615 | | 87.19 | 47,158 | |
| Pipeline/injection | 0.00 | 0 | | 2.07 | 2,066 | | 2.07 | 2,066 | | 2.07 | 2,066 | | 2.07 | 2,066 | |
| Sequestering | 0.00 | 0 | | 0.00 | -322,341 | | 0.00 | -322,234 | | 0.00 | -326,793 | | 0.00 | -322,341 | |
| Total (without sequestering) | 39.43 | 379,529 | | 85.00 | 59,259 | | 92.22 | 59,361 | | 169.88 | 54,836 | | 92.06 | 59,259 | |
| Power from gas turbine | 311.60 | ^a | | 282.10 | - | | 281.60 | - | | 269.00 | - | | 290.00 | - | |
| Power from steam turbine | 182.20 | - | | 113.40 | - | | 191.40 | - | | 170.10 | - | | 130.30 | - | |
| Gross power | 493.80 | - | | 395.50 | - | | 473.00 | - | | 439.10 | - | | 420.30 | - | |
| Net cycle energy | 454.37 | - | | 310.50 | - | | 380.78 | - | | 269.22 | - | | 328.24 | - | |
| No replacement power | | | | | | | | | | | | | | | |
| Atmospheric emissions of CO ₂ | - | 379,529 | | - | 59,259 | | - | 59,361 | | - | 54,836 | | - | 59,259 | |
| Change from base KRW | 0.00 | 0 | | -143.87 | 320,270 | | -73.59 | 320,168 | | -185.15 | 324,693 | | -126.13 | 320,270 | |
| CO ₂ emission (kg/kWh) | 0.84 | - | | 0.19 | - | | 0.16 | - | | 0.20 | - | | 0.18 | - | |
| CO ₂ sequestering (kWh/kg of CO ₂) | None | - | | 0.45 | - | | 0.23 | - | | 0.57 | - | | 0.39 | - | |
| Replacement power | | | | | | | | | | | | | | | |
| Atmospheric emissions of CO ₂ | - | - | | 143.87 | 143,871 | | 73.59 | 73,590 | | 185.15 | 185,149 | | 126.13 | 126,131 | |
| Change from base KRW | - | - | | - | 203,130 | | - | 132,951 | | - | 239,985 | | - | 185,390 | |
| CO ₂ emission (kg/kWh) | 0.00 | 0 | | 0.00 | -176,399 | | 0.00 | -246,578 | | 0.00 | -139,544 | | 0.00 | -194,139 | |
| CO ₂ sequestering (kWh/kg of CO ₂) | 0.84 | - | | 0.45 | - | | 0.29 | - | | 0.53 | - | | 0.41 | - | |
| | None | - | | 0.82 | - | | 0.30 | - | | 1.33 | - | | 0.65 | - | |

^a Not applicable.

FIGURE S.5 Net Power Generation and Associated CO₂ Inventory

S.9 Economics of CO₂ Recovery, Transport, and Sequestering

S.9.1 Direct Costs for IGCC/Shift/CO₂ Recovery

Summaries of the economics for the IGCC/CO₂ recovery systems appear in Section 13, while Table S.4 summarizes the economics for the KRW base case and for the four shift/CO₂ recovery systems surveyed. A levelized fixed charge of 20% is taken for the capital on the CO₂ recovery systems; and in a similar way, labor and maintenance have been included as fixed percentages of the capital. While the equipment installed in the base plant has not changed, the cost in terms of dollars per kilowatt-hour escalates because of the derating of the plant. The annual costs of operating the plant in terms of dollars per year use a fourth-quarter 1993 base. A somewhat higher levelized fixed charge of 23% was considered appropriate for the pipeline. To this levelized fixed charge, the annual costs of consumables and chemicals are added. Costs for power are taken consistently at 60 mills/kWh during the operation of the plant. The on-stream factor for the IGCC/CO₂ recovery system was taken to be 65%, as in the base case. Hence, no derating of plant availability is assumed as a consequence of the addition of shift and CO₂ recovery. Table S.5 summarizes these costs on the basis of CO₂ sequestered.

TABLE S.4 Comparative Costs of IGCC with Shift/CO₂ Recovery and Pipeline

| Component | Unit | KRW Base | MEA | Glycol | Methanol | K ₂ CO ₃ |
|---------------------------------------|-----------------------|---------------|---------------|---------------|---------------|--------------------------------|
| Base Plant Capital | \$/kW | 1342 | 1945 | 1591 | 2238 | 1842 |
| CO ₂ Control Capital | \$/kW | 0 | 647 | 529 | 861 | 615 |
| Power Plant Annual Cost | \$/yr | \$128,157,363 | \$148,835,848 | \$148,325,526 | \$160,519,522 | \$148,441,638 |
| Base Plant Power Cost | mills/kWh | 49.5 | 57.5 | 57.3 | 62.0 | 57.4 |
| Pipeline (500 km) | mills/kWh | 0.0 | 19.1 | 19.1 | 19.1 | 19.1 |
| Replacement Power | mills/kWh | 0.0 | 19.0 | 9.7 | 24.4 | 16.7 |
| Net Power Cost | mills/kWh | 49.5 | 95.6 | 86.1 | 105.6 | 93.1 |
| Coal Energy Input | 10 ⁶ Btu/h | 3940 | 3940 | 3940 | 3940 | 3940 |
| Busbar Power Output | MW | 493.80 | 395.50 | 473.00 | 439.10 | 420.30 |
| In Plant Power Use | MW | 36.63 | 80.13 | 87.35 | 164.98 | 87.19 |
| Net Plant Output | MW | 457.17 | 315.37 | 385.65 | 274.12 | 333.11 |
| Net Heat Rate | Btu/kWh | 8618 | 12493 | 10217 | 14373 | 11828 |
| Thermal Efficiency - HHV ^b | % | 39.62 | 27.33 | 33.42 | 23.76 | 28.87 |
| Out of Plant Power Use | MW | 2.80 | 4.87 | 4.87 | 4.90 | 4.87 |
| Net Energy Cycle Power | MW | 454.37 | 310.50 | 380.78 | 269.22 | 328.24 |
| Net Energy Cycle Heat Rate | Btu/kWh | 8671 | 12689 | 10347 | 14635 | 12003 |
| Thermal Efficiency - HHV | % | 39.38 | 26.91 | 33.00 | 23.33 | 28.45 |
| Net Energy Cycle Power | MW | 454.37 | 310.50 | 380.78 | 269.22 | 328.24 |
| Replacement Power | MW | 0.00 | 143.87 | 73.59 | 185.15 | 126.13 |
| Net Grid Power | MW | 454.37 | 454.37 | 454.37 | 454.37 | 454.37 |

^aAll power costs on the basis of net grid power.^bHHV, Higher heating value.

S.9.2 Direct Costs for Pipeline

Pipeline costs are identical for all of the CO₂ recovery systems. While these costs of \$28/t of CO₂ are quite defensible for a dedicated pipeline, this cost element has the most significant opportunity for optimization. An argument could be made for reducing this cost to \$10/t of CO₂ under a different set of design assumptions. The use of a dedicated pipeline for this single facility requires that the pipe be sized for the maximum hourly flow rate. This sizing means that the pipe is used 65% of the time, the same as the on-stream factor for the IGCC/shift/CO₂ recovery system. By considering the converse, the 35% off-stream time represents a significant economic penalty. If the same economic factors were to drive the transport of CO₂ that operate in the natural gas distribution network, the pipeline cost of \$1.57/standard cubic foot (scf) of CO₂ could come into line with the distribution costs for a natural gas pipeline of approximately \$0.50/scf of CO₂.

S.9.3 Costs of Replacement Electricity

The costs of replacement electricity have been included and charged at a rate of 60 mills/kWh. This replacement cost is taken as the fuel-cycle derating of the IGCC/CO₂ recovery system. This derating is simply the difference between the base-case fuel-cycle electric production

TABLE S.5 Economics of CO₂ Recovery, Transport, and Sequestering:
Cost Summary on Annual Basis

| Cost Factor | Amine | Low-Pressure Glycol | Chilled Methanol | Hot K ₂ CO ₃ |
|--|-------|---------------------|------------------|------------------------------------|
| Shift/CO ₂ recovery (10 ⁶ \$) | 34.3 | 36.2 | 86.2 | 47.6 |
| Cost (\$/t of CO ₂) | 19 | 20 | 47 | 26 |
| Cost (\$/1,000 scf of CO ₂) | 1.05 | 1.11 | 2.63 | 1.45 |
| Cost (\$/t of C) | 69 | 73 | 173 | 96 |
| Pipeline (500 km) (10 ⁶ \$) | 51.4 | 51.4 | 51.4 | 51.4 |
| Cost (\$/t of CO ₂) | 28 | 28 | 28 | 28 |
| Cost (\$/1,000 scf of CO ₂) | 1.57 | 1.57 | 1.57 | 1.57 |
| Cost (\$/t of C) | 103 | 103 | 103 | 103 |
| Replacement power at 60 mills/kWh (10 ⁶ \$) | 49.2 | 25.1 | 63.3 | 43.1 |
| Cost (\$/t of CO ₂) | 27 | 14 | 35 | 24 |
| Cost (\$/1,000 scf of CO ₂) | 1.50 | 0.77 | 1.93 | 1.31 |
| Cost (\$/t of C) | 99 | 51 | 127 | 87 |
| Total cost (10 ⁶ \$) | 134.8 | 112.8 | 200.9 | 142.1 |
| Total cost (\$/t of CO ₂) | 73 | 61 | 108 | 77 |
| Cost (\$/1,000 scf of CO ₂) | 4.11 | 3.44 | 6.13 | 4.33 |
| Total cost (\$/t of C) | 269 | 225 | 396 | 284 |

of 454.4 MW and the fuel-cycle costs for each process. Note that the power consumption already counted against the shift/CO₂ recovery system and also the pipeline power use appear here. When replacement costs for electricity are considered, the chilled methanol system, with a cost of \$35/t of CO₂, is clearly at a disadvantage in comparison with the other technologies. The low-pressure glycol system shows a distinct advantage, with a replacement power cost of \$14/t of CO₂.

S.9.4 Summation of Capital, Operating, and Replacement-Electricity Costs

The summation of the capital, operating, and replacement-electricity costs for the IGCC/shift/CO₂ recovery system and the pipeline costs ranges from \$65 to \$108/t of CO₂. The calculated base-case power cost of 49.5 mills/kWh is increased by nearly 74% in the most favorable case to 86.1 mills/kWh (shift with low-pressure glycol recovery). The ranking of systems from most favorable total cost to highest cost is as follows: glycol > amine > hot K₂CO₃ > chilled methanol. The low-pressure glycol system might actually appear to be at a disadvantage with respect to the amine CO₂ recovery until the impacts of power generation derating and the purchase of replacement power are taken into account. The amine and hot K₂CO₃ systems should be considered to be too close to distinguish for a process design review at this level of detail. Were

it to be included, the membrane separation technology would have a cost of nearly twice that of the amine recovery, making membrane separation technology the highest cost option.

S.10 References for Summary

Gallaspy, D.T., et al., 1990, *Assessment of Coal Gasification/Hot Gas Cleanup Based Advanced Gas Turbine Systems: Final Report*, DOE/MC/26019.3004 (DE91002084), prepared by Southern Company Services, Inc., Birmingham, Ala., et al., for the U.S. Department of Energy, Morgantown Energy Technology Center, Morgantown, W. Va., Dec.

Hangebrauck, R.P., et al., 1992, "Carbon Dioxide Sequestration," presented at the 1992 Greenhouse Gas Emissions and Mitigation Research Symposium, sponsored by the U.S. Environmental Protection Agency, Washington, D.C., Aug. 18-20.

1 Introduction

1.1 Background

The possibility of a change in global climate resulting from increasing levels of "greenhouse" gases is the subject of considerable debate and uncertainty; however, increasing atmospheric concentrations of carbon dioxide (CO_2) have the potential for significant impacts that may not be easily reversed. Because of these concerns, policies to limit CO_2 emissions are being discussed, both in the United States and in various international forums (Princiotta 1992). The options under consideration include calls for strong energy-conservation measures, increasing the efficiency of fossil-fuel energy conversion systems, the capture and sequestering of CO_2 , and the substitution of nonfossil energy sources for fossil-fuel combustion. Discussion of the issues has drawn considerable interest to power generating systems that minimize the production of CO_2 and are amenable to CO_2 capture.

Integrated gasification combined-cycle (IGCC) systems are becoming an increasingly attractive option among the emerging technologies because such systems combine several desired attributes. First, IGCC systems provide high energy-conversion efficiency, with the prospect of even higher efficiencies if higher-temperature turbines and hot-gas cleanup systems are developed. Second, very low emission levels for sulfur and nitrogen oxide species (SO_x and NO_x) have been demonstrated at such facilities as the Cool Water IGCC Plant in California. Third, IGCC plants produce flue-gas streams with concentrated CO_2 and high levels of carbon monoxide (CO), which can be easily converted to CO_2 . The base case chosen is a 458-MW IGCC system that uses an air-blown Kellogg-Rust-Westinghouse (KRW) gasifier with in-bed sulfur removal. Detailed consideration of the design and costs for IGCC systems appeared in a recent report from the Morgantown Energy Technology Center (Gallaspy et al. 1990b).

Recovery of CO_2 in IGCC systems appears to be less expensive than in conventional combustion systems. Recovery of CO_2 now can be accomplished in conjunction with removal of hydrogen sulfide (H_2S) by using several commercially available technologies. At the same time, advanced process concepts now under development offer the prospect for further significant improvements.

1.2 Goals, Objectives, and Approach

The objective of the project is to develop engineering evaluations of technologies for the capture, use, and disposal of CO_2 combined with IGCC power systems. Complementary evaluations will address CO_2 transportation, CO_2 use, and options for the long-term sequestering of unused CO_2 . Commercially available CO_2 -capture technology will provide performance and economic baselines for comparing innovative technologies (Figure 1.1).

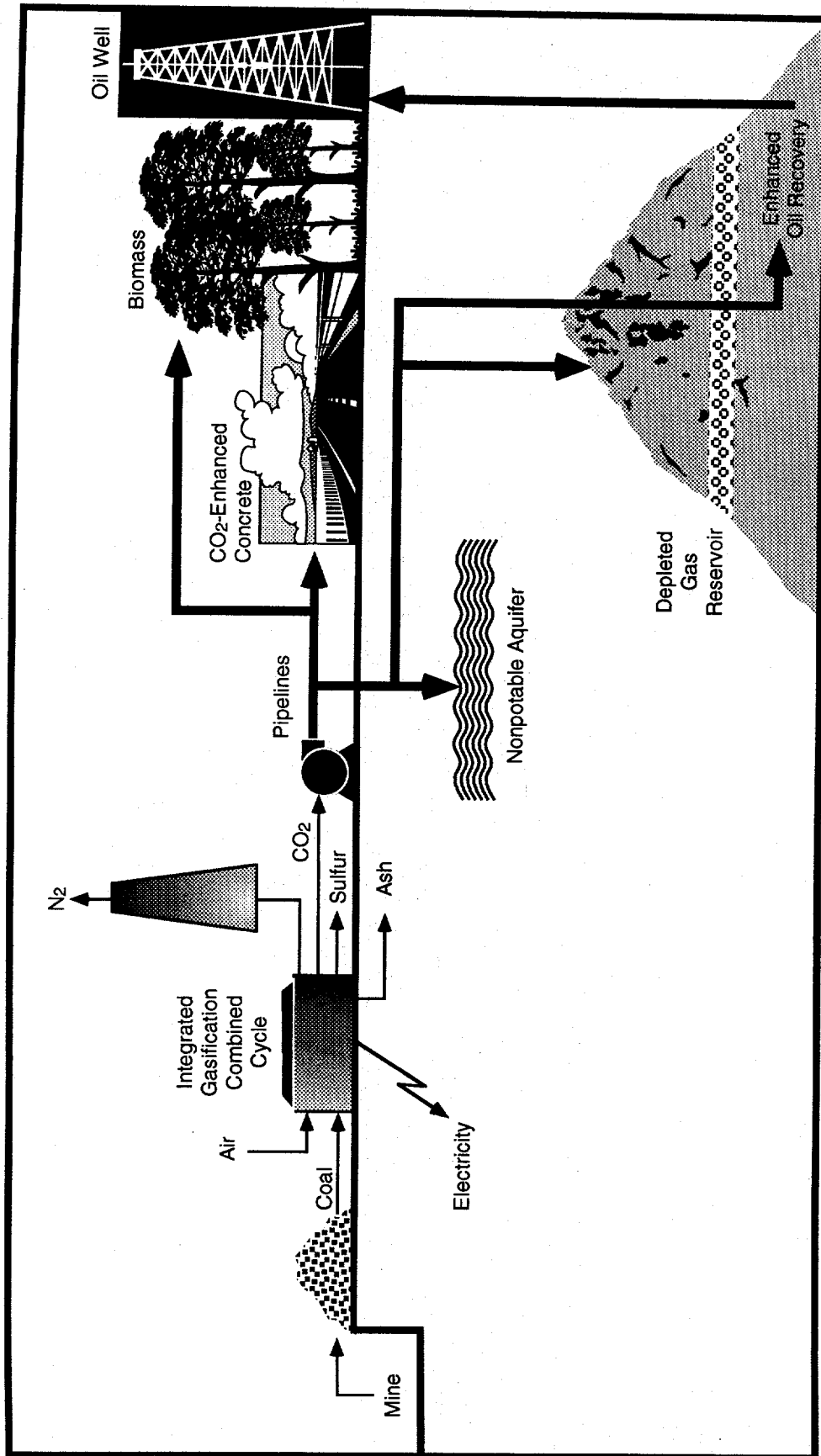


FIGURE 1.1 Scope of Project: To Estimate CO₂ Emissions from Mining through Sequestering

The intent is to provide the CO₂ budget, or an "equivalent CO₂" budget, associated with each of the individual energy-cycle steps, in addition to process design capital and operating costs. The value used for the "equivalent CO₂" budget is 1 kg of CO₂ per kilowatt-hour (electric). The base case is a 458-MW IGCC system that uses an air-blown KRW agglomerating fluidized-bed gasifier, Illinois No. 6 bituminous coal feed, and in-bed sulfur removal. Three commercial CO₂ recovery technologies are under study: (1) amine scrubbing, (2) a glycol solvent (Selexol), and (3) chilled methanol (Rectisol). Three emerging CO₂ recovery technologies are also being evaluated: (1) high-temperature CO₂ separation with calcium- or magnesium-based sorbents, (2) high-temperature molten-carbonate membranes for acid-gas removal, and (3) ambient-temperature facilitated-transport membranes for acid-gas separation.

