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KRW Oxygen-Blown Gasification Combined Cycle: Carbon Dioxide Recovery, Transport, and Disposal

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KRW Oxygen-Blown Gasification Combined Cycle: Carbon Dioxide Recovery, Transport, and Disposal

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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 (CO_2). This project emphasizes CO_2-capture technologies combined with integrated gasification combined-cycle (IGCC) power systems. Complementary evaluations address CO_2 transportation, CO_2 use, and options for the long-term sequestration of unused CO_2. Commercially available CO_2-capture technology is providing a performance and economic baseline against which to compare innovative technologies. The intent is to provide the CO_2 budget, or an "equivalent CO_2" 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 CO_2" budget is 1 kg of CO_2 per kilowatt-hour (electric). The base case is a 458-MW (gross generation) IGCC system that uses an oxygen-blown Kellogg-Rust-Westinghouse agglomerating fluidized-bed gasifier, Illinois No. 6 bituminous coal feed, and low-pressure glycol sulfur removal followed by Claus/SCOT treatment to produce a saleable product. Mining, feed preparation, and conversion result in a net electric power production for the entire energy cycle of 411 MW, with a CO_2 release rate of 0.801 kg/kWhe. For comparison, in two cases, the gasifier output was taken through water-gas shift and then to low-pressure glycol H_2S recovery, followed by either low-pressure glycol or membrane CO_2 recovery and then by a combustion turbine being fed a high-hydrogen-content fuel. Two additional cases employed chilled methanol for H_2S recovery and a fuel cell as the topping cycle with no shift stages. From the IGCC plant, a 500-km pipeline took the CO_2 to geological sequestering. For the optimal CO_2 recovery case, the net electric power production was reduced by 37.6 MW from the base case, with a CO_2 release rate of 0.277 kg/kWhe (when makeup power was considered). In a comparison of air-blown and oxygen-blown CO_2-release base cases, the cost of electricity for the air-blown IGCC was 56.86 mills/kWh, and the cost for oxygen-blown IGCC was 58.29 mills/kWh. For the optimal cases employing glycol CO_2 recovery, there was no clear advantage; the cost for air-blown IGCC was 95.48 mills/kWh, and the cost for the O_2-blown case was slightly lower, at 94.55 mills/kWh.

Summary

S.1 Background

Increasing atmospheric concentrations of carbon dioxide (CO_2) have the potential to cause significant climate-related impacts on ecosystems, food production, and economic development, as outlined in the U.S. Climate Change Action Plan (Clinton 1993). Because of these concerns, policies to limit CO_2 emissions are being explored by the United States and other signatories to the Framework Convention on Climate Change put forward at the June 1992 Rio de Janeiro Earth Summit.

For example, Norway has imposed a carbon tax ($50/metric ton of CO_2). As a result, Statoil (Trondheim, Norway) has submitted an engineering proposal for the disposal of CO_2 recovered during natural gas production (Smith 1994). The CO_2 sequestering is to be in an aquifer located 800 m below the sea bed 250 km offshore; as of the date of this publication, however, there has been no final decision to move forward. In Japan, work on disposing of CO_2 in the ocean continues. At the same time, now that this work has reached a more serious stage, there are some significant concerns being expressed by the Japanese government, which would rather see the CO_2 utilized. At present, the only signatories to the Rio Convention on Climate Change that are meeting the goal of maintaining 1990 CO_2 release levels are the United Kingdom, Denmark, and Germany (Stone 1994).

In October 1994, the U.S. Department of Energy (DOE) released greenhouse gas reporting guidelines, but for the present, participation is voluntary. The U.S. actions to stabilize CO_2 may include mandatory conservation — something like establishing Btu/kWh efficiency ratings for electric power plants similar to the fleet fuel efficiency standards for automobiles. Other options may include taking strong energy conservation measures, switching from coal to natural gas for electric power generation, capturing and sequestering CO_2, or substituting nonfossil energy sources for fossil fuel combustion. Discussion of the issues has drawn considerable interest in power generating systems that minimize the production of CO_2 and are amenable to CO_2 capture. In the event that natural gas would no longer be widely available at low prices, integrated gasification combined-cycle (IGCC) systems would be an attractive emerging electric power generating technology option because they provide high energy-conversion efficiency when current technology is used. They also offer the prospect of even higher efficiencies if higher-temperature turbines and hot-gas cleanup systems are developed. In addition, they have demonstrated very low emission levels for sulfur and nitrogen species. Finally, IGCC plants produce flue-gas streams with concentrated CO_2 and high levels of CO, which can be easily converted to CO_2 if the recovery and sequestering of CO_2 are mandated in the future.

The project objective is to develop engineering evaluations of technologies used to capture, use, and dispose of CO_2 when combined with oxygen (O_2)-blown Kellogg-Rust-Westinghouse (KRW) IGCC power systems. This study is an extension of earlier work done for the Morgantown Energy Technology Center (METC) that considered these questions for air-blown KRW IGCC power systems (Doctor et al. 1994).

S.2 Overview of Energy Cycle

The energy system definition for this study extends from the coal mine to the final geological repository for the CO_2, as shown in Figure S.1. The location of the IGCC plant is specified as the midwestern United States, and this study assumes it is 160 km by rail from the Old Ben No. 26 mine in Sesser, Illinois. Details of the IGCC portion of the system are taken from an Electric Power Research Institute (EPRI) report (Gallaspy 1990a), which describes an electric power station using an O_2-blown KRW gasifier, while a follow-up METC report (Gallaspy 1990b) describes a plant using an air-blown KRW gasifier with in-bed sulfur removal. In each case studied, the CO_2 recovery technologies have been integrated into that plant design as much as possible to limit efficiency losses. For each part of the energy system, CO_2 emissions have been either computed directly from process stream compositions or calculated from energy consumption on the basis of a "CO_2 equivalence" of 1 kg of CO_2 per kilowatt-hour (electric) (kWhe). In this way, a total CO_2 budget for the system can be derived and compared with a total CO_2 budget for other options, thereby taking into account effects outside the immediate plant boundary.

S.3 Mining, Preparation, and Transportation of Raw Materials

All seven cases presented here were adjusted to be on a consistent basis of 4,110 tons/d (stream day) of Illinois No. 6 coal from the Old Ben No. 26 mine. This bituminous 2.5%-sulfur coal contains 9.7% ash. The underground mine is associated with a coal preparation plant. The assumption is that the IGCC power plant is 160 km from the mine and the coal is shipped by rail on a unit train. The impact of coal mining and shipment on the energy budget is 2.41 MW of power use and 2,879 kg/h of CO_2 emissions.

Limestone is used for in-bed sulfur capture in the two air-blown gasifier cases. It is assumed that the limestone is extracted from a quarry about 160 km from the plant and transported by rail to the plant site. The impact of limestone mining and shipment on the energy budget is 0.27 MW of power use and 406 kg/h of CO_2 emissions.

S.4 Handling of Coal and Limestone

The coal preparation system for the O_2-blown IGCC plant includes equipment for unloading the coal from the unit train, passing it through magnetic separators, and then conveying it to a hammermill. From there, the coal is conveyed to storage silos from which it is recovered in a fluidized stream for use in the gasifier. The coal is not dried for the O_2-blown cases. The impact of coal preparation on the energy budget is 0.85 MW of power use and no CO_2 emissions (these will be combined with the overall emissions from the IGCC plant). Drying the coal was not considered for this case.