2 Mining

2.1 Extraction

So as to keep in perspective the significance of coal mining and transportation on the fuel-cycle CO_2 emissions, the emission rates typical for the United States are shown in Figure 2.1 (Marlund 1983; Ashton et al. 1990). These emission rates appear in Table 2.1; and the low extraction energy penalties, as compared with other fossil fuels, reflect both the abundance of coal and the ease of recovering it.

The IGCC base case employs a prepared feed of coal to the gasifier of 3,792 tons/d (143,335 kg/h); however, the overall extraction process incorporates energy use, CO_2 emission, and the generation of waste products. To begin, the coal losses as waste that will be incurred in this cycle appear in Table 2.2. The detailed analysis of these individual coal-handling steps

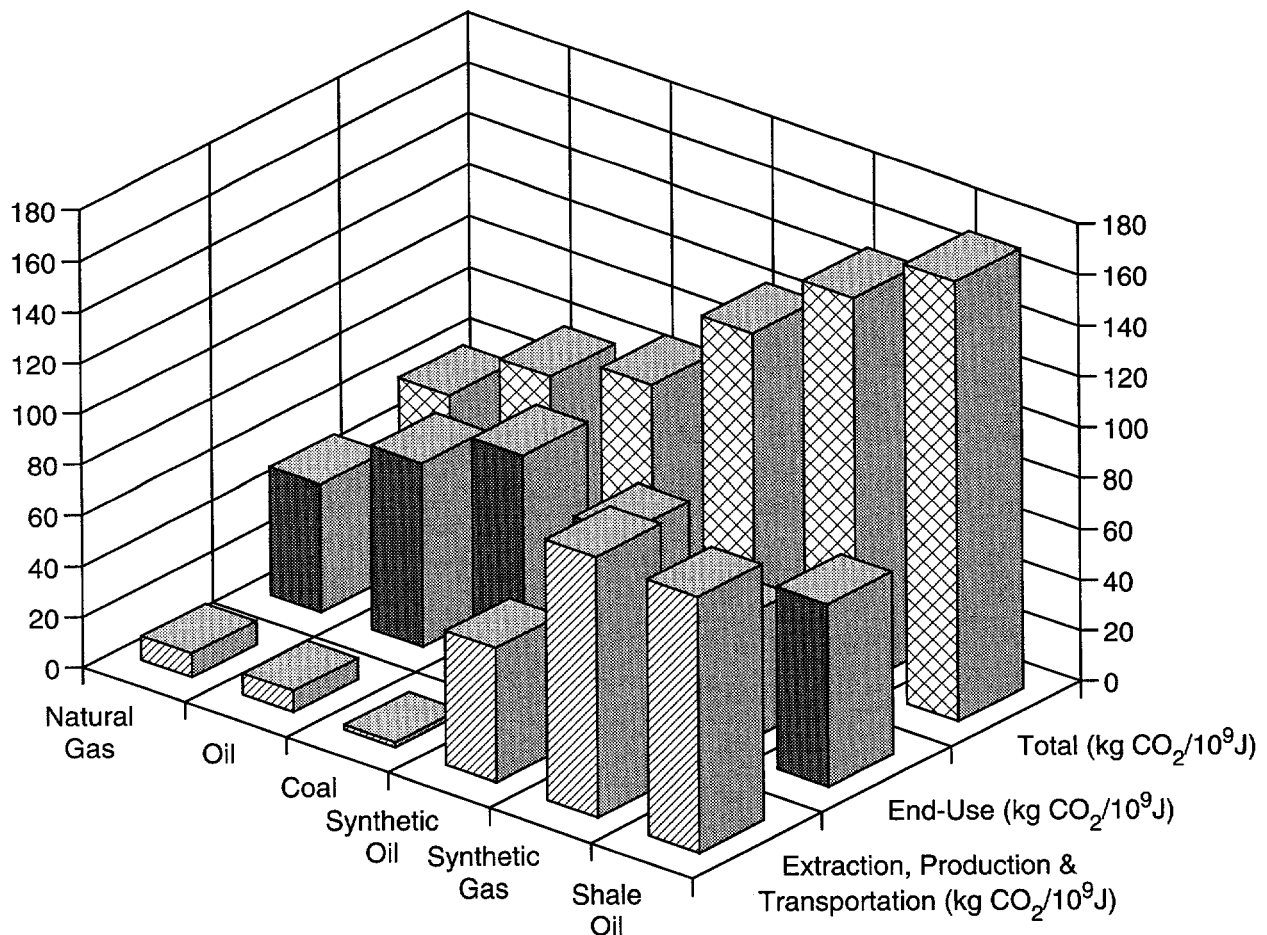


FIGURE 2.1 Rates of CO_2 Emissions for Fossil Fuels, Showing that Coal Has the Lowest CO_2 Emissions for Extraction of Any Fossil Fuel

TABLE 2.1 Typical CO₂ Emissions for U.S. Energy Use of Fossil Fuels

| Fuel | CO ₂ Emission (kg/10 ⁹ J) | | |
|---------------|---|---------|-------|
| | Extraction, Production, and Transportation | End Use | Total |
| Natural gas | 9.37 | 49.9 | 59.3 |
| Oil | 8.41 | 72.2 | 80.6 |
| Coal | 1.92 | 89.1 | 91.0 |
| Synthetic oil | 53.1 | 72.2 | 125.2 |
| Synthetic gas | 103.0 | 49.9 | 152.9 |
| Shale oil | 101.1 | 72.2 | 173.2 |

TABLE 2.2 Coal Losses and Feed Rates for IGCC Base Case

| Coal-Handling Step | % Loss | Feed Rate | |
|--------------------------|----------------|-----------|--------|
| | | kg/h | tons/d |
| Raw coal in mine | — ^a | 168,200 | 4,503 |
| Coal mine preparation | 10 | 153,200 | 4,053 |
| Coal transport (rail) | 3.5 | 152,667 | 4,039 |
| Coal preparation at IGCC | 6.5 | 143,335 | 3,792 |

^a Not applicable.

follows. Although the specific mine used as the basis for this study has an advanced coal-washing facility, the coal shipped to the IGCC plant is not assumed to undergo anything but minimal beneficiation.

2.1.1 Deep Mining

Illinois No. 6 coal was the basis for the KRW design study, and the ultimate analysis appears in Table 2.3. The coal was selected as being from Old Ben No. 26 Mine (Sesser, Illinois). This mine produces coal at approximately 2,500,000 tons/yr from a 651-ft deep-mining operation.

The energy costs associated with deep-mining this coal have been taken from a Bureau of Mines report (Katell et al. 1975). This analysis would lead to an energy consumption of 12.89 kWh/ton of coal, which equates to 2.42 MWh for the base case. An equivalent CO₂ emission of 2,418 kg/h is taken.

The breakdown of this power consumption (Table 2.4) can be estimated from costs that have been projected for the deep-mining of shale oil (Boyd 1980). The very large size of this shale-oil mine, 17.5×10^6 tons/yr, must be compared with the more typical size for an underground coal mine of $1-3 \times 10^6$ tons/yr. Because of this comparison, the Bureau of Mines report (Katell et al. 1975) showed a value that asymptotically approached 12.5 kWh/ton of coal, and this value will be used as the basis for adjusting these parameters. The significant cost increases in the energy for mining coal then appear to come in the areas of increased ventilation (for methane control) and increased dewatering (most coal seams are also excellent aquifers).

2.1.2 Coal-Cleaning Circuit

Additional energy costs must be added to this value. The Old Ben No. 26 Mine has a new coal-preparation plant that became operational in 1985. This preparation plant employs jigs, cyclones, centrifuges, crushers, breakers, and screens (Richardson 1987). This equipment is consistent with a heavy liquid cyclone circuit. The presumption is that coal leaving this circuit will be 28 mesh. An energy and material balance for a heavy liquid cyclone coal-cleaning circuit will use steam (0.065 kg/kg of coal) and electricity (2.388 W/kg of coal) for deeply cleaned coal (Gala et al. 1987); however, these values for electricity and CO₂ emissions were not included in the evaluation of the overall process cycle.

All coal-cleaning plants are designed to produce more than one product stream, and the coal transported to the IGCC plant will be sized at 2-4 in. for shipment. The production of deeply cleaned coal is a disadvantage for gasification because iron pyrite is a well-recognized catalyst for

TABLE 2.3 Base Coal Analysis:
Illinois No. 6 (Old Ben No. 26 Mine)

| Variable | Data |
|--|--------|
| Component, as received (weight %) | |
| Moisture | 11.12 |
| Carbon | 63.75 |
| Hydrogen | 4.50 |
| Nitrogen | 1.25 |
| Chlorine | 0.29 |
| Sulfur | 2.51 |
| Ash | 9.70 |
| Oxygen (by difference) | 6.88 |
| Total | 100.00 |
| Ash fusion temperature (reducing conditions) (°C) | |
| Initial deformation | 1,201 |
| Softening (H=W) | 1,238 |
| Softening (H=1/2 W) | 1,285 |
| Fluid | 1,324 |
| Higher heating value, as received | |
| MJ/kg | 27.11 |
| Btu/lb | 11,666 |

Source: Gallaspy et al. (1990b).

TABLE 2.4 Power Requirements for Underground Mining of Coal

| Operation | Horsepower (hp/ton) | Electricity (kWh/ton) ^a | Coal Change ^b | Estimated Electricity (kWh/ton) | KRW Base Electricity (kWh/ton) |
|------------------|------------------------|---------------------------------------|-----------------------------|---------------------------------------|--------------------------------------|
| Hoisting | 7.18 | 5.38 | nc | 5.38 | 5.55 |
| Drilling | 2.40 | 1.78 | nc | 1.78 | 1.84 |
| Ventilation | 1.23 | 0.93 | ++ | 1.94 | 2.00 |
| Dewatering | 1.51 | 1.13 | ++ | 2.35 | 2.42 |
| Break and convey | 0.86 | 0.64 | nc | 0.64 | 0.66 |
| Product pumps | 2.19 | 1.65 | — | 0 | 0 |
| General services | NA ^c | 0.41 | nc | 0.41 | 0.41 |
| Total | NA | 11.92 | NA | 12.50 | 12.89 |

^a Basis: shale oil at 70,000 tons/d (17.5×10^6 tons/yr).

^b nc, No change; ++, large increase; and —, large decrease.

^c NA, Not applicable.

gasification. While the primary crushers and tramp metal removal associated with this plant are used to prepare the coal product shipped to the IGCC plant, these energy costs have already been accounted for in the mining energy costs. No net additional impact occurs from this circuit.

2.2 Rail Transport

At the present time, the location of the IGCC plant has not been specified beyond being in the midwestern United States. In future work, predicting a site more accurately may be feasible (Fisher et al. 1991). The assumption will be made that the plant is located 161 km (100 mi) from Sesser, Illinois. The energy used in the rail transport of the coal comes to 0.08 W·h/kg coal·km (Hall et al. 1986). This figure translates to an energy cost of 1.966 MW for the rail transport of the coal. Only 5% of this energy consumption is not taken to be electric energy (0.1 MW; CO₂ emission of 100 kg/h). The remainder of the energy is supplied by the diesel fuel, with an equivalent CO₂ emission of 537 kg/h. This sum leads to a net CO₂ emission of 637 kg/h.

2.3 Impact of Plant Location on CO₂ Budget

One of the most significant questions that needs to be addressed in the event that CO₂ capture from power plants were to be necessary is this: What constitutes the optimal balance

among these parameters: locating the plant near the source of coal, near the region with the demand for the electric power, or near a reservoir that could accept the CO₂? The regional demand question will not be addressed in this study, thus reducing the question to: Is it better to locate the IGCC plant near the source of the coal (and minimize rail transport) or to locate the plant near a suitable disposal reservoir (and minimize the length of a CO₂-carrying pipeline)?

From a standpoint of energy use, the data from Hall et al. (1986) can be examined to provide a definitive answer. The energy consumption for rail transportation for the cleaned coal from the Old Ben No. 26 Mine should be 0.125 W·h/kg C·km. At the same time, a CO₂ pipeline should operate with an energy cost of 0.0187 W·h/kg C·km. This rail/pipeline energy cost ratio is 6.69:1. Hence, strictly from the standpoint of energy use, a shorter rail line and a longer pipeline appear optimal; however, the energy cost for building a pipeline, as compared with a new (or existing) rail line, has not been included.

2.4 Economics

Table 2.5 summarizes the power consumption and CO₂ emissions from the coal deep-mining operation, the coal-cleaning circuit, and rail transport. Because the coal is transported by diesel train, the power use on a national basis is nearly equal to the direct electric consumption in mining. The total electric power used from the start of mining to the IGCC plant's boundary is 2.52 MW, with an associated CO₂ emission of 3,055 kg/h.

2.5 Limestone Mining

Limestone is used as an in-bed sulfur-capture technique for the KRW base case. Limestone will be extracted from a surface mine, with the breakdown of power consumption (Table 2.6) estimated from projected costs for mining shale oil (Boyd 1980). The total limestone consumption for the IGCC system is 39,795 kg/h (87,733 lb/h = 1,052 tons/d). Hence, the total electric power consumption is 0.257 MW, with an equivalent CO₂ emission of 257 kg/h.

TABLE 2.5 Power Consumption and CO₂ Emissions from Coal Mining, Cleaning, and Transport

| Source | Power (MW) | CO ₂ Emission (kg/h) |
|------------------|----------------|---------------------------------|
| Deep coal mining | 2.42 | 2,418 |
| Coal cleaning | — ^a | — |
| Rail transport | 0.1 | 537 |
| Total | 2.52 | 3,055 |

^a Values for coal-cleaning circuit were not included.

TABLE 2.6 Power Requirements for Surface Mining of Limestone

| Operation | Horsepower (hp/ton) | Electricity (kWh/ton) ^a | Limestone Change ^b | Estimated Electricity (kWh/ton) |
|------------------|------------------------|---------------------------------------|----------------------------------|---------------------------------------|
| Hoisting | 7.18 | 5.38 | — | 1.35 |
| Drilling | 2.40 | 1.78 | nc | 1.78 |
| Dewatering | 1.51 | 1.13 | + | 1.70 |
| Break and convey | 0.86 | 0.64 | nc | 0.64 |
| General services | NA ^c | 0.41 | nc | 0.41 |
| Total | NA | NA | NA | 5.88 |

^a Basis: shale oil at 70,000 tons/d (17.5×10^6 tons/yr).

^b nc, No change; —, decrease; and +, increase.

^c NA, Not applicable.