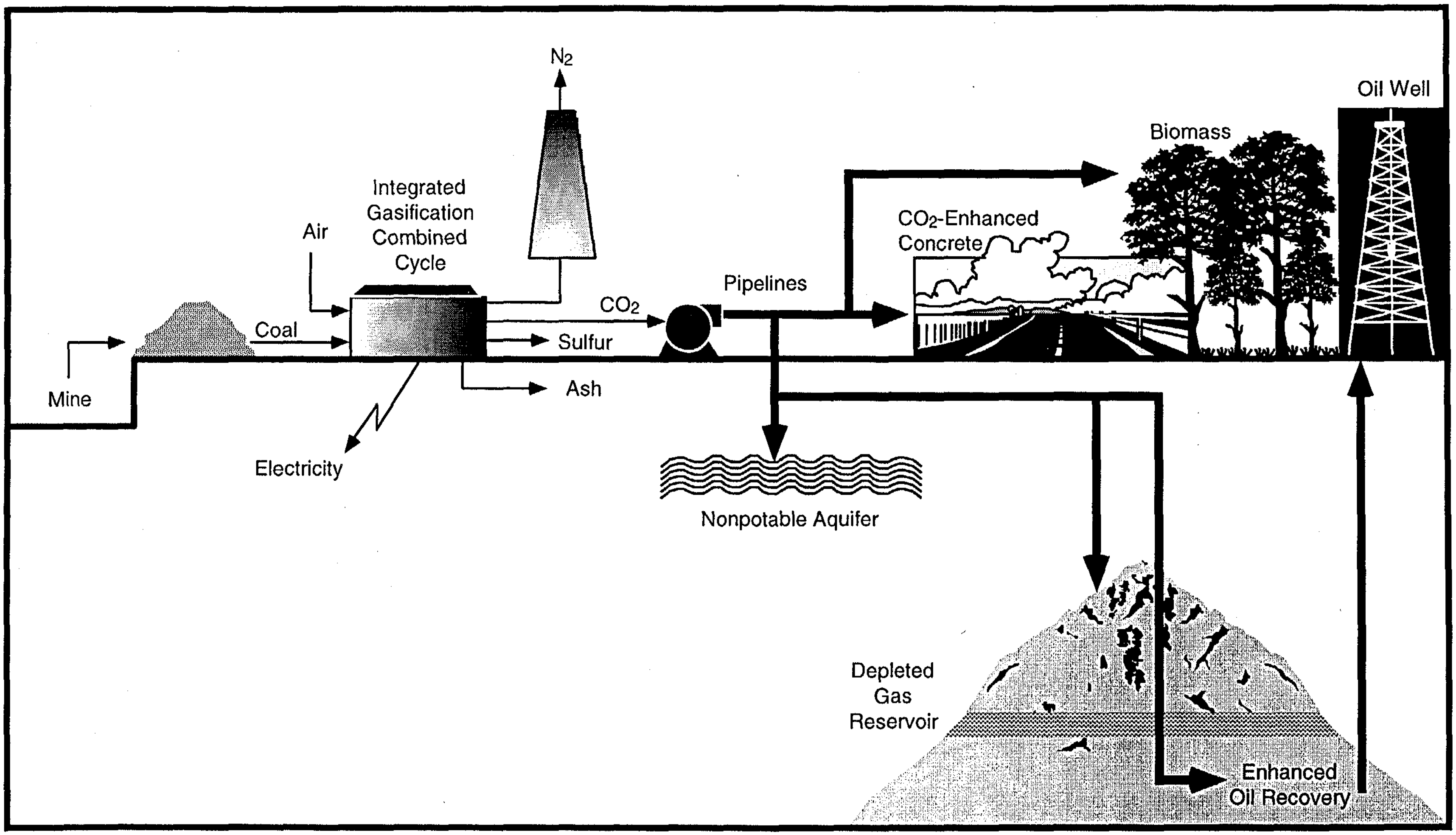

FIGURE S.1 Simplified Overview of the Energy Cycle Components for CO_2 Recovery

By way of contrast, the coal preparation system for the air-blown IGCC plant includes equipment for unloading the coal from the unit train, passing it through magnetic separators, and then conveying it to silos for 14-h storage. 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 amount of CO_2 being emitted from the energy cycle that is not reclaimed and presents a possible opportunity for further reduction. The coal is then held in a bunker for 2 h, from which it is pneumatically conveyed to surge bins ahead of the gasifier lockhoppers. The sulfator emits 11,374 kg/h of CO_2. Limestone is crushed in two pulverizers and 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 use for coal and limestone preparation is 3.49 MW.

S.5 Base Cases for Integrated Gasification Combined Cycle

S.5.1 Gasifier Island

The O_2-blown base case employs an air-separation plant producing 2,100 tons/d of 95% oxygen from a commercial package designed by Air Products. The KRW process is an O_2-blown, dry-ash, agglomerating, fluidized-bed process. A simplified schematic for this process appears in Figure S.2. Three parallel gasifier trains operating at 450 lb/in.^{2} gauge (psig) and 1,850°F are included in the design. Following gasification, cyclones recover 95% of the fines; gas cooling and high-efficiency particulate removal follow. For the base case, glycol H_2S recovery provides a feed to a conventional Claus tail-gas cleanup system. Hence, the significant differences between the O_2-blown and air-blown cases are that the O_2-blown cases cool the product gas for sulfur cleanup and produce a sulfur product for the market, while the air-blown cases employ hot-gas cleanup and produce a landfill product. The impact of the gasifier island operation on the energy budget is 36.82 MW of power use and 6,153 kg/h of CO_2 emissions for the O_2-blown base case.

The air-blown base case uses in-bed sulfur removal. A simplified schematic for this process appears in Figure S.3. The system includes two heavy-duty industrial gas turbines (2,300°F 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 not integrated into the later CO_2 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. Inlet gas temperatures are maintained at approximately 1,000°F. 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 impact of the gasifier island operation on the energy budget is 20.12 MW of power use and 137 kg/h of CO_2 emissions for the air-blown base case.