2.6 Limestone Rail Transport

At the present time, the location of the IGCC plant has not been specified beyond being in the midwestern United States. In future work, predicting a site more accurately may be feasible (Fisher et al. 1991). The assumption will be made that the plant is located 161 km (100 mi) from the limestone mine. By using the data from Bayley (Hall et al. 1986), this assumption translates to an energy cost of 0.08 W·h/kg, which is taken to be consumed as diesel fuel, with negligible electric consumption of 5% of this number. This leads to a direct CO₂ emission of 140 kg/h (Table 2.7) from diesel fuel, for a net CO₂ emission of 160 kg/h.

TABLE 2.7 Power Consumption and CO₂ Emissions from Limestone Mining and Transport

| Source | Power (MW) | CO ₂ Emission (kg/h) |
|--------------------------|---------------|---------------------------------------|
| Surface limestone mining | 0.257 | 257 |
| Transport | 0.02 | 160 |
| Total | 0.277 | 417 |

3 Kellogg-Rust-Westinghouse IGCC: Coal and Limestone Preparation

The coal and limestone handling system assumes unloading from a unit train to vibrating feeders that pass the material through magnetic separation before belt conveyors unload the material into a series of 14-h storage silos. The coal entering the preparation plant measures 2-4 in.

Coal cleaning results in significant reduction of both sulfur and ash; however, further pulverization of the coal must take place before the material is fed to the gasifiers. This pulverization will reduce the coal to 1/4-in. material. The coal is crushed and dried in a series of three fluidized-bed roller mills (with a fourth unit as a spare). Drying is accomplished by the hot flue gas (760°C) from the IGCC sulfator process. This drying results in a significant CO₂ emission from the energy cycle that was not reclaimed but presents a good opportunity for further reductions. The coal is then held in a 2-h bunker, from which it is pneumatically conveyed to surge bins ahead of the gasifier lock hoppers. The power requirements for this system were consistent with available design reports (Miller 1985; Condorelli et al. 1991), and a value of 18.3 W·h/kg of coal was used. Hence, for this case, the coal preparation directly consumes 2.35 MW, with an equivalent CO₂ emission of 2,351 kg/h. The sulfator emits CO₂ at 13,099 kg/h while consuming 2.6 MW.

Limestone is prepared by crushing through two pulverizers and also is pneumatically conveyed to a 24-h storage silo and 2-h storage bunker and is mixed with the coal in the gasifier surge bins. The power requirements for this system were consistent with available design reports, and a value of 30.9 W·h/kg of limestone was used. Hence, for this case, the direct power consumption is 1.23 MW, with an equivalent CO₂ emission of 1,230 kg/h. This information is summarized in Table 3.1.

TABLE 3.1 Power Consumption and Equivalent CO₂ Emissions for Coal and Limestone Preparation

| Source | Power (MW) | CO ₂ Emission (kg/h) |
|-----------------------|-------------------|---------------------------------|
| Coal preparation | 2.35 | — ^a |
| Sulfator gas | 2.6 | 13,099 |
| Limestone preparation | 1.23 | — ^a |
| Total | 6.18 ^b | 13,099 |

^a Carbon dioxide emissions are not counted because electricity comes from in-plant use.

^b This total is all in-plant energy use.

4 Kellogg-Rust-Westinghouse IGCC: Gasifier

In the case of direct coal combustion, the recovery of CO_2 is complicated by dilution of the flue gas with nitrogen (N_2) from the combustion air, as well as with excess air, which invariably accompanies coal combustion. For a representative boiler using bituminous coal, the mole fraction of CO_2 in the flue gas is in the range of 10-15% (Babcock & Wilcox Co. [B&W] 1975). For coal gasification, no excess air is provided. In fact, the air supplied is inadequate for complete combustion. The mole fraction of CO plus CO_2 in the fuel gas from the KRW gasifier is about 22% for an air-blown gasifier and 59% for an oxygen-blown gasifier (Gallaspy et al. 1990a). The CO can be converted to CO_2 via the water-gas shift reaction. Because of the higher CO_2 partial pressure, recovery of the CO_2 from this shifted synthesis gas requires a less costly recovery system than that which would be required for recovery following direct coal combustion in air. Sections 4 through 10 are intended to provide a first-order estimate of the effects of CO_2 recovery on the cost and performance of a gasifier.

4.1 Overview

4.1.1 Description of the Gasifier

The baseline system for gasification is the air-blown KRW gasifier, which is an agglomerating fluidized-bed gasifier. For the base-case system (case 1 [Gallaspy et al. 1990a]), hot-gas cleanup is employed.¹ The baseline capacity of the plant is 458 MW net. Figure 4.1 is a simplified flow diagram of the configuration in the base case with the system for CO_2 recovery added. Introduction of CO_2 recovery will reduce the net output of the plant because of additional energy losses incurred in various stages of CO_2 recovery. Specific energy penalties include the work of pumping and compression, the demand for heat or steam for solvent recovery, and thermal energy losses that are impractical to recover. In addition, some modifications to the combustion turbine will be required because of the accompanying change in the composition of the synthesis gas. No accounting of this latter effect is included in the present analysis.

Because no additional capacity for gasification has been added to compensate for energy losses due to the CO_2 recovery system, the output of the gasifier from the base case is taken to be the raw gas input to the CO_2 recovery system. The configuration of the gasifier in the base case uses four trains, each of which includes all major process equipment. Therefore, the preliminary design of the CO_2 recovery system described here also employs four trains, each at 25% of the necessary total capacity. The descriptions of the CO_2 recovery process, including equipment specifications, flow rates, and power requirements, refer to a single train. Descriptions of overall plant output and energy use refer to the aggregate of all four trains.

¹ The use of hot-gas cleanup on a system using solvent-based CO_2 recovery is of no value to the efficiency of the system because the temperature of the product gas must be reduced for effective absorption.

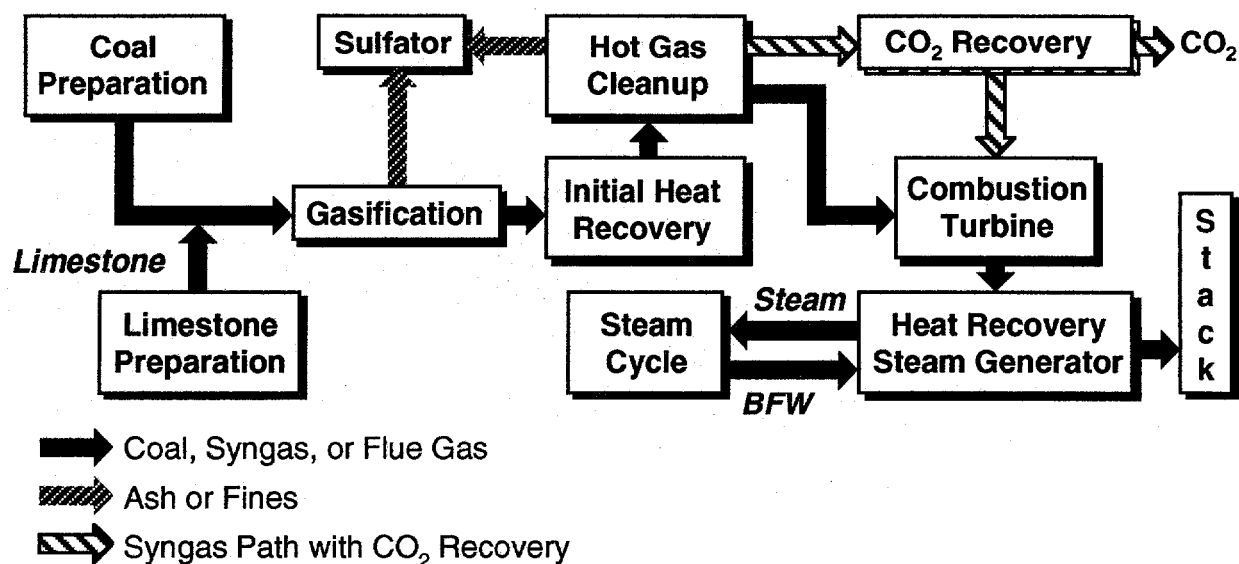
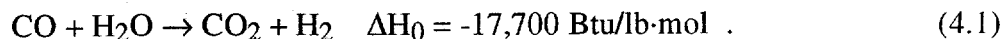


FIGURE 4.1 Primary Flows in KRW Base Case 1 (IGCC) and Location of CO₂ Recovery System

4.1.2 Shift Reaction

The raw synthesis gas from the KRW process is high in CO. Conversion of that CO to CO₂ in the combustion process would result in substantial dilution of the resulting CO₂ with N₂ from the combustion air and with water (H₂O) from the combustion reaction. If the CO₂ is removed prior to completing combustion, a substantial savings in the cost of the CO₂ recovery system is possible because of the reduced size of the gas processing equipment and the reduced flow rate of solvent. The CO in the raw gas must first be converted to CO₂ in a shift reactor according to Equation 4.1. The reaction is exothermic. The resulting CO₂ is recovered, leaving a fuel rich in hydrogen (H₂) for use in the gas turbine:



Minimizing the cost of heating and cooling requirements is an important issue for process integration. Raw gas is obtained from the KRW process at 811 K (1,000°F). The raw gas must be cooled prior to the shift reaction so that equilibrium favors the products in reaction 4.1. Table 4.1 lists the equilibrium constant (*K*) for reaction 4.1 as a function of temperature. As discussed subsequently, the shift reaction is accomplished in two stages, with additional cooling required between stages for further temperature reduction and for removal of the heat of reaction. The shift reactor is configured for 95% conversion of CO to CO₂.

TABLE 4.1 Temperature Sensitivity
of Equilibrium Constant
for Shift Reaction

| Temperature (°F) | K^a |
|---------------------|-------|
| 400 | 200.0 |
| 600 | 32.0 |
| 800 | 9.0 |
| 1,000 | 3.7 |
| 1,200 | 1.9 |

$$^a K = ([\text{CO}_2] [\text{H}_2]) / ([\text{CO}] [\text{H}_2\text{O}]).$$

Source: Adapted from Imperial
Chemical Industries, Ltd. (1970,
Appendix 5b).

4.1.3 Recovery Process

The CO₂ recovery processes examined in this study are absorption processes in which CO₂ is absorbed by a solvent. Depending on the process, CO₂ may chemically interact with the solvent. The general character of these processes, including gas composition changes and utility requirements, is clarified by an example. The Selexol® process is an absorption process that uses a proprietary solvent, the dimethyl ether of polyethylene glycol with additives.² The solubility of CO₂ in this solvent is highly dependent on temperature and pressure. In order to absorb significant amounts of CO₂ at economic rates of solvent flow, the process must be operated at near ambient temperature or below and at elevated pressure. Solvent regeneration is accomplished by flashing (pressure reduction) or by heating. Figure 4.2 shows the composition of the gas stream at various points in the process. The changes in composition are essentially the same as these changes for all of the absorption processes.

With 95% removal of CO₂ by the glycol solvent (Selexol) process following 95% conversion of CO in the shift reactor, the overall carbon removal by the system is 90%. This figure accounts for the presence of methane (CH₄) in the synthesis gas.

² Because the Selexol process is a proprietary process, much of the data required for even a preliminary design is not available in the open literature. Where engineering information is not available, this analysis relies on basic principles and assumptions. A representative of Union Carbide, the current holder of the Selexol license, has reviewed the preliminary design.

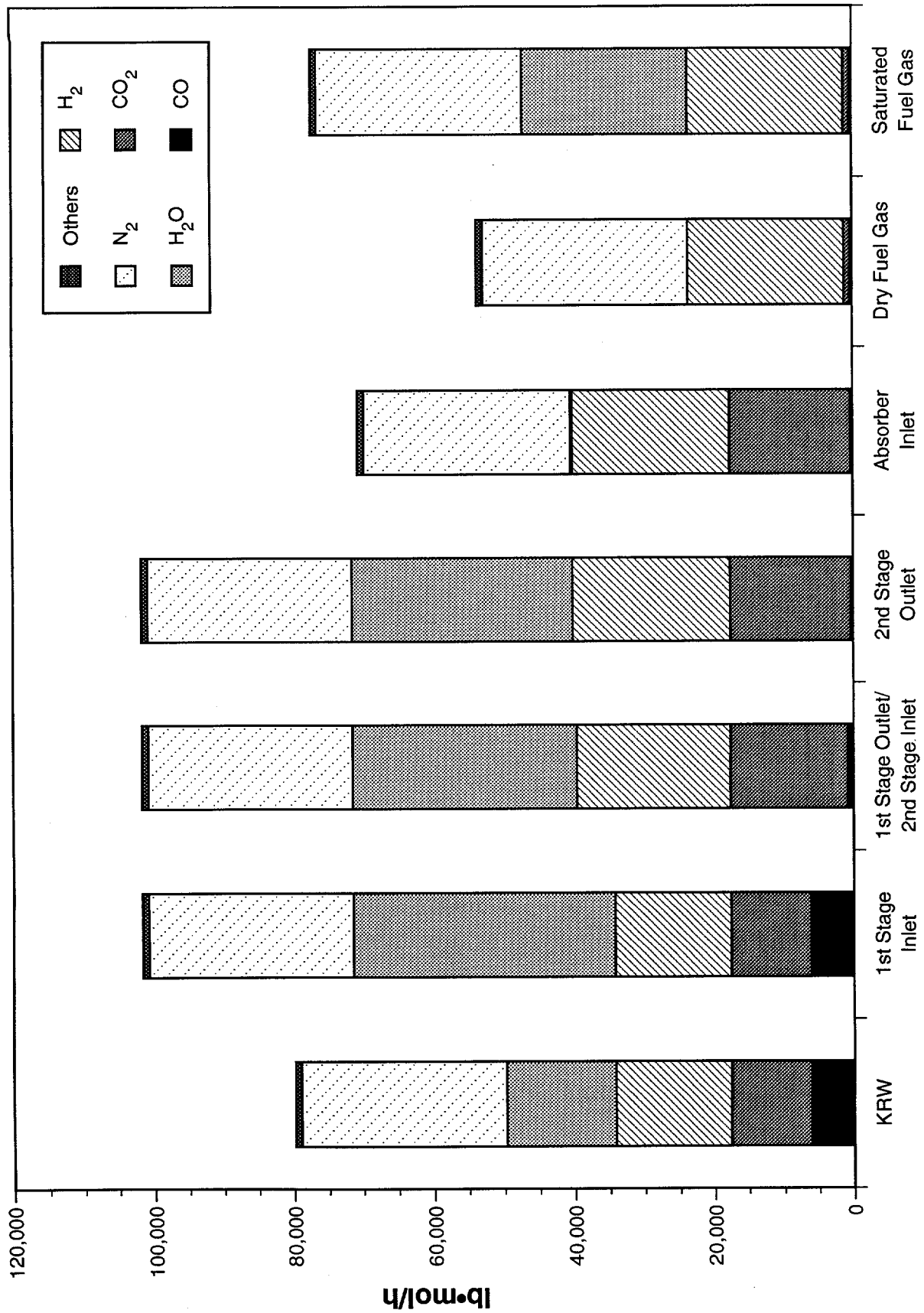


FIGURE 4.2 Gas Stream Composition at Various Stages in the Process

The reduction of CO₂ emissions that is achieved by 90% removal is evident from the composition of the gas turbine exhaust in the base (uncontrolled) case. That exhaust gas composition reflects the composition of fuel gas from the gasifier, vent gas from sulfur recovery, and combustion and cooling air for the turbine. Tables 4.2 and 4.3 display the composition of these streams and the flow rates of each component. Note that the CO₂ flow rate of 18,175 lb-mol/h at full load represents nearly 8% of the exhaust stream. Nitrogen is the primary component of the exhaust stream, as is expected for any air-fired combustion system.