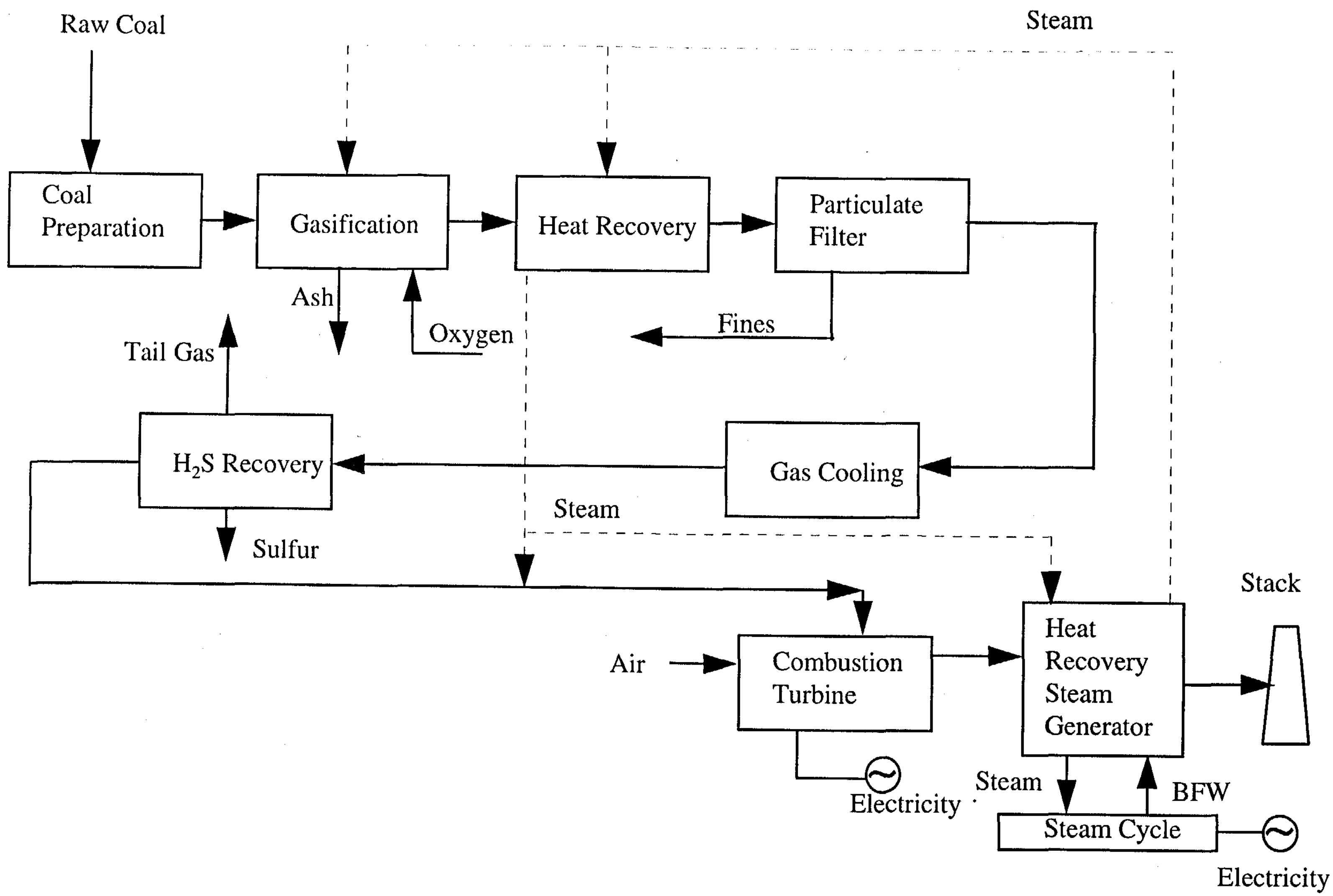

FIGURE S.2 Block Diagram of the Base-Case Oxygen-Blown KRW IGCC System

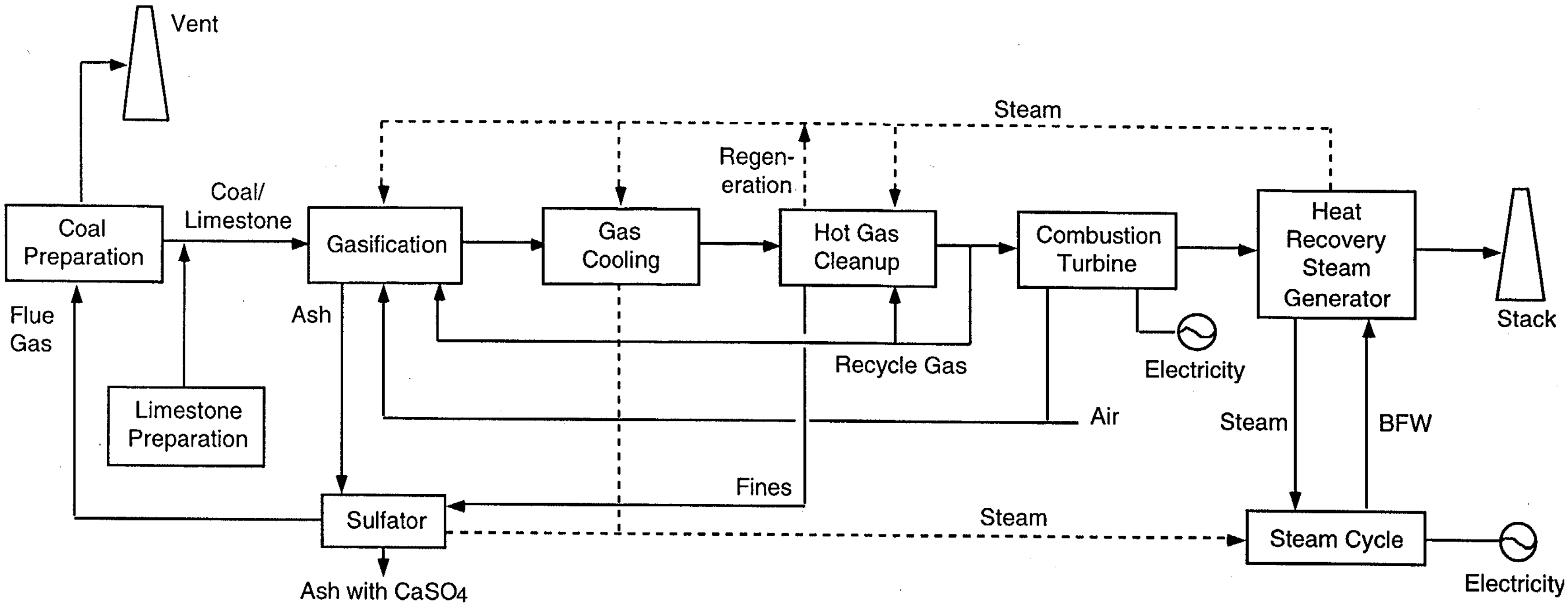

FIGURE S.3 Block Diagram of the Base-Case Air-Blown KRW IGCC System

S.5.2 Power Island

Both the O_2-blown and air-blown base cases employ a turbine topping cycle and a steam bottoming cycle based on two heavy-duty GE MS701F industrial gas turbines with a 2,300°F firing temperature. The impact on the energy budget of the power island operation is 7.02 MW of power use for the O_2-blown base case and 10.58 MW of power use for the air-blown base case. For the O_2-blown base case, gross power generation is 458.20 MW, with a net generation of 413.50 MW; for the air-blown base case, gross power generation is 479.63 MW, with a net power generation of 445.44 MW.

S.6 Integrated Gasification Combined Cycle with CO_2 Recovery

Several changes were made to the base-case IGCC plant to incorporate CO_2 recovery. For the turbine topping-cycle studies (Cases 1 and 2), these changes entailed processing the cleaned fuel gas through a "shift" reaction to convert the CO to CO_2, recovering the CO_2, and then combusting the low-CO_2 fuel gas in a modified turbine/steam cycle to produce electricity. Gas cleaning and sulfator performance were considered to be unaffected by these changes. In contrast, the fuel cell topping-cycle studies (Cases 3 and 4) required a highly cleaned gasifier without use of the water-gas shift reaction to be used by the fuel cells. A block diagram of the O_2-blown IGCC system with CO_2 recovery appears in Figure S.4, while the air-blown system with CO_2 recovery appears in Figure S.5.

The fuel gas from the KRW process is high in CO. Conversion of the CO to CO_2 in the combustion process would result in substantial dilution of the resulting CO_2 with nitrogen from the combustion air and with water from the combustion reaction. If the CO_2 is removed before combustion, a substantial savings in the cost of the CO_2 recovery system is possible because of reduced vessel size and solvent flow rate. The CO in the fuel gas must first be converted to CO_2 by the shift reaction:

$$CO + H_2O \implies CO_2 + H_2 .$$

The resulting CO_2 can then recovered, leaving a hydrogen-rich fuel for use in the gas turbine. The shift reaction is commonly accomplished in a catalyst-packed tubular reactor that uses a relatively low-cost iron-oxide catalyst. High CO_2 recovery is best achieved by staged reactors that allow for cooling between stages; hence, a two-stage system configured to achieve 95% conversion of CO to CO_2 was found to be optimal.

Commercial CO_2-removal technologies all 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 before CO_2 recovery is injected into the clean fuel-gas stream as it is heated by recovered heat following CO_2 removal. This option allows additional

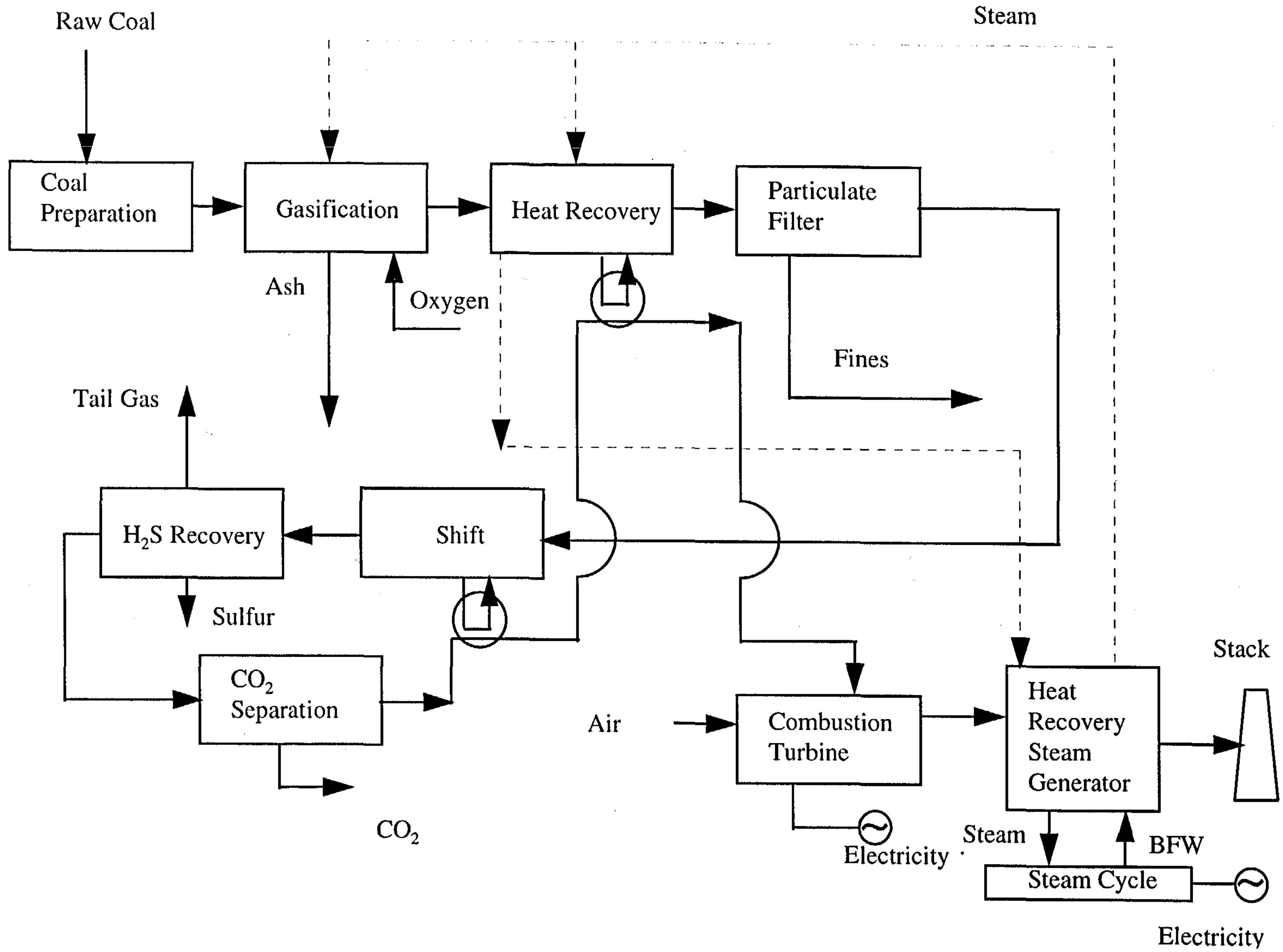

FIGURE S.4 Block Diagram of the Base-Case Oxygen-Blown KRW IGCC System Modified for CO_2 Recovery