4.2 Shift Reaction

The shift reaction is commonly accomplished in a catalyst-packed tubular reactor. A relatively low-cost iron oxide catalyst is effective in the temperature range of 650-1,100°F. Below that temperature, a copper oxide catalyst is required, at considerably higher unit cost. While the copper oxide cost is higher, this fact must be balanced against the fact that the equilibrium concentration of each product in Equation 4.1 strongly recommends lower temperatures. On the other hand, the reaction rate is reduced as temperature is reduced, requiring a larger vessel to

TABLE 4.2 Composition of Combustion Chamber Inlet Gas

| Variable | Fuel Gas | | Vent Gas | | Combustion Air | |
|------------------|----------|---------------|----------|---------------|----------------|---------------|
| | lb-mol/h | mole fraction | lb-mol/h | mole fraction | lb-mol/h | mole fraction |
| Gas | | | | | | |
| CO | 6,187 | 0.078 | 4 | 0.077 | 0 | 0.000 |
| CO ₂ | 11,461 | 0.144 | 8 | 0.144 | 0 | 0.000 |
| H ₂ | 16,436 | 0.206 | 11 | 0.208 | 0 | 0.000 |
| H ₂ O | 15,489 | 0.194 | 11 | 0.195 | 0 | 0.000 |
| N ₂ | 29,213 | 0.367 | 20 | 0.366 | 123,124 | 0.781 |
| Ar | 355 | 0.004 | 0 | 0.005 | 1,513 | 0.010 |
| CH ₄ | 514 | 0.006 | 0 | 0.007 | 0 | 0.000 |
| NH ₃ | 12 | 0.000 | 0 | 0.000 | 0 | 0.000 |
| H ₂ S | 2 | 0.000 | 0 | 0.000 | 0 | 0.000 |
| HCl | 0 | 0.000 | 0 | 0.000 | 0 | 0.000 |
| O ₂ | 0 | 0.000 | 0 | 0.000 | 32,932 | 0.209 |
| COS | 0 | 0.000 | 0 | 0.000 | 0 | 0.000 |
| SO ₂ | 0 | 0.000 | 0 | 0.000 | 0 | 0.000 |
| NO | 0 | 0.000 | 0 | 0.000 | 0 | 0.000 |
| Total gas flow | 79,669 | 1.000 | 55 | 1.000 | 157,569 | 1.000 |
| Temperature (°F) | 1,000 | | 524 | | 760 | |
| Pressure (psia) | 295 | | 750 | | 200 | |

TABLE 4.3 Composition of Turbine Inlet Gas and Turbine Exhaust Gas

| Variable | Combustion Gases | | Turbine Cooling Air | | Turbine Exhaust Gas | |
|------------------|------------------|---------------|---------------------|---------------|---------------------|---------------|
| | lb-mol/h | mole fraction | lb-mol/h | mole fraction | lb-mol/h | mole fraction |
| Gas | | | | | | |
| CO | 0 | 0.000 | 0 | 0.000 | 0 | 0.000 |
| CO ₂ | 18,175 | 0.080 | 0 | 0.000 | 18,175 | 0.076 |
| H ₂ | 0 | 0.000 | 0 | 0.000 | 0 | 0.000 |
| H ₂ O | 32,995 | 0.146 | 0 | 0.000 | 32,995 | 0.139 |
| N ₂ | 152,357 | 0.674 | 9,213 | 0.781 | 161,570 | 0.680 |
| Ar | 1,868 | 0.008 | 113 | 0.010 | 1,981 | 0.008 |
| CH ₄ | 0 | 0.000 | 0 | 0.000 | 0 | 0.000 |
| NH ₃ | 0 | 0.000 | 0 | 0.000 | 0 | 0.000 |
| H ₂ S | 0 | 0.000 | 0 | 0.000 | 0 | 0.000 |
| HCl | 0 | 0.000 | 0 | 0.000 | 0 | 0.000 |
| O ₂ | 20,567 | 0.091 | 2,464 | 0.209 | 23,031 | 0.097 |
| COS | 0 | 0.000 | 0 | 0.000 | 0 | 0.000 |
| SO ₂ | 2 | 0.000 | 0 | 0.000 | 2 | 0.000 |
| NO | 12 | 0.000 | 0 | 0.000 | 12 | 0.000 |
| Total gas flow | 225,976 | 1.000 | 11,790 | 1.000 | 237,766 | 1.000 |
| Temperature (°F) | 2,208 | | 760 | | 1,098 | |
| Pressure (psia) | 194 | | 200 | | 14.9 | |

provide the necessary reaction time. In view of these conflicting influences, high CO₂ recovery is best achieved in a staged reactor. In the first stage, iron oxide catalyst is used to convert about 87.1% of the CO. In the second stage, copper oxide catalyst is used to convert about 62.1% of the remaining CO, for an overall removal efficiency of 95%. This scheme for the shift reaction is represented in Figure 4.3. The temperatures and compositions of the streams and the heating and cooling requirements are summarized in Table 4.4.

To optimize the design of this two-stage system, the cost corresponding to a given temperature and fraction conversion in each stage must be estimated. The cost-estimating procedure is discussed subsequently for the selected design configuration, which is expected to be close to the optimal configuration. Comprehensive optimization has not been attempted; rather, a configuration that minimizes the cost of the catalysts is chosen. The total cost includes both capital and operating components. The operating cost includes hot and cold utilities, maintenance, and catalyst replacement. The capital cost is due primarily to the reaction vessels and initial catalyst. Therefore, estimating the volume of catalyst, which is determined by the space velocity required to

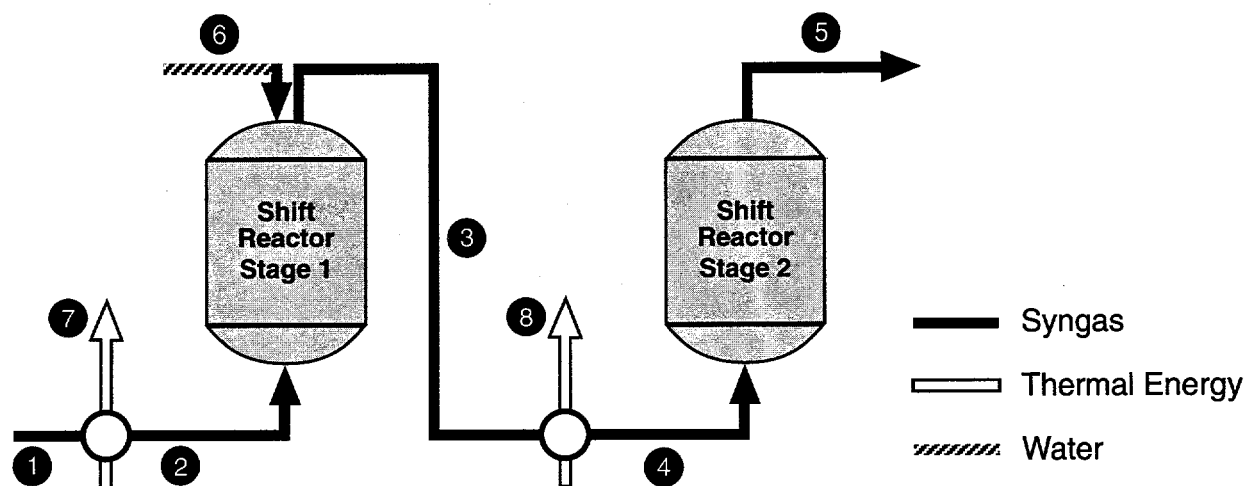


FIGURE 4.3 Main Components of Shift Reactor Process (numbers refer to streams)

TABLE 4.4 Composition and Enthalpy of Streams

| Stream | Composition (mole fraction) | | | | | | Stream Enthalpy (Btu/h) |
|--------|-----------------------------|-----------------|----------------|------------------|----------------|-------|-------------------------|
| | CO | CO ₂ | H ₂ | H ₂ O | N ₂ | Other | |
| 1 | 0.078 | 0.144 | 0.208 | 0.194 | 0.366 | 0.011 | 2.28×10^8 |
| 2 | 0.078 | 0.144 | 0.208 | 0.194 | 0.366 | 0.011 | 1.41×10^8 |
| 3 | 0.008 | 0.166 | 0.216 | 0.313 | 0.288 | 0.009 | 2.82×10^8 |
| 4 | 0.008 | 0.166 | 0.216 | 0.313 | 0.288 | 0.009 | 2.33×10^8 |
| 5 | 0.003 | 0.171 | 0.221 | 0.308 | 0.288 | 0.009 | 2.36×10^8 |
| 6 | 0 ^a | 0 | 0 | 1.0 | 0 | 0 | 1.17×10^8 |
| 7 | 0 | 0 | 0 | 0 | 0 | 0 | 8.71×10^7 |
| 8 | 0 | 0 | 0 | 0 | 0 | 0 | 4.89×10^7 |

^a Streams do not include these gases.

achieve a given conversion efficiency, is necessary. The space velocity (SV) is the ratio of the volumetric flow rate for the processed gas (Q) over the catalyst volume. Therefore:

$$\text{catalyst volume} = Q/SV . \quad (4.2)$$

A higher space velocity implies greater effectiveness for the catalyst and a consequent reduction in the amount of catalyst required. The space velocity is related to the fraction conversion (x) and the reaction rate (r) according to Equation 4.3:

$$SV^{-1} = \int_0^x \frac{dx}{r} . \quad (4.3)$$

Equation 4.4 is a reaction rate expression, which may be substituted into Equation 4.3 and integrated to yield Equation 4.5, relating the space velocity to the fraction of CO converted, the equilibrium constant, the reaction rate, and the initial concentrations (Moe 1962):

$$r = k \left[(A_0 - x)(B_0 - x) - \frac{(C_0 + x)(D_0 + x)}{K} \right] , \quad (4.4)$$

where

k = reaction rate constant,

A_0 = initial concentration of CO,

B_0 = initial concentration of water,

C_0 = initial concentration of CO_2 ,

D_0 = initial concentration of H_2 , and

x = fraction conversion of CO to CO_2 ;

and

$$\frac{k}{SV} = \frac{K}{\sqrt{-q}} \left\{ \ln \left[\frac{2wx - u - \sqrt{-q}}{2wx - u + \sqrt{-q}} \right] - \ln \left[\frac{-u - \sqrt{-q}}{-u + \sqrt{-q}} \right] \right\} , \quad (4.5)$$

where

$$w = K - 1,$$

$$\sqrt{-q} = \sqrt{u^2 - 4wv},$$

$$u = K(A_0 + B_0) + (C_0 + D_0), \text{ and}$$

$$v = K(A_0B_0) - (C_0D_0).$$

If the rate and equilibrium constants are known, Equation 4.5 can be used to estimate the space velocity. The following correlations can be used to estimate the rate and equilibrium constants (Moe 1962):

$$\log_{10} \frac{k}{A_p} = 6.947 - \frac{3,830}{^\circ R} \quad (4.6)$$

$$K = \exp\left(\frac{8,240}{^\circ R} - 4.33\right). \quad (4.7)$$

The reaction rate constant (k) is for a specific commercial iron-based catalyst (Girdler Catalyst type G-3A).

The variable A_p is a pressure-dependent activity factor, as shown in Figure 4.4. The assumption is made that the gasifier exit pressure, 295 psia, is maintained in the shift reactor, leading to an activity factor of about 3.95. For the sake of conservatism in the design, an activity factor of 3.0 has been used in these calculations. For the copper-based catalyst used in the second-stage shift reactor, a published correlation for reaction rate is provided by Equation 4.8 (Campbell et al. 1970):

$$\log_{10} \frac{k}{A_p} = 6.91 - \frac{3,062}{^\circ R}. \quad (4.8)$$

The pressure correction, A_p , as shown in Figure 4.4, is assumed to apply for this copper-based catalyst.

In addition to Equation 4.7, the equilibrium constant is defined by Equation 4.9,

$$K = \frac{[\text{CO}_2][\text{H}_2]}{[\text{CO}][\text{H}_2\text{O}]}, \quad (4.9)$$

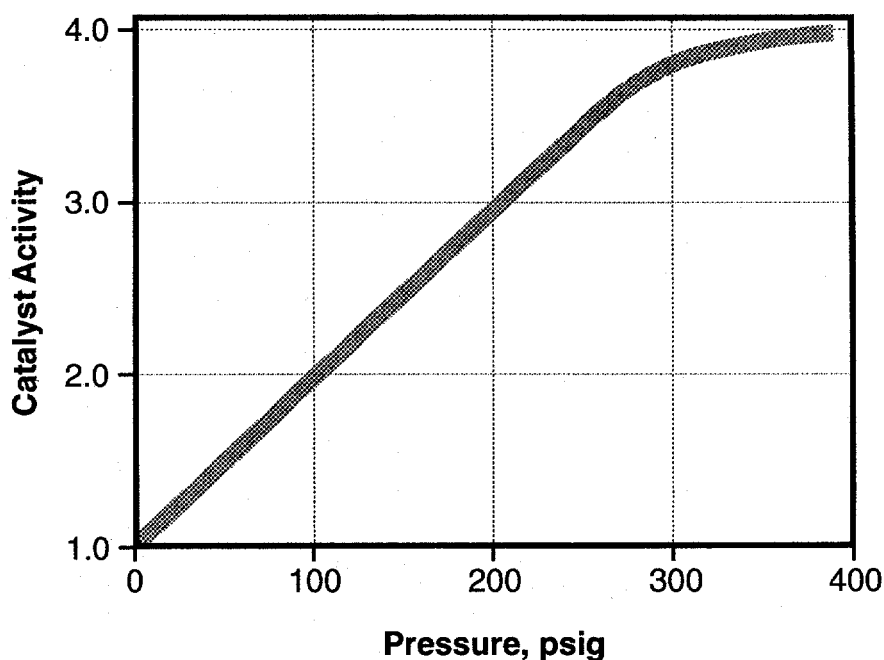


FIGURE 4.4 Catalyst Activity as a Function of Pressure (Source: Moe 1962) (Reproduced by permission of the American Institute of Chemical Engineers. ©1962 AIChE. All rights reserved.)

which can also be expressed as Equation 4.10:

$$K = \frac{(C_0 + x_e)(D_0 + x_e)}{(A_0 - x_e)(B_0 - x_e)}, \quad (4.10)$$

where

A_0 , B_0 , C_0 , and D_0 are as defined for Equation 4.4, and

x_e = fraction conversion at equilibrium.

Equation 4.9 shows that the addition of steam at a specific temperature (fixed K) will increase the amount of product. This effect is shown in Figure 4.5, a plot of conversion fraction as a function of the initial ratio of steam to CO. This strategy for increasing conversion has practical limitations because the increased water results in increased flow, effectively reducing residence time for a given reactor volume. This effect is particularly pronounced at higher temperatures.

The addition of liquid water, which is evaporated in the reactor, can provide some of the needed cooling to bring the raw gas to the desired temperature for the first-stage shift reaction. Additional cooling is provided by external heat exchange. Cooling for the second-stage shift reaction is accomplished entirely by heat exchange to avoid further dilution of the gas stream.

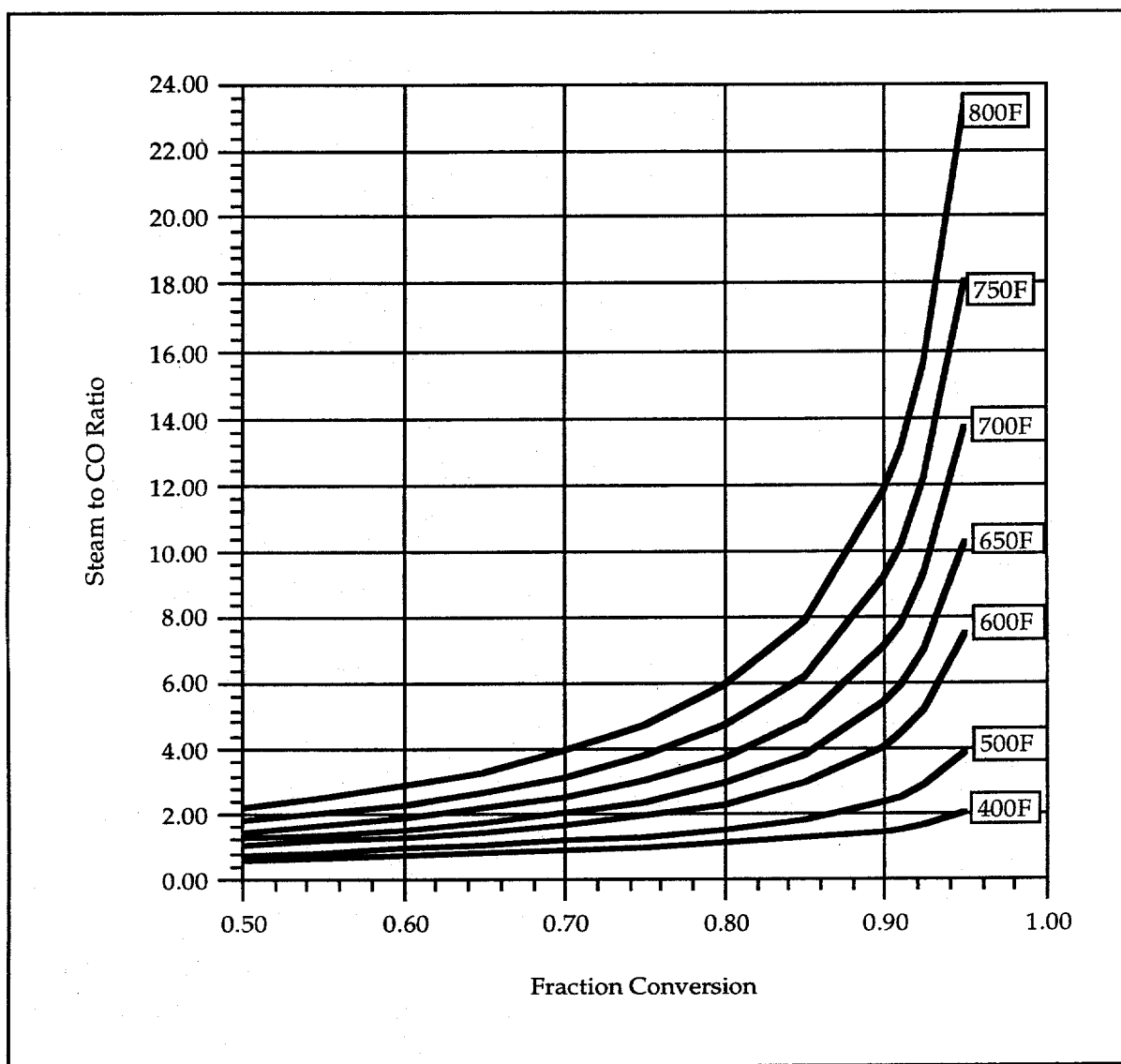


FIGURE 4.5 Effects of Steam-to-CO Ratio and Temperature on Equilibrium Conversion of CO to CO_2

With reference to Figure 4.5, 87% conversion in the first-stage reactor can be achieved at a temperature of about 680°F with a steam-to-CO ratio of about 6 if equilibrium is attained. The actual conversion accomplished is somewhat less, depending on the amount of catalyst provided and the temperature. The relationship is as indicated in Equation 4.5.