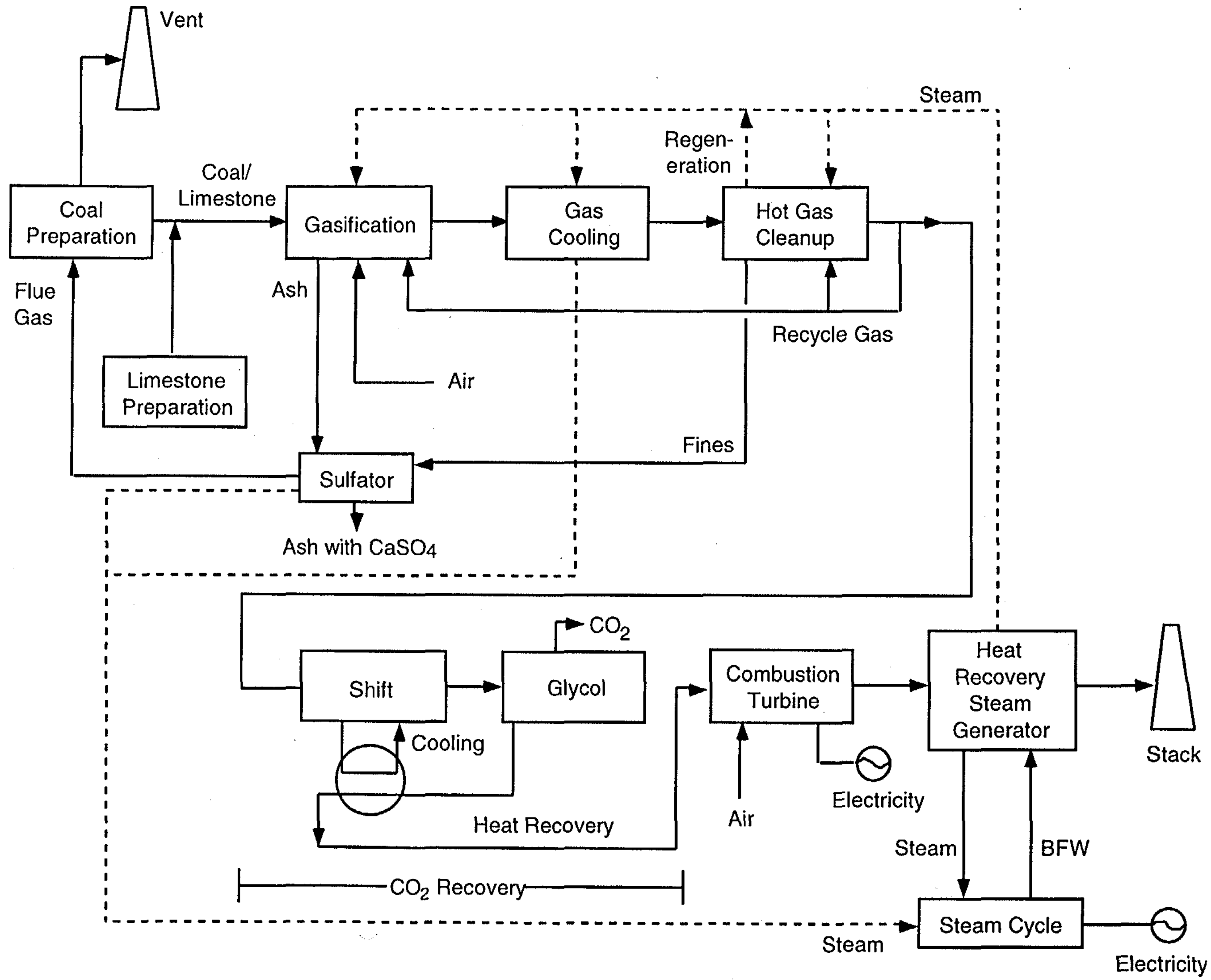

FIGURE S.5 Block Diagram of the Base-Case Air-Blown KRW IGCC System Modified for CO_2 Recovery heat to be absorbed before combustion and increases the mass flow rate through the gas turbine. The balance of the thermal energy is used in the heat recovery steam generator for feedwater heating and steam generation.

Commercial CO_2-recovery processes operate by absorption of the CO_2 in a liquid solvent and subsequent regeneration of the solvent to release the CO_2. 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 above. Furthermore, lower temperatures favor absorption, thereby reducing the necessary solvent flow rate. This situation 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, since it is usually accomplished by flashing (pressure reduction) and/or heating. If flashing is employed, repressurization of the solvent is required. Heating is generally accomplished by the extraction of steam from the steam cycle.

In addition to supplying data on an oxygen-blown base case and an air-blown base case (both without CO_2 recovery), this study evaluates five CO_2 recovery power cycles: four oxygen-blown cases and the optimal air-blown case discussed in our previous study, ANL/ESD-24 (Doctor et al. 1994).

	Case 1	Case 2	Case 3	Case 4	ANL/ESD-24
Gasifier oxidant	Oxygen	Oxygen	Oxygen	Oxygen	Air
H_2S recovery	Glycol	Glycol	Methanol	Methanol	In-bed/ZnTi
CO_2 recovery	Glycol	Membrane	Glycol	Membrane	Glycol
Topping cycle	Turbine	Turbine	Fuel cell	Fuel cell	Turbine
Bottoming cycle	Steam	Steam	Steam	Steam	Steam

For the optimal O_2-blown CO_2 recovery case (Case 1), the net electric power production was reduced by 37.6 MW from the base case, with a 0.277-kg/kWhe CO_2 release rate (when makeup power was considered). The low-pressure glycol system, which does not require compression of the synthesis gas before absorption, appears to be the best system studied.

S.7 Pipeline Transport of CO_2

Once the CO_2 has been recovered from the fuel-gas stream, its transportation, utilization, and/or disposal remain significant issues. In a previous study for METC (Doctor et al. 1994), the issues associated with the transport and sequestering of CO_2 were considered in greater detail; they serve as the basis for this work. The CO_2 represents a large-volume, relatively low-value by-product that cannot be sequestered in the same way as most coal-utilization wastes (i.e., by landfilling). Large volumes of recovered CO_2 are likely to be moved by pipeline, and if sequestering were required, new pipelines would likely need to be constructed. In some cases, existing pipelines could be used, perhaps in a shared mode with other products. Costs for pipeline construction and use vary greatly on a regional basis within the United States. The recovered CO_2 represents more than 3 million normal cubic meters per day of gas volume. It is assumed that the transport and sequestering process releases approximately 2% of the recovered CO_2.

S.8 Sequestering of CO_2

Proposals have been made to dispose of CO_2 in the ocean depths. However, many questions of engineering and ecological concern associated with such options remain unanswered, and the earliest likely reservoir is a land-based geological repository (Hangebrauck et al. 1992). A portion of the CO_2 can be used for enhanced oil recovery, which sequesters a portion of the CO_2, or the CO_2 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 have been prepared; they take into account 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 CO_2 gas for a 30-year period without requiring any additional compression costs for operation. The pipeline transport and sequestering process represents approximately 26 mills/kWh for the CO_2-recovery cases.

S.9 Energy Consumption and CO_2 Emissions

Data on the energy consumption and CO_2 emissions for the O_2-blown base case are provided in Table S.1. These can be compared with data on the optimal case that employs low-pressure glycol CO_2 recovery and a turbine topping cycle (i.e., Case 1) provided in Table S.2.

S.10 Economic Summary

A comparison of the cost of electricity for the CO_2 release base cases revealed that the cost for the air-blown IGCC was 58.29 mills/kWh, and the cost for the O_2-blown case was 56.86 mills/kWh (Table S.3). There was no clear advantage for the optimal cases employing glycol CO_2 recovery; the cost for the air-blown IGCC was 95.48 mills/kWh, and the cost for the O_2-blown case was slightly lower, at 94.55 mills/kWh.