Figure 4.6 shows the effect of temperature on space velocity for the first-stage shift reactor conversion. Note that minimum catalyst volume is achieved at maximum space velocity. Thus, for a specified conversion requirement, the optimum temperature is defined by the maximum on the corresponding curve. This relationship represents the trade-off between reaction rate, which is favored by high temperature, and the extent of conversion, which is favored by low temperature. Figure 4.7 shows similar results for the second-stage reactor, where the extent of conversion is

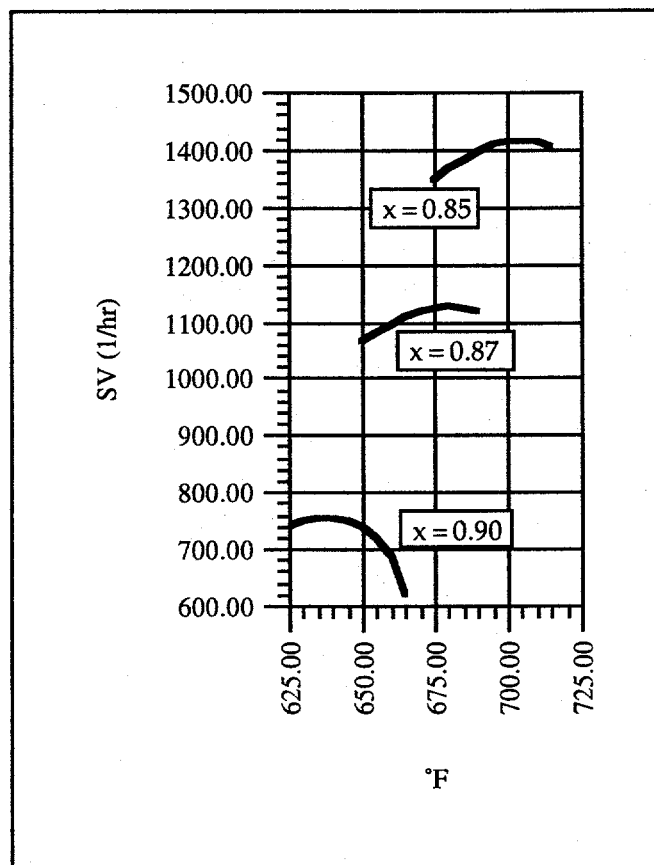


FIGURE 4.6 First-Stage Space Velocity (SV) as a Function of Temperature

complementary to the values in the first stage, so that an overall conversion of 95% is attained. The optimum choice depends on the relative cost of the first-stage and second-stage catalysts. The unit cost of the second-stage catalyst is about five times the unit cost of the first-stage catalyst. Therefore, because the volume of catalyst is inversely proportional to the space velocity, the total cost of catalyst is proportional to the expression, $1/SV_1 + 5/SV_2$, where SV_1 and SV_2 are, respectively, the first-stage and second-stage space velocities. For the three first-stage conversion options presented here, this expression is a minimum for the intermediate case; that is, 87% conversion in the first stage and 62% conversion in the second stage will minimize the total cost of catalyst.

4.3 Heat Recovery and Process Integration

Preparation of the CO₂-rich synthesis gas for processing by the CO₂ recovery system involves substantial cooling because of the high gasifier exit temperature and the exothermic shift reaction. This cooling is in addition to the initial cooling of the synthesis gas from 1,600°F to 1,000°F in a heat recovery unit that is used in the base plant to control the gas inlet temperature to

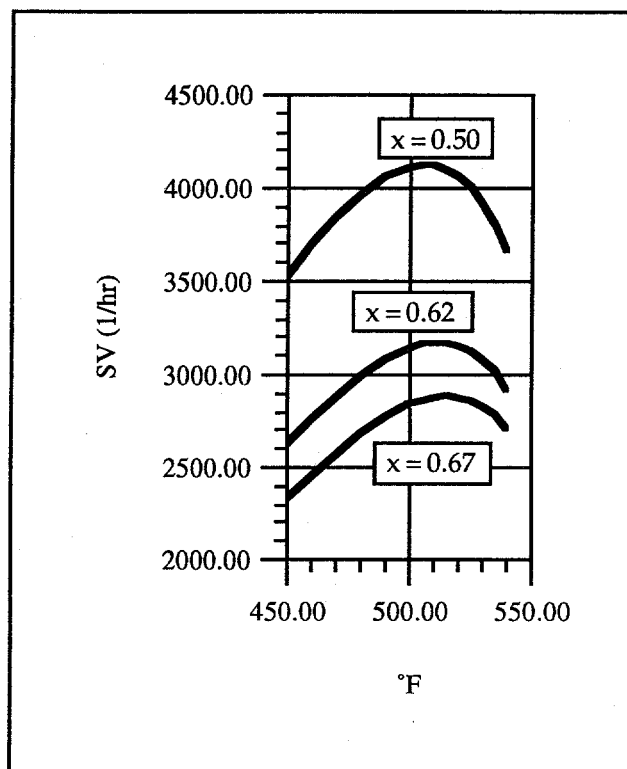


FIGURE 4.7 Second-Stage Space Velocity (SV) as a Function of Temperature

the combustion turbine. The heat removed during cooling must be recovered and integrated into the system to minimize the energy penalty associated with CO₂ recovery. Table 4.5 lists the temperatures at which thermal energy is available from the gas preparation processes and the rate at which the energy is available. Three options have been evaluated for recovery of this thermal energy. A fourth option takes integration one step further by better using the thermal energy from the initial cooling of the high-temperature synthesis gas.

- *Option 1: Steam Generation Only.* All of the available thermal energy is used for feedwater heating, steam generation, and superheating to produce steam for the high-pressure steam turbine. The heat recovery steam generator (HRSG) can be designed to use heat from these process streams.
- *Option 2: Fuel Gas Preheating with Supplemental Steam Generation.* The recovered thermal energy is applied as far as possible for preheating the fuel gas stream to the gas turbine. That thermal energy cannot be completely consumed in fuel gas preheating because the fuel gas will approach the temperature at which heat is available (1,000°F maximum) before all of the available thermal energy is absorbed. Heat that cannot be used for fuel gas preheating is used in the HRSG.

TABLE 4.5 Heat Available from Preparation of CO₂-Rich Gas for a Glycol Solvent Process

| Variable | Initial Syngas Cooling | Heat Recovery | | |
|--|------------------------------|--------------------------------|----------------------------|--------------------------------|
| | | Before First-Stage Shift | Between Shift Stages | After Second-Stage Shift |
| Temperature for CO ₂ -rich gas stream (°F) | | | | |
| Inlet | 1,600 | 1,000 | 680 | 480 |
| Outlet | 1,600 | 482 | 460 | 100 |
| Thermal energy available (10 ⁶ Btu/h) | | | | |
| | 1,299.3 | 348.2 | 195.8 | 864.7 |

- *Option 3: Fuel Gas Saturation and Preheating.* Moisture that has condensed out of the synthesis gas stream during the final cooling stage of the gas preparation is injected into the fuel gas as it is heated by recovered heat. This procedure allows additional heat to be absorbed as preheating for the fuel gas and increases the mass flow rate through the gas turbine. The balance of the available thermal energy is used in the HRSG for feedwater heating and steam generation. The saturated gas can be heated to about 500°F.
- *Option 4: Saturation and Higher-Temperature Preheating of Fuel Gas.* In the base plant (no CO₂ recovery), synthesis gas is cooled from 1,600°F to 1,000°F in a heat recovery unit that provides heat to the steam cycle. By this means, the gas supply to the combustor is limited to 1,000°F. In the CO₂ recovery cases, the treated and saturated fuel gas can be heated to 1,000°F with use of this higher-temperature source and only the remaining thermal energy applied to the steam cycle.

Table 4.6 displays the allocation of available thermal energy to steam for the high-pressure steam turbine and to fuel for the gas turbine for each option. The net plant output is the output from the steam and gas turbines less the power demand of the CO₂ recovery system and other internal plant demand. Table 4.7 summarizes the turbine output and internal demands for a plant designed with each of the heat recovery options and for the base case, which does not include CO₂ recovery. Note that option 4 results in substantially higher net output. The values for power output have been calculated for the low-pressure glycol solvent case, which requires an absorber feed temperature of 100°F. Option 4 offers a similar advantage for the other CO₂ recovery processes, although the net plant output for those processes will differ from that shown in Table 4.7. Therefore, option 4 has been selected as the heat recovery strategy for all of the CO₂ recovery processes.

TABLE 4.6 Options for Heat Recovery from CO₂-Rich Gas Preparation

| Option | Thermal Energy (10 ⁶ Btu/h) | |
|--|---|-------------|
| | To Steam | To Fuel Gas |
| Option 1: Steam generation only | 2,247.9 | 0 |
| Option 2: Preheat fuel gas; raise steam with excess heat | 2,082.9 | 165.0 |
| Option 3: Saturate and preheat fuel gas | 1,574.4 | 673.5 |
| Option 4: Higher-temperature saturation and preheating of fuel gas | 1,272.1 | 975.8 |

TABLE 4.7 Turbine Power Output, Plant Power Use, and Net Power Output for Base Case and Four Heat Recovery Options

| Power Variable | Shift Reaction Heat Recovery Options | | | | |
|--|--------------------------------------|-----------------------|-----------------------------------|-------------------------|--|
| | Base Case | Steam Only (Option 1) | Steam and Fuel Preheat (Option 2) | Fuel Preheat (Option 3) | Higher-Temperature Fuel Preheat (Option 4) |
| Power output (MW) | | | | | |
| Gas turbine | 311.6 | 201.1 | 222.9 | 250.7 | 281.6 |
| Steam turbine | 182.2 | 202.7 | 199.6 | 195.8 | 191.4 |
| Power use (MW) | | | | | |
| CO ₂ recovery ^a | 0 | (40.2) | (40.2) | (40.2) | (40.2) |
| In-plant use ^a | (35.4) | (35.4) | (35.4) | (35.4) | (35.4) |
| Net power output (MW) | 458.4 | 328.2 | 346.9 | 370.9 | 397.4 |
| Net plant efficiency based on HHV ^b of coal (%) | 39.7 | 28.4 | 30.0 | 32.1 | 34.4 |
| Heat rates (Btu/kWh) | 8,595 | 12,004 | 11,357 | 10,622 | 9,914 |

^a Parentheses indicate power consumption.

^b HHV, Higher heating value.

The results in Table 4.7 depend on the following assumptions:

1. The efficiencies of the gas turbine, compressor, and steam turbine are the same as in the base case;
2. The inlet conditions for the gas turbine are controlled by adjusting the excess air to the combustion chamber so that the turbine peak operating temperature of 2,200°F is maintained;
3. The power consumption for in-plant power use exclusive of the CO₂ recovery system is the same for all options as in the base case; and
4. Cooling air requirement for the gas turbine is the same in all cases.

Note that in the humidified fuel gas from option 4, the sum of H₂ and CO has approximately the same mole fraction as the original synthesis gas and also has approximately the same heating value. Therefore, the humidified fuel gas from option 4 is expected to perform satisfactorily in the gas turbine.

The gas turbine, air compressor, and steam cycle efficiencies used for evaluation of the alternative CO₂ recovery systems are based on those reported by KRW for the reference case. Table 4.8 shows the stream conditions for the inlet and outlet of the air compressor. Two outlet streams are shown. The first is that reported for the base-case design. The second is that which would obtain for isentropic expansion. The ratio of the enthalpy change for the reported conditions to the enthalpy change for the isentropic conditions (1.137) is assumed to apply for variations on the compressor operation encountered in this analysis. Therefore, the enthalpy change for other compressor loadings is calculated as 1.137 times the isentropic enthalpy change, which is estimated from thermodynamic considerations.

A similar approach has been taken for estimating the performance of the gas turbine. In Table 4.8, the stream conditions for the base case are summarized. Again, the ratio of the reported enthalpy change to the isentropic enthalpy change for the given pressure ratio is used to characterize the turbine performance for differing inlet stream conditions encountered in this analysis. Several adjustments to the reported stream conditions have been made. First, the gas composition, flow rates, and heats of reaction were used to check the reported turbine inlet temperature of 2,235°F. On the basis of this heat balance, the turbine inlet temperature in the base case is actually 2,049°F, as shown in Table 4.8. The reported outlet temperature of 1,098°F includes dilution by air supplied for blade cooling. The amount of air thus supplied was calculated by a mass balance, and the cooling effect was deducted to yield the adjusted outlet temperature of 1,114°F for the base case. The assumption is made that only one-half of the air supplied for cooling does work during its expansion.

TABLE 4.8 Air Compressor and Gas Turbine Stream Conditions for Base Case

| Stream | Temperature (°F) | Pressure (psia) | ΔH^a (10^9 Btu/h) |
|-------------------------|---------------------|--------------------|---------------------------------|
| Air compressor inlet | 90 | 14.7 | 0 |
| Air compressor outlet | | | |
| Reported | 760 | 200 | 1.038 |
| Isentropic | 684 | 200 | 0.913 |
| Inlet to gas turbine | 2,049 | 194 | 0 |
| Outlet from gas turbine | | | |
| Reported | 1,114 | 14.9 | -2.121 |
| Isentropic | 772 | 14.9 | -2.821 |

^a ΔH , Change in enthalpy.

Performance of the steam cycle, including the HRSG and the steam turbine, is also expected to be similar for the base case and for the CO₂ recovery cases. In the base case, enthalpy of the gas turbine exhaust is 2.121×10^9 Btu/h. The steam turbine output is 182.2 MW. These values define an overall heat rate of 11,530 Btu/kWh. The assumption is made that this heat rate applies for the alternative plant designs employing CO₂ recovery.

5 Integrated Gasification Combined Cycle with Amine CO₂ Recovery

5.1 Feed Gas Preparation and Process Integration

As discussed in Section 4.3, the most significant issue in process integration is the effective use of process heat from cooling of the synthesis gas as required for the shift reaction and gas preparation for the CO₂ absorber. The most effective use of this thermal energy is preheating of the fuel gas feed to the gas turbine. With the addition of moisture to that feed, it can absorb much of the available heat. The balance is used for feedwater heating, raising steam for the steam cycle, and solvent regeneration. Because of differences in the absorber inlet gas specifications among the alternative CO₂ recovery systems, some variation exists in the amount of heat available. For the amine process, the CO₂-rich gas to the absorber must be cooled to 100°F. The amount and allocation of thermal energy available from various processes are listed in Table 5.1.