S.11 References for Summary

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TABLE S.1 Energy Consumption and CO_2 Emissions for Oxygen-Blown Base Case with No CO_2 Recovery

	Electricity MW	CO2 release kg/h
Mining and Transport		
Raw Coal in Mine	-2.36	2,356
Coal Rail Transport	-0.05	523
Subtotal	-2.41	2,879
IGCC Power Plant		
Coal Preparation	-0.85	0
Gasifier Island	-36.82	6,153
Power Island	-7.02	320,387
Subtotal	-44.70	326,540
Power - Gas Turbine	298.80	
Power - Steam Turbine	159.40	
GROSS Power	**458.20**	
NET Power	**413.50**	
Pipeline/Sequester	0.00	0
Energy Cycle Power Use	**-47.11**	
NET Energy Cycle	**411.09**	**329,419**
CO2 emission rate/net cycle	**0.801**	**kg CO2/kWh**
Power use/CO2 in reservoir	**N/A**	**kWh/kg CO2**

TABLE S.2 Energy Consumption and CO_2 Emissions for Optimal Oxygen-Blown Case with CO_2 Recovery: Case 1

	Electricity MW	CO2 release kg/h
Mining and Transport		
Raw Coal in Mine	-2.36	2,356
Coal Rail Transport	-0.05	523
Subtotal	-2.41	2,879
IGCC Power Plant		
Coal Preparation	-0.85	0
Gasifier Island	-36.82	6,153
Power Island	-7.02	320,387
Glycol Circulation	-5.80	-260,055
Glycol Refrigeration	-4.50	
Power Recovery Turbines	3.40	
CO2 Compression (to 2100psi)	-17.30	
Subtotal	-68.90	66,485
Power - Gas Turbine	284.80	
Power - Steam Turbine	161.60	
GROSS Power	**446.40**	
NET Power	**377.50**	
Pipeline/Sequester		
Pipeline CO2		260,055
Pipeline booster stations	-1.64	1,637
Geological reservoir (2% loss)	0.00	-254,854
Subtotal	-1.64	6,839
Energy Cycle Power Use	-72.95	
NET Energy Cycle	373.45	76,202
Derating from O2-Base Case	37.64	
Make-up Power	37.64	37,637
TOTAL	411.09	113,840
CO2 emission rate/net cycle	**0.277**	**kg CO2/kWh**
Power use/CO2 in reservoir	**0.148**	**kWh/kg CO2**

TABLE S.3 Summary of Comparative Costs of IGCC Systems

Case		BASE	BASE	Case #1	Case #2	Case#3	Case #4	ESD-24/Glycol
Gasifier Oxidant		Oxygen	Air	Oxygen	Oxygen	Oxygen	Oxygen	Air
H2S Recovery		Glycol	In-Bed/ZnTi	Glycol	Glycol	Methanol	Methanol	In-Bed/ZnTi
CO2 Recovery		none	none	Glycol	Membrane	Glycol	Membrane	Glycol
Topping Cycle		Turbine	Turbine	Turbine	Turbine	Fuel Cell	Fuel Cell	Turbine
Bottoming Cycle		Steam	Steam	Steam	Steam	Steam	Steam	Steam
Component	Unit							
Base Plant Capital	$/kW	$1,332	$1,253	$1,485	$1,703	$2,560	$2,746	$1,487
CO2 Control Capital	$/kW	$0	$0	$202	$602	$145	$905	$246
Total Plant Capital	**$/kW**	**$1,332**	**$1,253**	**$1,687**	**$2,305**	**$2,705**	**$3,651**	**$1,733**
Power Plant Annual Cost	$K	$137,253	$144,212	$203,238	$242,336	$249,786	$287,547	$204,288
Power Cost								
Base Plant Power Cost	mills/kWh	58.29	56.86	70.64	101.62	102.45	132.19	71.46
Pipeline Cost	mills/kWh	0	0	23.91	27.35	26.53	28.76	24.02
Net Power Cost	**mills/kWh**	**58.29**	**56.86**	**94.55**	**128.97**	**128.98**	**160.95**	**95.48**
Coal Energy Input	10^6Btu/h	3839	3839	3839	3839	3839	3839	3839
Gross Power Output	MW	458.20	479.63	446.40	417.60	418.50	413.20	460.88
In Plant Power Use	MW	44.70	34.19	68.90	87.60	78.39	99.40	85.11
Net Plant Output	**MW**	**413.50**	**445.44**	**377.50**	**330.00**	**340.11**	**313.80**	**375.77**
Net Heat Rate	Btu/kWh	9284	8618	10170	11633	11288	12234	10216
Thermal Efficiency - HHV	%	36.78%	39.62%	33.58%	29.35%	30.25%	27.91%	33.42%
Out of Plant Power Use	MW	2.41	4.18	4.05	3.87	4.05	4.12	4.47
Net Energy Cycle Power	**MW**	**411.09**	**441.26**	**373.45**	**326.13**	**336.06**	**309.68**	**371.30**
Net Energy Cycle Heat Rate	Btu/kWh	9339	8700	10280	11771	11424	12397	10339
Thermal Efficiency - HHV	%	36.56%	39.25%	33.21%	29.01%	29.89%	27.54%	33.02%
Net Energy Cycle Power	MW	411.09	441.26	373.45	326.13	336.06	309.68	371.30
Net Replacement [Added] Power	MW	0.00	(30.17)	37.64	84.96	75.03	101.41	39.79
Net Grid Power	**MW**	**411.09**	**411.09**	**411.09**	**411.09**	**411.09**	**411.09**	**411.09**

1 Introduction

1.1 Background

Argonne National Laboratory report ANL/ESD-24, *Gasification Combined Cycle: Carbon Dioxide Recovery, Transport, and Disposal* (Doctor et al. 1994), provides a comparison of carbon dioxide (CO_2) recovery options for an integrated gasification combined-cycle (IGCC) plant using an air-blown Kellogg-Rust-Westinghouse (KRW) gasifier that employs an in-bed sorbent system for sulfur recovery. The comparison focuses on the relative energy penalty, capital investment, and CO_2 reduction for five commercial CO_2 recovery processes. The potential for two advanced processes is also discussed in that report. The comparison of energy penalty and CO_2 emission reduction is based on the full energy system, including mining, transportation, coal preparation, conversion, and gas treatment. Emissions associated with replacement power to compensate for the energy penalty of the CO_2 recovery processes are included in the accounting. Compared with CO_2 recovery from a conventional coal plant, the essential advantage of coupling a CO_2 recovery system to a coal-gasification-based power plant is that removal of CO_2 from gasifier fuel gas is more economical than removal of CO_2 from flue gas produced by conventional coal combustion. Primarily, this economy results from the lesser dilution of the fuel gas with atmospheric nitrogen. Thus, a substantially smaller volume of gas must be processed, and the CO_2 concentration in that gas is higher than in postcombustion flue gas. This advantage is expected to be more pronounced for a gasifier that uses oxygen rather than air as the oxidant. Further advantage is derived from the higher operating pressure associated with gasification in general and with the oxygen-blown case in particular.

Because of the dilution with nitrogen, air-blown gasifiers produce low-Btu gas, which has a heating value in the range of 90 to 170 Btu per standard cubic foot (scf). Oxygen-blown gasifiers produce a medium-Btu gas, which has a heating value of about 250 to 400 Btu/scf. In the air-blown case, substantially more of the energy value of the coal is manifested as sensible heat in the fuel gas. Losses associated with heat recovery and the cost of heat recovery equipment are therefore more important in the air-blown case. Thus, the economic value of high-temperature gas cleanup is greater in the air-blown case. The oxygen-blown cases considered here use low-temperature gas cleanup processes for sulfur removal. The air-blown cases considered in ANL/ESD-24 use a high-temperature system for sulfur removal.

1.2 Goals, Objectives, and Approach

The present volume supplements ANL/ESD-24. Four additional cases have been analyzed for this supplement. Table 1.1 summarizes the plant configurations for these cases. All four cases employ an oxygen-blown KRW gasifier with cold gas cleanup. Two cases use a gas turbine topping cycle and two cases use a fuel cell topping cycle. For the fuel cell cases, chilled methanol is used for H_2S recovery because of tight specifications (H_2S at less than 1 part per million, volume [ppmv]) imposed to protect the fuel cell. For the gas turbine cases, a glycol-

TABLE 1.1 Alternative Plant Configurations

Case	H_2S Recovery	CO_2 Recovery	Topping Cycle	Bottoming Cycle
1	Glycol	Glycol	Gas turbine	Steam
2	Glycol	Membrane	Gas turbine	Steam
3	Chilled methanol	Glycol	Fuel cell	Steam
4	Chilled methanol	Membrane	Fuel cell	Steam

based physical absorption system is used for H_2S recovery. These systems are analyzed for energy penalty and costs associated with the CO_2 recovery system and for net CO_2 removal. A comparison with the air-blown cases described in the earlier report is also provided.

2 Mining

2.1 Mining, Preparation, and Transportation of Raw Materials

All seven cases presented here were adjusted to be on a consistent basis of 4,110 tons/d (stream day) of Illinois No. 6 coal from the Old Ben No. 26 mine. The underground mine is associated with a coal preparation plant. It is assumed that the IGCC power plant is 160 km from the mine and the coal is shipped by rail on a unit train. The ultimate analysis for this coal appears in Table 2.1. The impact on the energy budget of coal mining and shipment is 2.41 MW of power use and 2,879 kg/h of CO_2 emissions.

Limestone is used for in-bed sulfur capture in the two air-blown gasifier cases. It is assumed that the limestone is extracted from a quarry about 160 km from the plant and transported by rail to the plant site. The impact on the energy budget of limestone mining and shipment is 0.27 MW of power use and 406 kg/h of CO_2 emissions.

2.2 Coal and Limestone Handling

The coal preparation system for the O_2-blown IGCC plant includes equipment for unloading the coal from the unit train, passing it through magnetic separators, and then conveying it to a hammermill. From there, the coal is conveyed to storage silos from which it is recovered in a fluidized stream for use in the gasifier. The coal is not dried for the O_2-blown cases. The impact on the energy budget of coal preparation is 0.85 MW of power use and no CO_2 emissions (these will be combined with the overall emissions form the IGCC plant.) Drying the coal was not considered for this case.