In the monoethanolamine (MEA) process, solvent regeneration requires the use of a reboiler, which imposes a substantial demand for thermal energy. This demand is reflected in the net power output. The reboiler demand is met by a combination of thermal energy sources,

TABLE 5.1 Heat Recovery and Allocation for MEA Process

| Process | Enthalpy Change Available from Process (10 ⁶ Btu) | Allocation to Fuel Gas Preheating (10 ⁶ Btu) | Allocation to Steam Cycle (10 ⁶ Btu) | Allocation to Solvent Regeneration (10 ⁶ Btu) |
|--|---|--|---|---|
| Initial synthesis gas cooling to 1,000°F | 1,299.3 | 369.8 | 929.5 | 0 |
| Cooling before first-stage shift | 32.9 | 0 | 0 | 32.9 |
| Cooling after first-stage shift | 68.8 | 0 | 0 | 68.8 |
| Cooling after second-stage shift | 847.0 | 580.2 | 0 | 266.8 |
| Low-pressure turbine steam extraction | 1,114.3 | 0 | 0 | 1,114.3 |

including CO₂-rich gas cooling after the second-stage shift reaction, heat of compression of the CO₂ product, and extraction of steam from the low-pressure turbine. The net plant output, including credit for this heat recovery, is listed in Table 5.2.

5.2 Carbon Dioxide Recovery by Amines

An overview of the amine-based recovery system is provided by Figure 5.1 (the system flow diagram) and by Table 5.3 (the corresponding line lists). A brief overview of the system is provided subsequently. More detailed information and assumptions for each stream are provided in Table 5.4.

The amine process uses a 30% solution of aqueous MEA as the solvent. Early applications of this solvent were restricted to a 20% solution because of the corrosiveness of the solvent. The application of corrosion inhibitors has allowed an increase in solvent concentration to 30%. The gas feed to the absorber (stream 1) has been cooled to 100°F. The rich solvent stream (stream 3) is at an elevated temperature because of the exothermic reaction of the solvent with CO₂.

TABLE 5.2 Turbine Power Output, Plant Power Use, and Net Power Output for MEA Process

| Power Variable | Power (MW) | |
|---|------------|--------|
| | Base Case | MEA |
| Power output | | |
| Gas turbine | 311.6 | 282.1 |
| Steam turbine | 182.2 | 113.4 |
| Internal power consumption ^a | | |
| CO ₂ recovery | | |
| CO ₂ compression | 0 | (34.8) |
| Solvent circulation | 0 | (4.8) |
| Solvent refrigeration | 0 | 0 |
| Others | 0 | 0 |
| Gasification system | (35.4) | (35.4) |
| Net power output | 458.4 | 320.5 |
| Energy penalty | 0 | 137.9 |

^a Parentheses indicate power consumption.

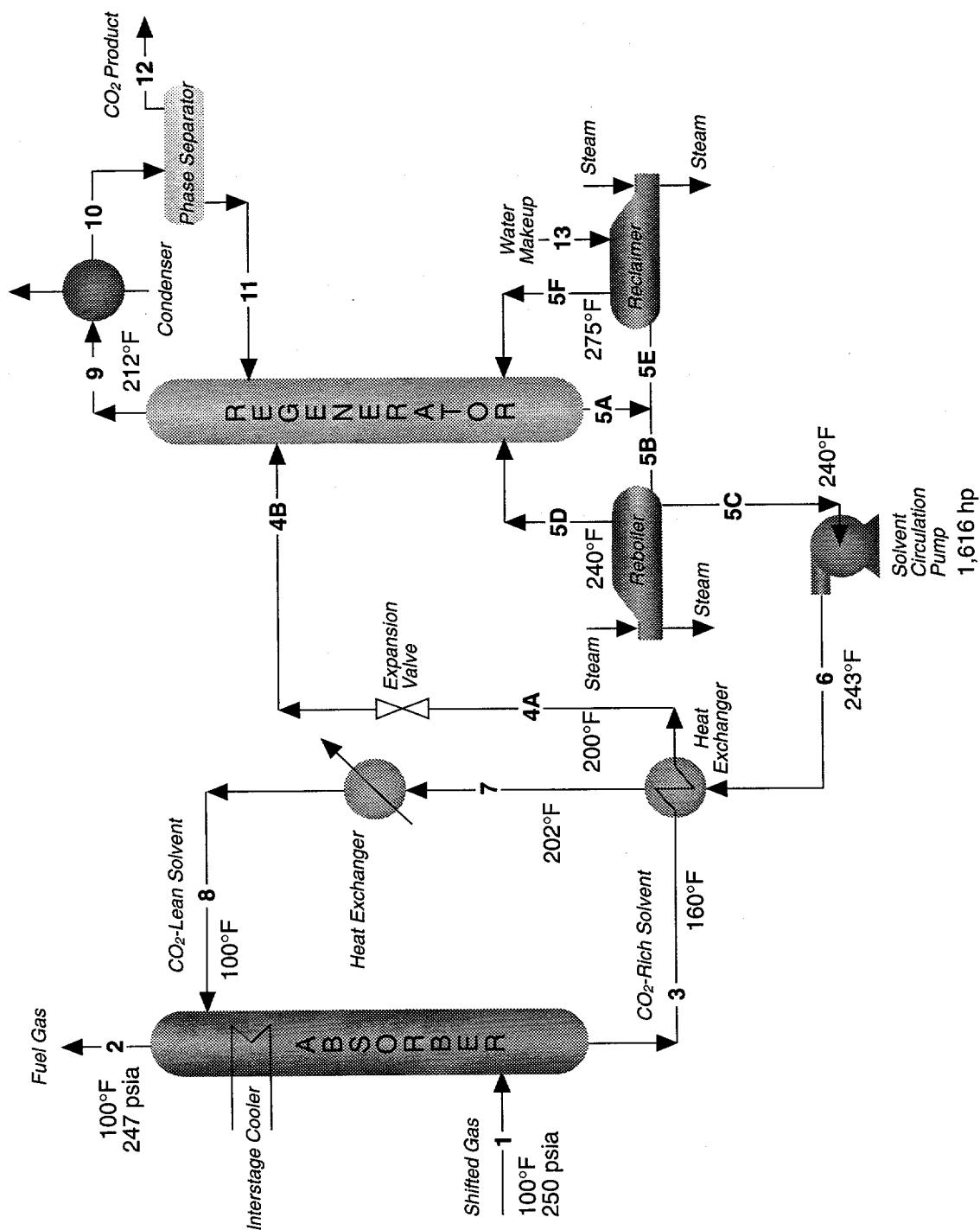


FIGURE 5.1 Summary of MEA Process (numbers refer to streams)

TABLE 5.3 Stream Flows of MEA Process^a

| Stream Data | | Stream 1 | Stream 2 | Stream 3 | Stream 4A | Stream 4B | Stream 5A |
|---|--|------------------------------|------------------------|---------------------------|-----------------------------------|---------------------------------|---------------------------|
| Description of stream | | Shifted gas absorber feed | Fuel gas | Rich MEA from absorber | Rich MEA to expansion valve | Rich MEA feed to stripper | Lean MEA from stripper |
| Gases (lb-mol/h) | | | | | | | |
| CO | | 77.37 | 77.37 | 0.00 | 0.00 | 0.00 | 0.00 |
| CO ₂ | | 4,335.99 | 216.80 | 5,448.13 | 5,448.13 | 5,448.13 | 4,201.09 |
| H ₂ | | 5,611.86 | 5,611.86 | 0.00 | 0.00 | 0.00 | 0.00 |
| H ₂ O | | 61.91 | 51.81 | 10.10 | 10.10 | 3,098.43 | 5,309.33 |
| N ₂ | | 7,306.65 | 7,306.65 | 0.00 | 0.00 | 0.00 | 0.00 |
| Ar | | 88.60 | 88.60 | 0.00 | 0.00 | 0.00 | 0.00 |
| CH ₄ | | 128.77 | 128.77 | 0.00 | 0.00 | 0.00 | 0.00 |
| NH ₃ | | 2.99 | 2.99 | 0.00 | 0.00 | 0.00 | 0.00 |
| H ₂ S | | 0.40 | 0.00 | 0.40 | 0.40 | 0.40 | 0.00 |
| HCl | | 0.04 | 0.04 | 0.00 | 0.00 | 0.00 | 0.00 |
| O ₂ | | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| COS | | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| SO ₂ | | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| Total gas flow | | 17,614.58 | 13,484.88 | 5,458.63 | 5,458.63 | 8,546.96 | 9,510.43 |
| Liquids (lb-mol/h) | | | | | | | |
| 30% MEA | | 0.00 | 0.00 | 74,080.50 | 74,080.50 | 70,992.17 | 465,725.26 |
| Temperature (°F) | | 100.00 | 100.00 | 160.00 | 200.00 | 132.08 | 220.00 |
| Pressure (psia) | | 250.00 | 247.00 | 247.00 | 245.00 | 20.00 | 20.00 |
| Enthalpy of stream (Btu/h) (reference, 32°F) | | 9.96 × 10 ⁶ | 7.26 × 10 ⁶ | 1.96 × 10 ⁸ | 2.57 × 10 ⁸ | 2.08 × 10 ⁸ | 1.78 × 10 ⁹ |

TABLE 5.3 Stream Flows of MEA Process^a (Cont.)

| Stream Data | Stream 5B | Stream 5C | Stream 5D | Stream 5E | Stream 5F | Stream 6 |
|--|------------------------|------------------------------------|-----------------------------------|------------------------|------------------------|---------------------------------|
| Description of stream | Lean MEA to reboiler | Lean MEA to absorber heat exchange | Recycle to stripper from reboiler | MEA to reclaimer | Recycle from reclaimer | Lean MEA after circulation pump |
| Gases (lb-mol/h) | | | | | | |
| CO | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| CO ₂ | 4,075.06 | 1,328.94 | 2,746.13 | 126.03 | 126.03 | 1,328.94 |
| H ₂ | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| H ₂ O | 5,150.05 | 0.00 | 5,150.05 | 159.28 | 433.90 | 0.00 |
| N ₂ | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| Ar | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| CH ₄ | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| NH ₃ | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| H ₂ S | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| HCl | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| O ₂ | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| COS | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| SO ₂ | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| Total gas flow | 9,225.11 | 1,328.94 | 7,896.18 | 285.31 | 559.93 | 1,328.94 |
| Liquids (lb-mol/h) | | | | | | |
| 30% MEA | 451,753.50 | 74,080.50 | 377,673.00 | 13,971.76 | 13,971.76 | 74,080.50 |
| Temperature (°F) | 220.00 | 240.00 | 240.00 | 220.00 | 275.00 | 242.75 |
| Pressure (psia) | 20.00 | 20.00 | 20.00 | 20.00 | 20.00 | 250.00 |
| Enthalpy of stream (Btu/h) (reference, 32°F) | 1.72 × 10 ⁹ | 3.11 × 10 ⁸ | 1.68 × 10 ⁹ | 5.33 × 10 ⁷ | 7.74 × 10 ⁷ | 3.15 × 10 ⁸ |

TABLE 5.3 Stream Flows of MEA Process^a (Cont.)

| Stream Data | Stream 7 | Stream 8 | Stream 9 | Stream 10 | Stream 11 | Stream 12 |
|--|-------------------------------|------------------------|---|---|------------------------|-------------------------------|
| Description of stream | Lean MEA after heat exchanger | Lean MEA after cooler | CO ₂ -rich gas from stripper | CO ₂ -rich gas after condenser | Recycle to stripper | CO ₂ -rich product |
| Gases (lb-mol/h) | | | | | | |
| CO | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| CO ₂ | 1,328.94 | 1,328.94 | 4,119.19 | 4,119.19 | 0.00 | 4,119.19 |
| H ₂ | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| H ₂ O | 0.00 | 0.00 | 8,238.38 | 8,238.38 | 7,953.66 | 284.72 |
| N ₂ | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| Ar | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| CH ₄ | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| NH ₃ | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| H ₂ S | 0.00 | 0.00 | 0.40 | 0.40 | 0.00 | 0.40 |
| HCl | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| O ₂ | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| COS | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| SO ₂ | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| Total gas flow | 1,328.94 | 1,328.94 | 12,357.97 | 12,357.97 | 7,953.66 | 4,404.32 |
| Liquids (lb-mol/h) | | | | | | |
| 30% MEA | 74,080.50 | 74,080.50 | 32.60 ^b | 32.60 ^b | 32.60 ^b | 0.29 ^b |
| Temperature (°F) | 201.71 | 100.00 | 212.00 | 100.00 | 100.00 | 100.00 |
| Pressure (psia) | 250.00 | 250.00 | 14.70 | 14.70 | 14.70 | 14.70 |
| Enthalpy of stream (Btu/h) (reference, 32°F) | 2.54 × 10 ⁸ | 1.02 × 10 ⁸ | 1.78 × 10 ⁸ | 1.27 × 10 ⁷ | 9.78 × 10 ⁶ | 2.49 × 10 ⁶ |

^a Flow rates are for one of four trains.^b MEA only.

TABLE 5.4 Descriptions of Streams of MEA Process^a

| Stream and Characteristic | Data | Comments on Stream Calculations |
|---|--------|---|
| Stream 1: Gas feed to absorber | | |
| Temperature (°F) | 100 | In MEA process, mean absorption temperature is 77°F. Therefore, feed gas is cooled to 100°F to limit solvent flow rate. Absorption pressure is 250 psia, which is exit pressure at shift reactor. Water content corresponds to vapor pressure of water at 100°F. |
| Pressure (psia) | 250 | |
| Flow rate (lb-mol/h) | 17,614 | |
| CO ₂ (mole fraction) | 0.2462 | |
| Stream 2: Fuel gas | | |
| Temperature (°F) | 100 | Composition of this stream corresponds to 95% removal of CO ₂ from synthesis gas and water content corresponding to vapor pressure of water at prevailing temperature of 100°F. Gases leave absorber at temperature equal to that of solvent entering absorber at top. Because of small quantities of H ₂ S present, assumption is made that all H ₂ S is absorbed. Other gases are not absorbed and leave absorber along with fuel gas. |
| Pressure (psia) | 247 | |
| Flow rate (lb-mol/h) | 13,485 | |
| CO ₂ (mole fraction) | 0.0161 | |
| Stream 3: Rich MEA solution | | |
| Temperature (°F) | 160 | Flow rate of MEA is calculated on basis of CO ₂ concentration in feed and fuel gas and in lean MEA. Assumption is made that lean MEA solution will have CO ₂ at 0.16 mol/mol of MEA. Temperature of rich MEA solution increases because of heat of reaction (825 Btu/lb of CO ₂ absorbed) of CO ₂ and MEA. Temperature is estimated from this heat of reaction and from enthalpies of streams around absorber. Concentration of CO ₂ in rich MEA is 0.656 mol/mol of MEA. Interstage cooling is used to limit the temperature rise to 160°F. |
| Pressure (psia) | 247 | |
| Flow rate (lb-mol/h) | 79,539 | |
| CO ₂ (mole fraction) | 0.0685 | |
| Stream 4A: Rich MEA feed to expansion valve | | |
| Temperature (°F) | 200 | Rich MEA solution is heated from 160°F to 200°F in lean-rich solvent heat exchanger to decrease reboiler load. |
| Pressure (psia) | 245 | |
| Flow rate (lb-mol/h) | 79,539 | |
| CO ₂ (mole fraction) | 0.0685 | |

TABLE 5.4 (Cont.)

| Stream and Characteristic | Data | Comments on Stream Calculations |
|---|--------|---|
| Stream 4B: Rich MEA feed to stripper | | |
| Temperature (°F) | 132 | Rich MEA is flashed from 245 to 20 psia before stripper to release dissolved CO ₂ . Along with CO ₂ , water will also be flashed. Amount of CO ₂ released is assumed to be one-third of net CO ₂ absorbed. From partial pressures of water and CO ₂ at prevailing conditions, amount of water flashed is calculated. From heat of vaporization of water and heat of desorption of CO ₂ , temperature of solution is calculated. |
| Pressure (psia) | 20 | |
| Flow rate (lb-mol/h) | 79,539 | |
| CO ₂ (mole fraction) | 0.0685 | |
| Stream 5C: Lean MEA from reboiler | | |
| Temperature (°F) | 240 | Carbon dioxide level in lean MEA is 0.16 mol/mol of MEA. Reboiler is operated at temperature of 240°F. This temperature ensures desired loading of CO ₂ in lean MEA. |
| Pressure (psia) | 20 | |
| Flow rate (lb-mol/h) | 75,409 | |
| CO ₂ (mole fraction) | 0.0176 | |
| Stream 6: Lean MEA after circulation pump | | |
| Temperature (°F) | 243 | Lean MEA is at pressure of 20 psia and is pressurized to absorber pressure of 250 psia by using circulation pump of 1,616 hp (efficiency, 75%), increasing temperature of solvent to 243°F. |
| Pressure (psia) | 250 | |
| Flow rate (lb-mol/h) | 75,409 | |
| CO ₂ (mole fraction) | 0.0176 | |
| Stream 7: Lean MEA after lean-rich solvent heat exchanger | | |
| Temperature (°F) | 202 | Lean solvent is cooled against rich MEA from absorber to temperature of 202°F in lean-rich solvent heat exchanger. |
| Pressure (psia) | 250 | |
| Flow rate (lb-mol/h) | 75,409 | |
| CO ₂ (mole fraction) | 0.0176 | |
| Stream 8: Lean MEA after lean solvent cooler | | |
| Temperature (°F) | 100 | Lean MEA is further cooled to temperature of 100°F in heat exchanger by using cooling water. Cooling water is assumed to be available at 70°F. |
| Pressure (psia) | 250 | |
| Flow rate (lb-mol/h) | 75,409 | |
| CO ₂ (mole fraction) | 0.0176 | |