By way of contrast, the coal preparation system for the air-blown IGCC plant includes equipment for unloading the coal from the unit train, passing it through magnetic separators, and then conveying it to silos for 14-h storage. 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 amount of CO_2 being emitted from the energy cycle that is not reclaimed and presents a possible 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 lockhoppers. The sulfator emits 11,374 kg/h of CO_2. Limestone is crushed in two pulverizers and 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 for coal and limestone preparation is 3.49 MW.

TABLE 2.1 Analysis of Coal from Illinois No. 6 Seam, Old Ben No. 26 Mine

Component	Ultimate Analysis as-Received (wt %)	Property	Value
Moisture	11.12	Temperature of ash fusion (reducing conditions) (°C)	
Carbon	63.75	Initial deformation	1,201
Hydrogen	4.50	Softening (H = W)	1,238
Nitrogen	1.25	Softening (H = 1/2W)	1,285
Chlorine	0.29	Fluid	1,324
Sulfur	2.51		
Ash	9.70	Higher heating value (J/kg)	27.13×10^6
Oxygen (by diff.)	6.88		
Total	100.0		

3 Oxygen-Blown Base Case with No CO_2 Recovery

3.1 Design Basis

Figure 3.1 provides an overview of the base-case plant configuration, which does not incorporate CO_2 recovery. This layout is typical of an oxygen-blown IGCC with cold-gas cleanup in which H_2S is removed by an acid gas removal system following gas cooling. The base-case analysis performed by Southern Company Services and others with sponsorship from the Electric Power Research Institute (EPRI 1990) assumes the use of Selexol®, a commercial glycol-based process, for this H_2S removal. The cleaned gas is then saturated and reheated with steam before it is used in the gas turbine. The turbine exhaust gas is used to raise steam for a Rankine cycle steam plant. Steam from the heat recovery steam generator is also supplied to the gasifier. Oxidant is provided by an air separation plant. Three KRW gasifiers with the capacity to provide 42% of plant requirements are used to ensure high reliability.

The oxygen is produced by cryogenic distillation in a separate air plant that is not integrated with the gasifier and power generation systems except through direct use of the oxygen product. Opportunities for integration do exist but are not incorporated in current plans for oxygen-blown gasifiers. The KRW gasifier is an agglomerating fluidized-bed reactor that operates at 450 $lb/in.^2$ gauge (psig) and 1,850°F. Operation in the agglomerating regime enhances overall plant performance (EPRI 1990; Takematsu 1991). The KRW process has been demonstrated in extensive pilot scale tests, but no commercial demonstration unit has been built. One commercial-scale air-blown unit is under construction.

Hot gas from the gasification reactor contains ash, char, and sulfur species that must be removed before combustion. Ninety-five percent of the ash and char are removed in cyclones after the initial cooling of the hot (1,850°F) raw gas to 1,350°F. Following further cooling to 450°F, the remaining fines are removed by sintered metal filters. Final cooling to 100°F is accomplished by water quench prior to acid gas removal by the Selexol process. The concentrated H_2S stream from the Selexol process is treated in a Claus unit for sulfur recovery. Design sulfur recovery is 96.4%.

3.2 Material Balance

Material flows are summarized in Table 3.1,[1] which provides a comparison of the reference oxygen-blown base case with an air-blown base case using in-bed sulfur capture.

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[1] Design specifications used in this report are a combination of specifications from two documents. *Assessment of Coal Gasification/Hot Gas Cleanup Based on Advanced Gas Turbine Systems* (Gallaspy 1990b) provided the design basis for the air-blown systems reviewed in ANL/ESD-24. This document also includes limited information on one oxygen-blown case, an update of a design evaluated in an earlier report, *Southern Company Service's Study of a KRW-Based GCC Power Plant* (Gallaspy 1990a). This earlier report has been relied on for certain design details, although flows have been scaled to agree with the updated plant specifications in the former report. The update is primarily a result of a substantial increase in the performance rating of the GE gas turbine selected as part of the design basis.

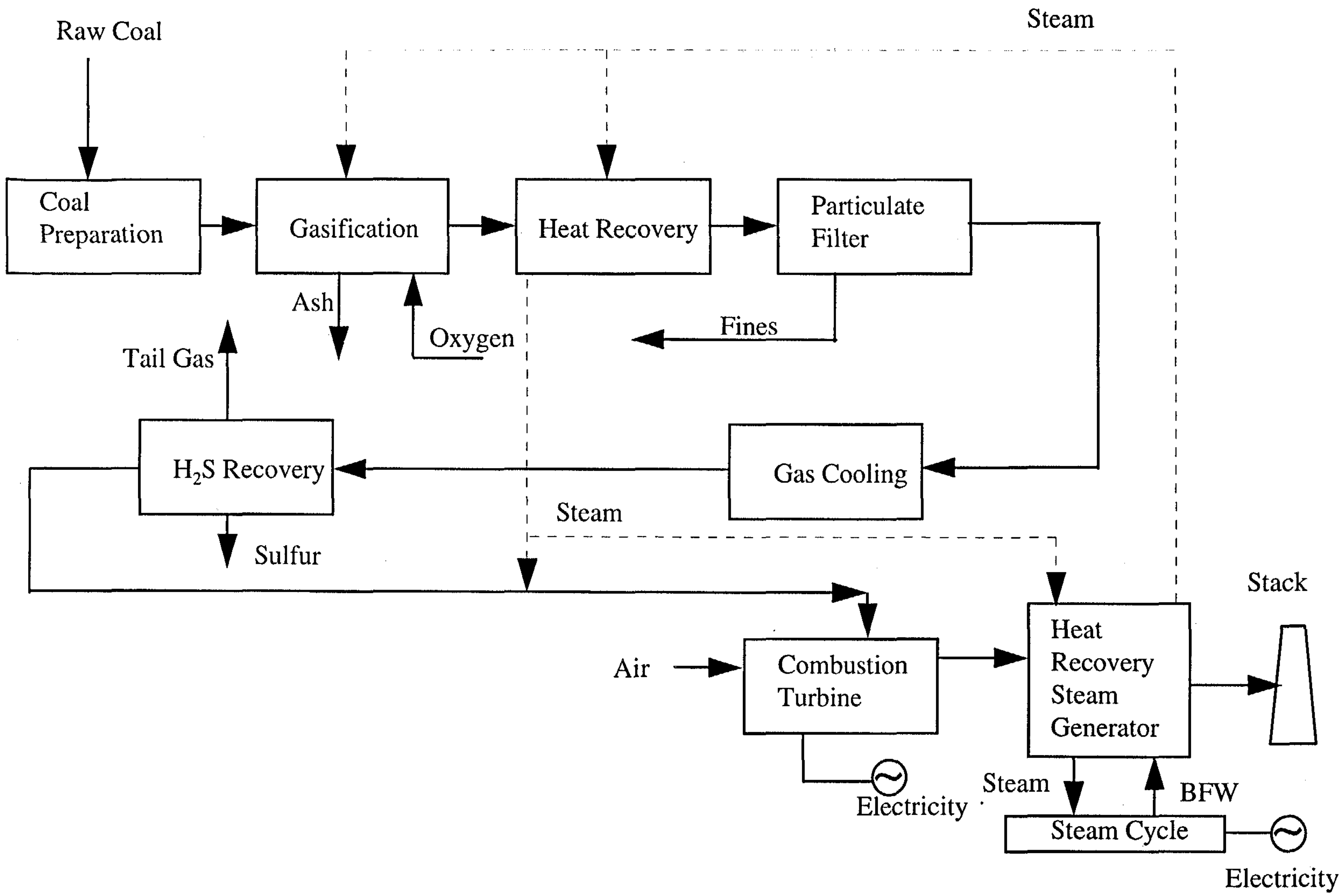

FIGURE 3.1 Block Diagram of the Base-Case Oxygen-Blown KRW IGCC System

TABLE 3.1 Material Flows for Oxygen-Blown and Air-Blown Base Cases

Material Flow (tons/d)	Oxygen-Blown Base Case	Air-Blown Base Case
Coal (prepared)	3,845	3,792
Limestone	0	1,053
Air	0	12,888
Oxygen	2,347	0
Solid waste	492	1,231
Sulfur	78	0
CO_2 (gasifier only)	8,586	9,600
SO_2 (gasifier only)	6.92	1.24
Net power output (MW)	413.5	458.4

3.3 Gas Turbine, Steam Cycle, and Plant Performance

Nominal capacity of the reference plant is 413.5 MW net, including 298.8 MW from the gas turbines and 159.4 MW from the steam cycle minus 44.7 MW for station service load. The net plant heat rate is 9,039 Btu/kWh at full load. The power island incorporates two GE MS7001F combustion turbines, two heat recovery steam generators, and one reheat steam turbine.