TABLE 5.4 (Cont.)

| Stream and Characteristic | Data | Comments on Stream Calculations |
|--|--------|--|
| Stream 9: CO ₂ -rich gas from stripper | | |
| Temperature (°F) | 212 | Remaining CO ₂ in solvent after flashing is released by application of heat. To maintain low partial pressure of CO ₂ at top of stripper, reflux ratio of 2 is used. Therefore, additional water has to be evaporated from stripper along with CO ₂ . Composition of this stream represents amounts of CO ₂ and water flashed and remaining CO ₂ and water required for reflux ratio of 2. Temperature at top of stripper is 208°F. |
| Pressure (psia) | 14.7 | |
| Flow rate (lb-mol/h) | 12,391 | |
| CO ₂ (mole fraction) | 0.3324 | |
| Stream 10: CO ₂ -rich gas after condenser | | |
| Temperature (°F) | 100 | Gas rich in CO ₂ , along with water vapor and small amounts of MEA, is condensed in heat exchanger by using cooling water to temperature of 100°F. |
| Pressure (psia) | 14.7 | |
| Flow rate (lb-mol/h) | 12,391 | |
| CO ₂ (mole fraction) | 0.3324 | |
| Stream 11: Recycle to stripper | | |
| Temperature (°F) | 100 | Condensed water and MEA are recycled to top of stripper. |
| Pressure (psia) | 14.7 | |
| Flow rate (lb-mol/h) | 7,986 | |
| CO ₂ (mole fraction) | 0 | |
| Stream 12: CO ₂ product stream | | |
| Temperature (°F) | 100 | This stream is final CO ₂ -rich gas for further compression and disposal. Water vapor in this stream corresponds to vapor pressure of water at 100°F. |
| Pressure (psia) | 14.7 | |
| Flow rate (lb-mol/h) | 4,405 | |
| CO ₂ (mole fraction) | 0.9352 | |
| Stream 13: Water makeup | | |
| Temperature (°F) | 100 | More water is removed through fuel gas and final CO ₂ product than is supplied with feed gas. Therefore, water has to be added to system to maintain water content in MEA solution. |
| Pressure (psia) | 14.7 | |
| Flow rate (lb-mol/h) | 275 | |
| CO ₂ (mole fraction) | 0 | |

^a Flow rates are for one of four trains.

Solvent regeneration is accomplished through pressure reduction and heating. A portion of the heat required is obtained through exchange of heat between the rich solvent (stream 3) and the lean solvent (stream 6). The resulting stream is delivered to the regenerator through an expansion valve. Most of the heat of regeneration is obtained from steam supplied to a recirculating reboiler. The sources of steam or heat for steam supply are listed in Table 5.1. The cooled solvent enters the absorber at 100°F. (The operating temperature of the absorption column is maintained at 160°F by interstage cooling by circulating cooling water.) The regeneration process releases the CO₂, which is cooled to condense accompanying solvent vapors and to prepare it for compression to pipeline disposal conditions.

5.3 Power System Performance Effects of the MEA Process

The fact that the synthesis gas must be cooled before processing by the CO₂ recovery system significantly affects the net power system output. This cooling substantially reduces the enthalpy of the synthesis gas, including the enthalpy contributed by the exothermic shift reaction. As discussed in Section 4.3, much of the thermal energy removed from the gas stream can be recovered, and the most effective use of that energy is preheating of humidified fuel gas. This effectiveness is shown in Table 4.7, where several alternative process integration strategies are compared. The heat demand for solvent regeneration (stripping) in the MEA process cannot be met entirely by heat recovered during cooling of the synthesis gas. In fact, most of the heat for stripping is obtained by diverting steam from the low-pressure turbine, as listed in Table 5.1. The overall power balance for the plant is summarized in Table 5.2. The energy penalty associated with CO₂ recovery includes both reduced output from the turbogenerator units and electricity consumption for pumping, compression, and refrigeration in the CO₂ recovery system. For the MEA process, the total penalty is 137.9 MW.

5.4 Major Equipment Specification and Preliminary Cost Estimation

5.4.1 Heat Exchanger before First-Stage Shift Reactor

The temperature of the gases from the KRW process following the initial cooling from 1,600°F in a heat recovery unit is 1,000°F. These gases have to be cooled to a temperature of 482°F, the optimum temperature for the first-stage shift reactor. The difference between the enthalpies of gases at these two temperatures gives the heat exchanger load. Most of this heat is used for raising saturated steam from saturated water at 417.4°F and 300 psia. This steam is used in the first-stage shift reactor. The remaining heat is used in the reboiler for the stripper. The heat transfer area required can be calculated from

$$Q = U \cdot A \cdot \Delta T , \quad (5.1)$$

where

Q is the amount of heat transferred (Btu/h),

U is the overall heat transfer coefficient (Btu/h/ft²/°F),

A is the heat transfer area (ft²), and

ΔT is the logarithmic mean temperature difference, defined as

$$\Delta T = \frac{\Delta T_1 - \Delta T_2}{\ln(\Delta T_1/\Delta T_2)}, \quad (5.2)$$

where

ΔT_1 is the temperature of inlet hot fluid minus the temperature of outlet cold fluid, and

ΔT_2 is the temperature of outlet hot fluid minus the temperature of inlet cold fluid (assuming countercurrent flow).

The area required for the two operations (raising steam and reboiling) is estimated separately because the approach temperature and the overall heat transfer coefficients are different. For raising steam from saturated water, a boiling film coefficient of 1,000 Btu/h/ft²/°F is used (Perry and Chilton 1973, p. 10-45, Table 10-10). The overall heat transfer coefficient for the reboiler operation is a gas-liquid heat transfer coefficient and is assumed to be 80 Btu/h/ft²/°F. The detailed calculations are shown in Table 5.5.

5.4.2 Heat Exchanger between First and Second Shift Reactors

The shift reaction is an exothermic reaction, and the heat released in the first stage has to be removed to obtain an optimum temperature for the second-stage shift reactor. The gases have to be cooled from 680°F to 460°F. The heat load of the heat exchanger is again the difference between the enthalpies of the gases. A major portion of the available heat is used in heating water from 100°F to saturated water conditions at 300 psia and 417.4°F, and the remaining heat is used in the reboiler. The surface area is calculated by using Equation 5.1, and the details are given in Table 5.5.

TABLE 5.5 Sizing and Cost Estimation of Major Equipment for MEA Process

1. Heat Exchanger before First Stage for Raising Steam

| | | |
|--|------------|----------|
| Q = Load (Btu/h) | 78,838,137 | |
| Tha = Inlet temperature of hot fluid (°F) | 1,000 | |
| Thb = Outlet temperature of hot fluid (°F) | 531 | |
| Pressure of hot gases (psia) | 290 | |
| Tca = Inlet temperature of cold fluid (saturated water) (°F) | 417 | |
| Tcb = Outlet temperature of cold fluid (saturated vapor) (°F) | 417 | |
| ΔT_1 | 583 | |
| ΔT_2 | 114 | |
| Log mean temperature difference (°F) | 287 | |
| Overall heat transfer coefficient (Btu/h/ft ² /°F) | | |
| (overall heat transfer coefficient for boiling) ^a | 260 | |
| Heat transfer area (ft ²) | 1,057 | |
| Operating pressure (psia) | 300 | |
| Pressure factor | 1.165 | |
| Module factor | 3.2 | |
| (includes all supporting equipment and connections and installation) | | |
| Purchased cost of heat exchanger in 1987 | \$23,000 | |
| (mild steel construction; shell and tube floating head) | | |
| CE index for process equipment in 1987 ^b | 320 | |
| CE index for process equipment in 1993 | 360.4 | |
| Installed cost of heat exchanger in 1993 | | \$96,569 |

2. Gas-Liquid Heat Exchanger before First Stage for Reboiler

| | | |
|--|-----------|----------|
| Q = Load (Btu/h) | 8,219,649 | |
| Tha = Inlet temperature of hot fluid (°F) | 531 | |
| Thb = Outlet temperature of hot fluid (°F) | 482 | |
| Pressure of hot gases (psia) | 285 | |
| Tca = Inlet temperature of cold fluid (MEA) (°F) | 220 | |
| Tcb = Outlet temperature of cold fluid (MEA) (°F) | 240 | |
| ΔT_1 | 291 | |
| ΔT_2 | 262 | |
| Log mean temperature difference (°F) | 276 | |
| Overall heat transfer coefficient (Btu/h/ft ² /°F) | 40 | |
| (hot fluid: gas, cold fluid: MEA) ^c | | |
| Heat transfer area (ft ²) | 744 | |
| Operating pressure (psia) | 285 | |
| Pressure factor | 1.165 | |
| Module factor | 3.2 | |
| (includes all supporting equipment and connections and installation) | | |
| Purchased cost of heat exchanger in 1987 | \$19,000 | |
| (mild steel construction; shell and tube floating head) | | |
| CE index for process equipment in 1987 | 320 | |
| CE index for process equipment in 1993 | 360.4 | |
| Installed cost of heat exchanger in 1993 | | \$79,775 |

TABLE 5.5 (Cont.)

3. Heat Exchanger after First Stage for Saturated Water

| | | |
|--|------------|-----------|
| Q = Load (Btu/h) | 31,747,489 | |
| Tha = Inlet temperature of hot fluid (°F) | 680 | |
| Thb = Outlet temperature of hot fluid (°F) | 537 | |
| Pressure of hot gases (psia) | 275 | |
| Tca = Inlet temperature of water (°F) | 100 | |
| Tcb = Outlet temperature of saturated water (°F) | 417 | |
| Saturated water pressure (psia) | 300 | |
| ΔT_1 | 263 | |
| ΔT_2 | 437 | |
| Log mean temperature difference (°F) | 342 | |
| Overall heat transfer coefficient (Btu/h/ft ² /°F) | 40 | |
| (hot fluid: gas, cold fluid: water) ^c | | |
| Heat transfer area (ft ²) | 2,319 | |
| Operating pressure (psia) | 275 | |
| Pressure factor | 1.165 | |
| Module factor | 3.2 | |
| (includes all supporting equipment and connections and installation) | | |
| Purchased cost of heat exchanger in 1987 | \$40,000 | |
| (mild steel construction; shell and tube floating head) | | |
| CE index for process equipment in 1987 | 320 | |
| CE index for process equipment in 1993 | 360.4 | |
| Installed cost of heat exchanger in 1993 | | \$167,946 |

4. Gas-Liquid Heat Exchanger after First Stage for Reboiler

| | | |
|--|------------|-----------|
| Q = Load (Btu/h) | 17,195,044 | |
| Tha = Inlet temperature of hot fluid (°F) | 537 | |
| Thb = Outlet temperature of hot fluid (°F) | 460 | |
| Pressure of hot gases (psia) | 285 | |
| Tca = Inlet temperature of cold fluid (MEA) (°F) | 220 | |
| Tcb = Outlet temperature of cold fluid (MEA) (°F) | 240 | |
| ΔT_1 | 297 | |
| ΔT_2 | 240 | |
| Log mean temperature difference (°F) | 267 | |
| Overall heat transfer coefficient (Btu/h/ft ² /°F) | 40 | |
| (hot fluid: gas, cold fluid: liquid) ^c | | |
| Heat transfer area (ft ²) | 1,608 | |
| Operating pressure (psia) | 285 | |
| Pressure factor | 1.165 | |
| Module factor | 3.2 | |
| (includes all supporting equipment and connections and installation) | | |
| Purchased cost of heat exchanger in 1987 | \$32,000 | |
| (mild steel construction; shell and tube floating head) | | |
| CE index for process equipment in 1987 | 320 | |
| CE index for process equipment in 1993 | 360.4 | |
| Installed cost of heat exchanger in 1993 | | \$134,357 |

TABLE 5.5 (Cont.)

5. Gas-Liquid Heat Exchanger after Second Stage for Reboiler

| | | |
|--|------------|------------------|
| Q = Load (Btu/h) | 66,700,501 | |
| Tha = Inlet temperature of hot fluid (°F) | 480 | |
| Thb = Outlet temperature of hot fluid (°F) | 417 | |
| Pressure of hot gases (psia) | 285 | |
| Tca = Inlet temperature of cold fluid (MEA) (°F) | 220 | |
| Tcb = Outlet temperature of cold fluid (MEA) (°F) | 240 | |
| ΔT_1 | 240 | |
| ΔT_2 | 197 | |
| Log mean temperature difference (°F) | 218 | |
| Overall heat transfer coefficient (Btu/h/ft ² /°F) | 40 | |
| (hot fluid: gas, cold fluid: liquid) ^c | | |
| Heat transfer area (ft ²) | 7,649 | |
| Operating pressure (psia) | 285 | |
| Pressure factor | 1.165 | |
| Module factor | 3.2 | |
| (includes all supporting equipment and connections and installation) | | |
| Purchased cost of heat exchanger in 1987 | \$90,000 | |
| (mild steel construction; shell and tube floating head) | | |
| CE index for process equipment in 1987 | 320 | |
| CE index for process equipment in 1993 | 360.4 | |
| Installed cost of heat exchanger in 1993 | | \$377,879 |

6. Gas-Gas Heat Exchanger after Second Stage for Fuel Gas Heating

| | | |
|--|-------------|--------------------|
| Q = Load (Btu/h) | 145,047,122 | |
| Tha = Inlet temperature of hot fluid (°F) | 417 | |
| Thb = Outlet temperature of hot fluid (°F) | 120 | |
| Pressure of hot gases (psia) | 285 | |
| Tca = Inlet temperature of cold fluid (fuel gas) (°F) | 100 | |
| Tcb = Outlet temperature of cold fluid (fuel gas) (°F) | 401 | |
| ΔT_1 | 16 | |
| ΔT_2 | 20 | |
| Log mean temperature difference (°F) | 18 | |
| Overall heat transfer coefficient (Btu/h/ft ² /°F) ^d | 5 | |
| Heat transfer area (ft ²) | 1,599,152 | |
| Operating pressure (psia) | 285 | |
| Pressure factor | 1.165 | |
| Module factor | 3.2 | |
| (includes all supporting equipment and connections and installation) | | |
| Purchased cost of heat exchanger in 1987 | \$419,706 | |
| (mild steel construction; shell and tube floating head) | | |
| CE index for process equipment in 1987 | 320 | |
| CE index for process equipment in 1993 | 360.4 | |
| Installed cost of heat exchanger in 1993 | | \$1,762,202 |

TABLE 5.5 (Cont.)

| | | |
|--|-----------|-----------|
| 7. Gas-Liquid Heat Exchanger after Second Stage | | |
| Q = Load (Btu/h) | 4,415,036 | |
| Tha = Inlet temperature of hot fluid (°F) | 120 | |
| Thb = Outlet temperature of hot fluid (°F) | 100 | |
| Pressure of hot gases (psia) | 285 | |
| Tca = Inlet temperature of cold fluid (water) (°F) | 70 | |
| Tcb = Outlet temperature of cold fluid (water) (°F) | 100 | |
| ΔT_1 | 20 | |
| ΔT_2 | 30 | |
| Log mean temperature difference (°F) | 25 | |
| Overall heat transfer coefficient (Btu/h/ft ² /°F) | 40 | |
| (hot fluid: gas, cold fluid: liquid) ^c | | |
| Heat transfer area (ft ²) | 4,475 | |
| Operating pressure (psia) | 285 | |
| Pressure factor | 1.165 | |
| Module factor | 3.2 | |
| (includes all supporting equipment and connections and installation) | | |
| Purchased cost of heat exchanger in 1987 | \$60,000 | |
| (mild steel construction; shell and tube floating head) | | |
| CE index for process equipment in 1987 | 320 | |
| CE index for process equipment in 1993 | 360.4 | |
| Installed cost of heat exchanger in 1993 | | \$251,920 |
| 8. First-Stage Shift Reactor | | |
| Catalyst volume (ft ³) | 1,000 | |
| Reactor volume (ft ³) (1.2 times catalyst volume) | 1,200 | |
| Reactor volume (gal) | 8,963 | |
| Pressure factor | 1 | |
| Module factor | 3.05 | |
| CE index for process equipment in 1987 | 320 | |
| CE index for process equipment in 1993 | 360.4 | |
| Purchased cost of reactor in 1987 | \$9,000 | |
| Installed cost of reactor in 1993 | | \$30,916 |
| 9. Second-Stage Shift Reactor | | |
| Catalyst volume (ft ³) | 340 | |
| Reactor volume (ft ³) (1.2 times catalyst volume) | 408 | |
| Reactor volume (gal) | 3,041 | |
| Pressure factor | 1 | |
| Module factor | 3.05 | |
| CE index for process equipment in 1987 | 320 | |
| CE index for process equipment in 1993 | 360.4 | |
| Purchased cost of reactor in 1987 | \$5,000 | |
| Installed cost of reactor in 1993 | | \$17,175 |

TABLE 5.5 (Cont.)