3.4 Economics

A summary of capital and operating costs is provided in Section 9.

4 Case 1 — Gas Turbine Topping Cycle and Glycol CO_2 Recovery

As noted in the introduction, two topping cycle options have been studied: gas turbines and fuel cells. Two CO_2-recovery options have been investigated for use in conjunction with the gas turbine topping cycle: a glycol-based absorption system and a two-stage membrane system. Detailed design, performance, and cost information is presented in this section for the gas turbine option with glycol-based CO_2 recovery. A glycol system is also used for sulfur recovery.

4.1 Design Basis

Figure 4.1 shows the addition of a glycol-based CO_2 recovery system to the reference IGCC plant. The membrane system occupies a similar position in the overall scheme, although stream conditions differ somewhat for the two recovery options. The CO_2 recovery follows H_2S recovery, which is preceded by a shift reaction to convert the CO-rich synthesis gas to a hydrogen-rich gas diluted by CO_2. This shift is accomplished in two stages for economical use of catalysts and is integrated with the power cycle by heat exchange with the CO_2-lean fuel gas. The role of these processes is clarified in Figure 4.2, which displays the gas composition at various process stages. Note the dramatic increase in CO_2 during the shift reaction and the simultaneous reduction in CO. The removal of CO_2 is evident by contrast of the absorber inlet concentration and the dry fuel gas product. Nominally 90% CO_2 recovery is accomplished by a combination of 95% conversion of CO in the shift and 95% recovery of the resulting CO_2 in the gylcol process. Somewhat less recovery is accomplished in the membrane case because of membrane performance limitations. Table 4.1 is a summary of principal material flows for the base case and for this design option.

4.2 Shift Reactor

The shift reactor relies on steam in the presence of a catalyst to convert CO to CO_2. Catalyst performance is temperature sensitive, so that reduction in gas stream temperature is required for efficient conversion. Economic use of catalysts dictates that the shift reaction be carried out in two stages. In the first stage, an iron-based catalyst is used, which is effective above 650°F. In the second stage, a copper-based catalyst is used, which is effective at lower temperatures. Cooling is required before both stages to remove sensible heat and heat of reaction associated with the shift reaction. The effective use of the heat removed in cooling the gas is an important design consideration. The shift system design is discussed in detail in ANL/ESD-24. In that report, it is demonstrated that a considerable overall cycle efficiency advantage is gained by allocating as much of the sensible heat as possible to the cleaned fuel gas feed to the turbine. A similar design is incorporated here. This involves the optimization of the two catalytic reactors and of the heat integration. Figure 4.3 is a flow diagram of the shift reactor system showing the heat integration. The high-temperature heating and humidification of the fuel gas stream is

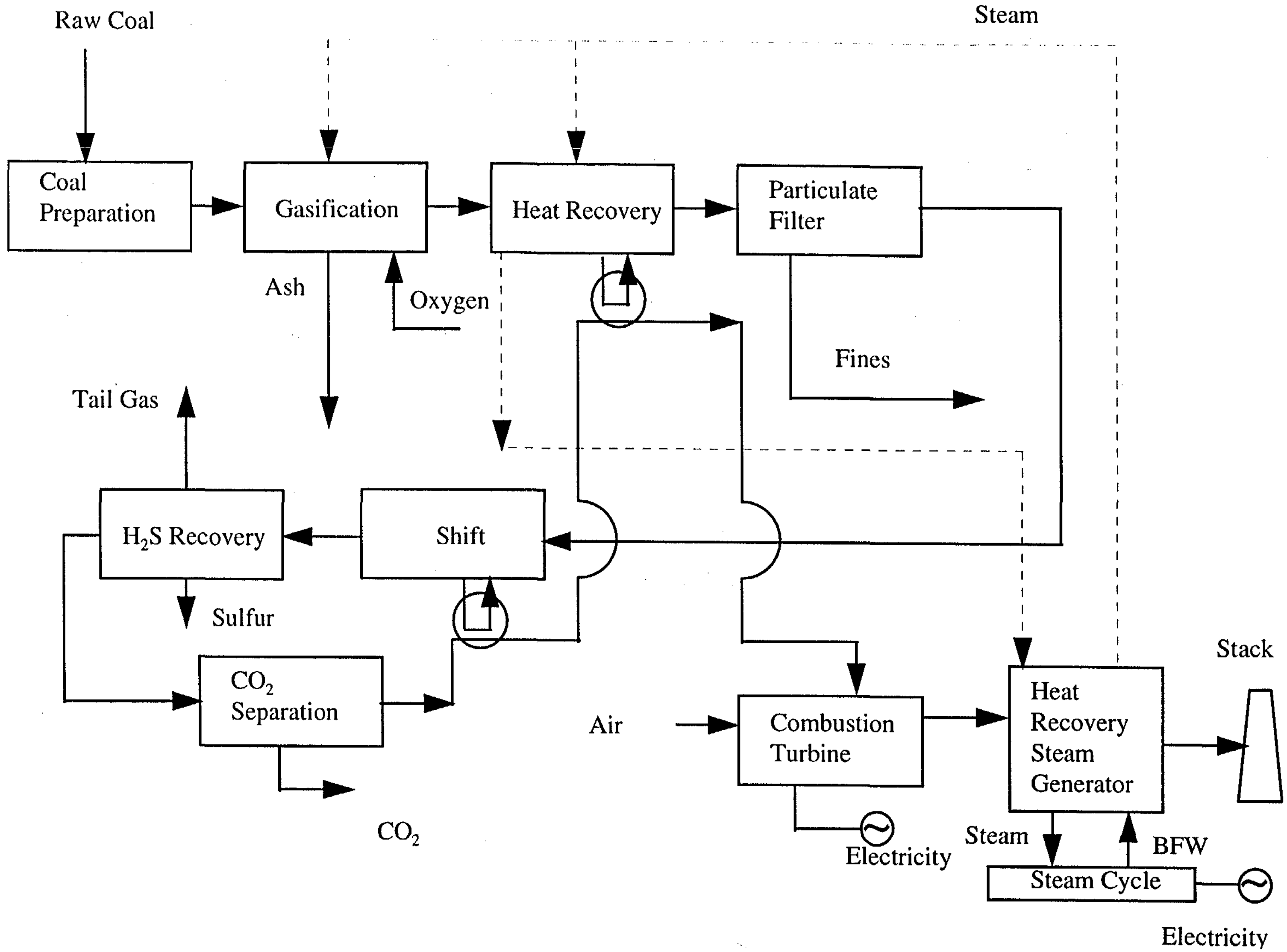

FIGURE 4.1 Block Diagram of the Base-Case Oxygen-Blown KRW IGCC System Modified for CO_2 Recovery