10. Cost of Shift Catalyst

| | | |
|---|-------|------------------|
| Volume of catalyst in first stage (ft ³) | 1,000 | |
| Volume of catalyst in second stage (ft ³) | 340 | |
| Cost of high-temperature catalyst (\$/ft ³) | \$50 | |
| Cost of low-temperature catalyst (\$/ft ³) | \$250 | |
| Total cost of catalyst | | \$135,000 |

11. CO₂ Absorption Column

| | | |
|---|---------|------------------|
| Diameter of tower (ft) | 12 | |
| HETP (ft) | 3 | |
| No. of theoretical stages | 12 | |
| Absorber tower height (ft) | 40 | |
| (4 ft for inlet and outlet and for gas and liquid distributors) | | |
| Volume of packing (ft ³) | 4,073 | |
| Pressure factor | 2 | |
| Cost per foot of column height (\$/ft) | \$1,400 | |
| (mild steel construction) | | |
| Module factor | 4.16 | |
| CE index for process equipment in 1987 | 320 | |
| CE index for process equipment in 1993 | 360.4 | |
| 1993 installed cost of absorber | | \$524,742 |
| Cost of packing (\$/ft ³) | \$63.5 | |
| (2-in. pall rings-metal) | | |
| Total cost of packing | | \$258,645 |

12. CO₂ Stripping Column

| | | |
|---|---------|------------------|
| Diameter of tower (ft) | 15 | |
| HETP (ft) | 3 | |
| No. of theoretical stages | 15 | |
| Absorber tower height (ft) | 49 | |
| (4 ft for inlet and outlet and for gas and liquid distributors) | | |
| Volume of packing (ft ³) | 7,955 | |
| Pressure factor | 2 | |
| Cost per foot of column height (\$/ft) | \$1,650 | |
| (mild steel construction) | | |
| Module factor | 4.16 | |
| CE index for process equipment in 1987 | 320 | |
| CE index for process equipment in 1993 | 360.4 | |
| 1993 installed cost of absorber | | \$757,597 |
| Cost of packing (\$/ft ³) | \$63.5 | |
| (2-in. pall rings-metal) | | |
| Total cost of packing | | \$505,165 |

TABLE 5.5 (Cont.)

13. Heat Exchanger for Lean Solvent Cooling

| | | |
|--|-------------|------------------|
| Q = Load (Btu/h) | 151,957,348 | |
| Tha = Inlet temperature of hot fluid (°F) | 202 | |
| Thb = Outlet temperature of hot fluid (°F) | 100 | |
| Pressure of hot gases (psia) | 250 | |
| Tca = Inlet water temperature (°F) | 70 | |
| Tcb = Outlet water temperature (°F) | 150 | |
| ΔT_1 | 52 | |
| ΔT_2 | 30 | |
| Log mean temperature difference (°F) | 40 | |
| Overall heat transfer coefficient (Btu/h/ft ² /°F) | 150 | |
| (hot fluid: MEA, cold fluid: water) ^c | | |
| Heat transfer area (ft ²) | 25,408 | |
| Operating pressure (psia) | 250 | |
| Pressure factor | 1.165 | |
| Module factor | 3.2 | |
| (includes all supporting equipment and connections and installation) | | |
| Purchased cost of heat exchanger in 1987 | \$207,385 | |
| (mild steel construction; shell and tube floating head) | | |
| CE index for process equipment in 1987 | 320 | |
| CE index for process equipment in 1993 | 360.4 | |
| Installed cost of heat exchanger in 1993 | | \$870,740 |

14. Interstage Cooler for Absorber

| | | |
|--|------------|------------------|
| Q = Load (Btu/h) | 57,747,672 | |
| Tha = Inlet temperature of hot fluid (°F) | 198 | |
| Thb = Outlet temperature of hot fluid (°F) | 160 | |
| Pressure of hot gases (psia) | 250 | |
| Tca = Inlet water temperature (°F) | 70 | |
| Tcb = Outlet water temperature (°F) | 160 | |
| ΔT_1 | 38 | |
| ΔT_2 | 90 | |
| Log mean temperature difference (°F) | 60 | |
| Overall heat transfer coefficient (Btu/h/ft ² /°F) | 150 | |
| (hot fluid: MEA, cold fluid: water) ^c | | |
| Heat transfer area (ft ²) | 6,384 | |
| Operating pressure (psia) | 250 | |
| Pressure factor | 1 | |
| Module factor | 3.2 | |
| (includes all supporting equipment and connections and installation) | | |
| Purchased cost of heat exchanger in 1987 | \$80,000 | |
| (mild steel construction; shell and tube floating head) | | |
| CE index for process equipment in 1987 | 320 | |
| CE index for process equipment in 1993 | 360.4 | |
| Installed cost of heat exchanger in 1993 | | \$288,320 |

TABLE 5.5 (Cont.)

15. Lean-Rich MEA Heat Exchanger

| | | |
|---|------------|------------------|
| Q = Load (Btu/h) | 61,345,644 | |
| Tha = Inlet temperature of hot fluid (°F) (lean MEA) | 243 | |
| Thb = Outlet temperature of hot fluid (°F) (lean MEA) | 202 | |
| Tca = Inlet temperature of cold fluid (°F) (rich MEA) | 160 | |
| Tcb = Outlet temperature of cold fluid (°F) (rich MEA) | 200 | |
| ΔT_1 | 43 | |
| ΔT_2 | 42 | |
| Log mean temperature difference (°F) | 42 | |
| Overall heat transfer coefficient (Btu/h/ft ² /°F) (hot fluid: gas, cold fluid: water) ^c | 150 | |
| Heat transfer area (ft ²) | 9,658 | |
| Operating pressure (psia) | 250 | |
| Pressure factor | 1 | |
| Module factor | 3.2 | |
| (includes all supporting equipment and connections and installation) | | |
| Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head) | \$110,000 | |
| CE index for process equipment in 1987 | 320 | |
| CE index for process equipment in 1993 | 360.4 | |
| Installed cost of heat exchanger in 1993 | | \$396,440 |

16. Stripper Overhead Condenser

| | | |
|---|------------|------------------|
| Q = Load (Btu/h) | 11,803,945 | |
| Tha = Inlet temperature of hot fluid (°F) | 212 | |
| Thb = Outlet temperature of hot fluid (°F) | 100 | |
| Pressure of hot gases (psia) | 14.7 | |
| Tca = Inlet water temperature (°F) | 70 | |
| Tcb = Outlet water temperature (°F) | 165 | |
| ΔT_1 | 43 | |
| ΔT_2 | 30 | |
| Log mean temperature difference (°F) | 36 | |
| Overall heat transfer coefficient (Btu/h/ft ² /°F) (hot fluid: gas, cold fluid: water) ^c | 40 | |
| Heat transfer area (ft ²) | 7,793 | |
| Operating pressure (psia) | 14.7 | |
| Pressure factor | 1 | |
| Module factor | 3.2 | |
| (includes all supporting equipment and connections and installation) | | |
| Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head) | \$90,000 | |
| CE index for process equipment in 1987 | 320 | |
| CE index for process equipment in 1993 | 360.4 | |
| Installed cost of heat exchanger in 1993 | | \$324,360 |

TABLE 5.5 (Cont.)

17. Stripper Reboiler

| | | |
|--|-------------|-----------|
| Q = Load (Btu/h) | 278,576,337 | |
| Tha = Inlet temperature of hot fluid (°F) | 501 | |
| Thb = Outlet temperature of hot fluid (°F) | 267 | |
| Pressure of hot gases (psia) | 50 | |
| Tca = Inlet temperature of MEA liquid (°F) | 220 | |
| Tcb = Outlet temperature of MEA liquid (°F) | 240 | |
| ΔT_1 | 261 | |
| ΔT_2 | 47 | |
| Log mean temperature difference (°F) | 125 | |
| Overall heat transfer coefficient (Btu/h/ft ² /°F) ^e | 110 | |
| Heat transfer area (ft ²) | 20,288 | |
| Operating pressure (psia) | 300 | |
| Pressure factor | 1.165 | |
| Module factor | 3.2 | |
| (includes all supporting equipment and connections and installation) | | |
| Purchased cost of heat exchanger in 1987 | \$177,958 | |
| (mild steel construction; shell and tube floating head) | | |
| CE index for process equipment in 1987 | 320 | |
| CE index for process equipment in 1993 | 360.4 | |
| Installed cost of heat exchanger in 1993 | | \$747,186 |

18. Solvent Circulation Pump

| | | |
|---|----------|-----------|
| Horsepower | 1,616 | |
| Size exponent | 0.79 | |
| Purchased cost of 300-hp pump in 1987 | \$30,000 | |
| (includes motor, coupling, base; cast iron, horizontal) | | |
| Module factor | 1.5 | |
| CE index for process equipment in 1987 | 320 | |
| CE index for process equipment in 1993 | 360.4 | |
| Installed cost of solvent pump in 1993 | | \$191,700 |

19. CO₂ Product Gas Compressors

| | | |
|--|-----------|--|
| Compressor 1 (hp) | 4,051 | |
| Compressor 2 (hp) | 3,804 | |
| Compressor 3 (hp) | 3,803 | |
| Purchased cost of centrifugal compressor 1 in 1987 | \$800,000 | |
| Purchased cost of centrifugal compressor 2 in 1987 | \$750,000 | |
| Purchased cost of centrifugal compressor 3 in 1987 | \$750,000 | |
| (includes electric motor drive and gear reducer) | | |
| Size factor for compressor | 1 | |

TABLE 5.5 (Cont.)

| | | |
|--|-------|---------------------|
| 19. (Cont.) | | |
| Module factor ^f | 2.6 | |
| CE index for process equipment in 1987 | 320 | |
| CE index for process equipment in 1993 | 360.4 | |
| Installed cost of compressor 1 in 1993 | | \$2,342,600 |
| Installed cost of compressor 2 in 1993 | | \$2,196,188 |
| Installed cost of compressor 3 in 1993 | | \$2,196,188 |
| Total Cost | | \$14,653,610 |
| Total Cost of Four Trains | | \$58,614,400 |

^a Source: Fraas (1982, Figure 4.5).

^b CE, Chemical engineering.

^c Source: Perry and Chilton (1973, p. 10-44, Table 10-10).

^d Source: B&W (1975, p. 19-10).

^e Source: Anada et al. (1982, p. 116).

^f Source: Garrett (1989).

5.4.3 Heat Exchanger after Second-Stage Shift Reactor

The shifted gases from the second stage are at a temperature of 480°F. These gases have to be cooled to a temperature of 100°F. This cooling is achieved in three stages. In the first stage, the gases are cooled to about 417.4°F in the reboiler; in the second stage, the gases are cooled to about 120°F by heating CO₂-lean fuel gas from the absorber at 100°F to a temperature of 401°F. The final cooling to 100°F is achieved by circulating cooling water. The heat exchanger areas required for these operations are estimated, and the details are given in Table 5.5.

5.4.4 First-Stage Shift Reactor

The volume of the reactor is assumed to be 20% greater than the volume of the catalyst. The volume of the catalyst is estimated to be 1,000 ft³ from the kinetics and the equilibrium considerations.

5.4.5 Second-Stage Shift Reactor

The volume of the reactor is assumed to be 20% greater than the volume of the catalyst. The volume of the catalyst is estimated to be 340 ft^3 from the kinetics and the equilibrium considerations.

5.4.6 Cost of Catalyst

The costs of the catalysts are assumed to be $\$50/\text{ft}^3$ and $\$250/\text{ft}^3$ for the iron- and copper-based catalysts, respectively.

5.4.7 Absorption Tower

The diameter of the absorption tower is calculated for 60% flooding velocity. Diameters are calculated at the top and bottom of the tower, and the larger of the two is taken as the tower diameter. The diameter of the tower is 12 ft. The height of the tower is calculated from the number of theoretical stages required and the height equivalent to a theoretical plate (HETP). The assumptions are made that 12 theoretical stages are required and that HETP is 3 ft. An additional 4 ft is provided for liquid and gas distributors.

5.4.8 Stripping Column

The diameter is again calculated for 60% flooding velocity. The diameter of the tower is 15 ft. The number of theoretical stages required is 15, and the HETP is 3 ft. Therefore, the total height of the column is 49 ft, including 4 ft for gas and liquid distributors.

5.4.9 Lean Solvent Cooler

The lean MEA solution is at a temperature of 202°F after cooling with rich MEA leaving the absorber. The MEA is further cooled to 100°F by using cooling water. The heat load is $152 \times 10^6 \text{ Btu/h}$, and cooling water is available at 70°F . The area required is calculated by using Equation 5.1, and the details are shown in Table 5.5.

5.4.10 Interstage Cooler for Absorber

The temperature of the solution in the absorbing column increases to nearly 198°F because of the heat of absorption of CO_2 into the solution. The solution is cooled to about 160°F by using

cooling water. The amount of heat to be removed is 57.75×10^6 Btu/h, and the detailed calculations are shown in Table 5.5.

5.4.11 Lean-Rich MEA Heat Exchanger

The lean MEA solution from the reboiler is at 243°F. This solution can be cooled to 202°F by rich MEA coming from the bottom of the absorber, which is at 160°F. The temperature of the rich MEA increases to 200°F. The heat transferred is 61.3×10^6 Btu/h. The detailed calculations are shown in Table 5.5.

5.4.12 Stripper Overhead Condenser

The stripper overhead contains CO₂, as well as water that is evaporated. Therefore, to separate the CO₂ and water vapor, the gases have to be cooled. The condensed water will be recycled to the top of the stripper. The gases are at a temperature of 212°F and are cooled to 100°F by using cooling water. The heat load is 11.8×10^6 Btu/h, and the heat transfer area required is 7,793 ft².

5.4.13 Stripper Reboiler

The rich solution from the lean-rich MEA heat exchanger is at 200°F, and flashing of this solution will release some CO₂. The remaining CO₂ is desorbed by using heat. The heat necessary for the desorption and for the evaporation of water is supplied by the reboiler. For this process, superheated steam that is condensing at 280°F is used. The heat load is 366.3×10^6 Btu/h, and the heat transfer area is 20,288 ft².

5.4.14 Solvent Circulation Pump

The lean MEA solution from the stripper is at 14.7 psia and is pumped back to the absorber operating pressure of 250 psia. The horsepower (hp) of this circulating pump is estimated as flow work by using the expression

$$\text{pump hp} = H_s \cdot \frac{G}{1,714 \cdot \eta} \quad (5.3)$$

where

H_s = total dynamic head (psia),

G = flow rate of solvent (gal/min),

η = efficiency of the pump, and

1,714 = unit conversion factor.

5.4.15 Carbon Dioxide Product Gas Compressors

The CO₂ from the stripper is at 14.7 psia. This gas has to be compressed to the pipeline pressure of 1,000 psia. This compression is done in three stages, with interstage cooling. The values for the horsepower of the three compressors are 4,051 hp, 3,804 hp, and 3,803 hp. The horsepower (hp) of the compressors is estimated by using the expression

$$\text{hp} = \frac{0.00436}{\eta} \cdot Q_1 \cdot P_1 \cdot \left(\frac{k}{k-1} \right) \cdot \left[\left(\frac{P_2}{P_1} \right)^{(k-1)/k} - 1 \right], \quad (5.4)$$

where

Q_1 = inlet flow rate of gases (ft³/min),

P_1 = inlet pressure (psia),

P_2 = outlet pressure (psia),

k = C_p/C_v = 1.395 (assumed),

η = overall efficiency of compressor = 82% (assumed), and

0.00436 = unit conversion factor.