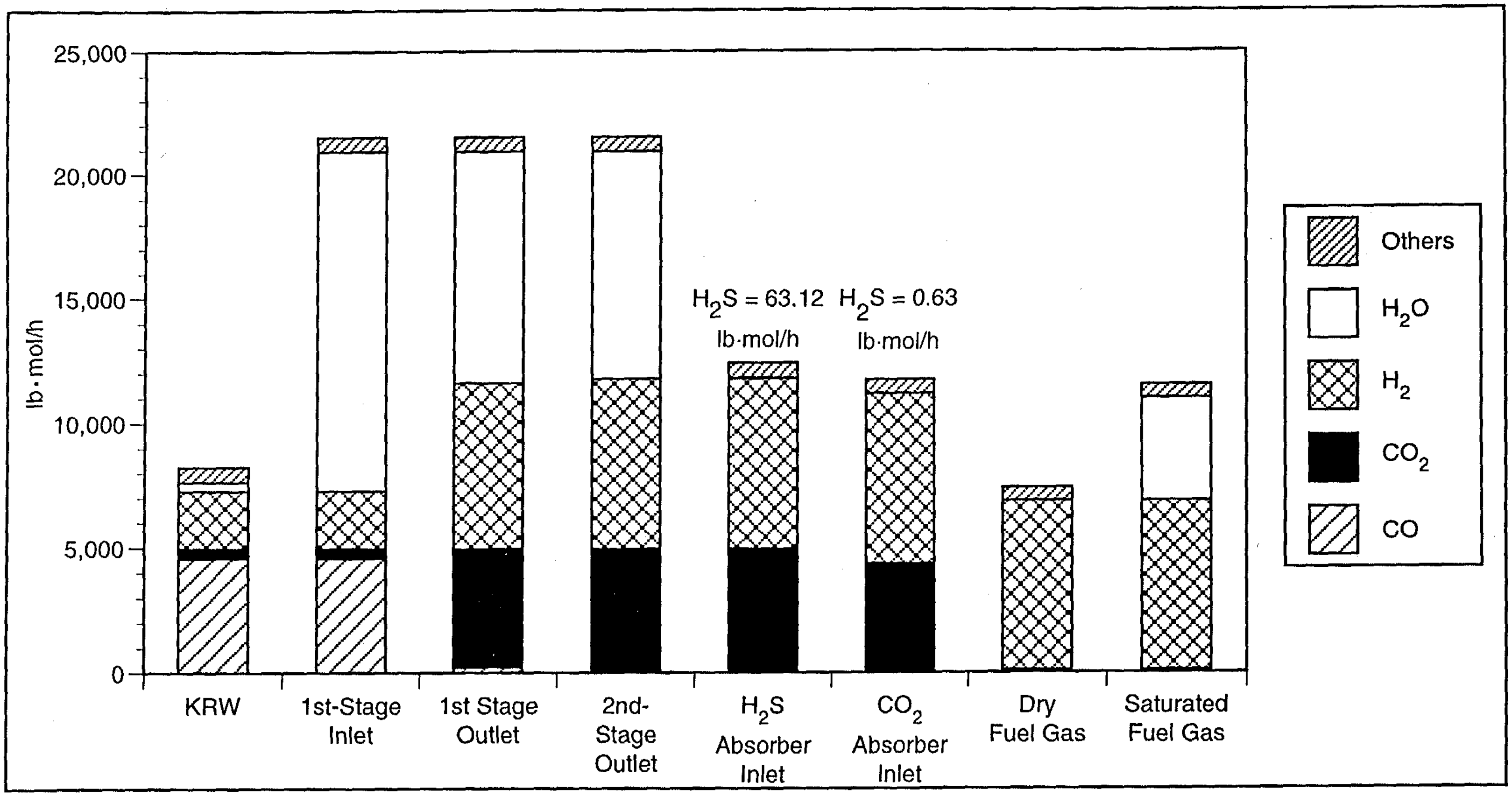

FIGURE 4.2 Gas Stream Composition at Various Stages in the Process in Case 1

TABLE 4.1 Material Flows for Oxygen-Blown Base Case and Case 1

Material Flow (tons/d)	Base Case	Case 1
Coal (prepared)	3,845	3,845
Limestone	0	0
Air	0	0
Oxygen	2,347	2,347
Solid waste	492	492
Sulfur	78	78
CO_2 (gasifier only)	8,586	898
SO_2 (gasifier only)	6.92	6.92
Net power output (MW)	413.5	377.47

accomplished with the initial cooling of the synthesis gas. The allocation of available enthalpy is summarized in Table 4.2. Details on the gas stream composition and the other streams shown in Figure 4.3 are provided in Table 4.3.

4.3 Glycol Process for CO_2 and H_2S Recovery

Of the several commercial options for CO_2 recovery investigated in ANL/ESD-24, the glycol process had the most favorable economics and the lowest energy penalty. The design analyzed here is based on a commercial version of the glycol process; it is called Selexol®. Lack of design data for this proprietary process makes system optimization to commercial standards impossible, but the key features of a commercial system are well-represented by this analysis. A glycol process has also been employed for H_2S recovery in the two gas turbine cases. Figure 4.4 is a flow diagram of the glycol process for H_2S removal. The material balances for the flows represented in that figure are summarized in Table 4.4. Key assumptions for these stream flow calculations are presented in Table 4.5. A similar set of exhibits defines the glycol system for CO_2 recovery. A significant difference between the two systems is the use of thermal stripping for solvent recovery in the H_2S case and flash recovery in the CO_2 case. Figure 4.5 shows the glycol recovery process for the CO_2. The stream flow data and stream calculation descriptions are summarized in Tables 4.6 and 4.7, respectively.

4.4 Gas Turbine, Steam Cycle, and Plant Performance

The application of CO_2 recovery by the glycol process results in a reduction in net plant output of 36 MW or 8.7% of the reference case plant output. Table 4.8 lists the gas turbine output, steam cycle output, and internal plant consumption for the base case (no CO_2 recovery) and for the glycol-based CO_2 recovery case. The most significant losses are a reduction in gas turbine output and the consumption of power for CO_2 compression.

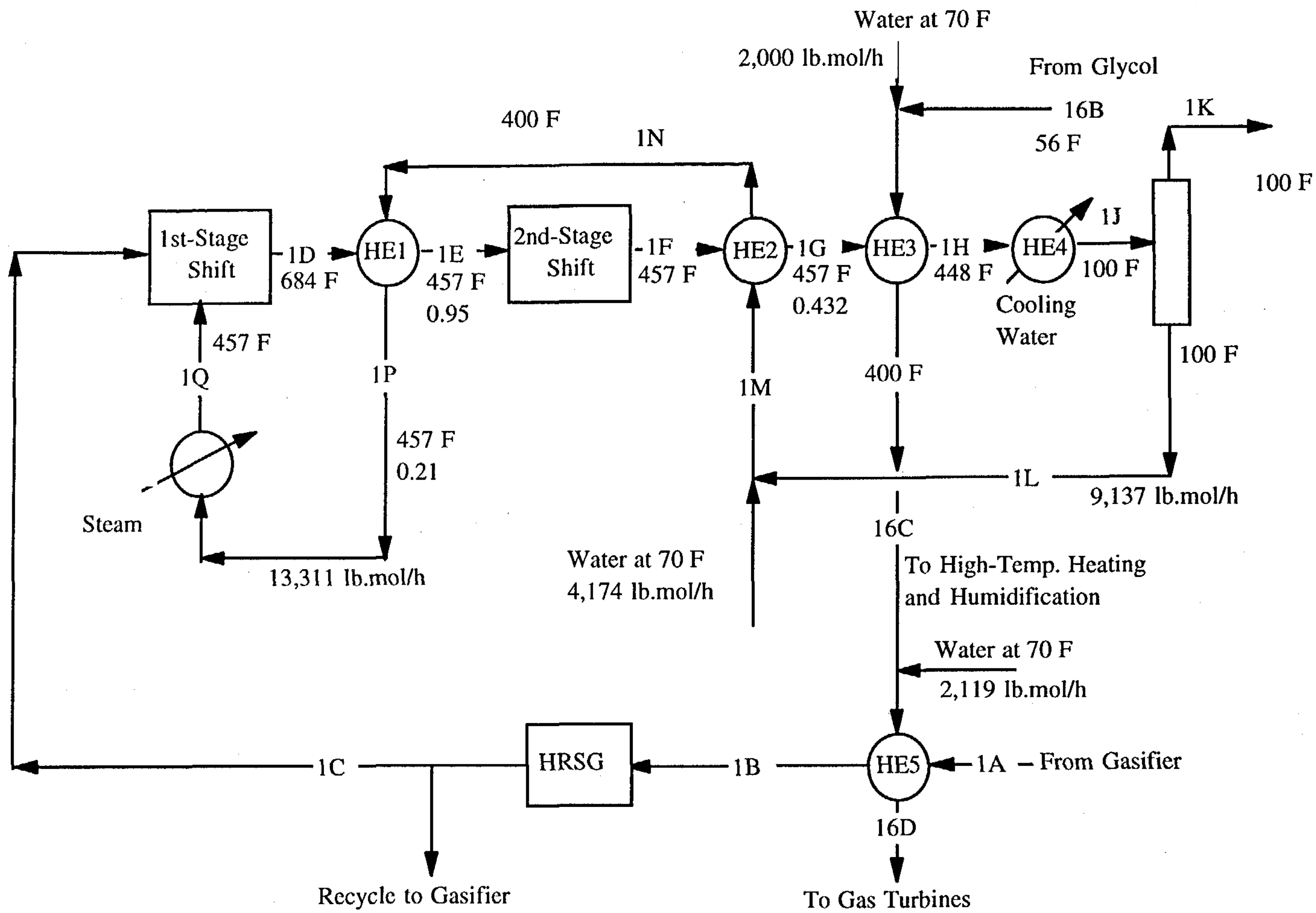

FIGURE 4.3 Flow Diagram of Shift System and Associated Heat Integration in Case 1

TABLE 4.2 Heat Recovery and Allocation (10^6 Btu/h) for Gas Turbine/Glycol Process in Case 1

Process	Enthalpy Change Available from Process	Allocation to Fuel Gas Preheating	Allocation for Raising Steam for Shift System	Allocation to Steam Cycle
Initial gas cooling to 460°F	513.89	344.28	123.89	45.71
Cooling after first-stage shift	168.21	0.00	168.21	0.00
Cooling after second-stage shift	673.27	177.41	215.65	280.22

TABLE 4.3 Stream Flows of Shift System of Gas Turbine/Glycol Process in Case 1

Stream Data	Stream 1A	Stream 1B	Stream 1C	Stream 1D	Stream 1E	Stream 1F
Description of stream	Raw gas from KRW gasifier	Raw gases after gas-gas heat exchanger	Raw gases to shift system	Raw gas from 1st-stage shift	Raw gases from heat exchanger 1	Raw gas from 2nd-stage shift
Gases (lb·mol/h)						
CO	8,887.28	8,887.28	4,558.89	227.94	227.94	45.59
CO_2	769.57	769.57	394.76	4,725.71	4,725.71	4,908.06
H_2	4,513.52	4,513.52	2,315.43	6,646.37	6,646.37	6,828.72
H_2O	711.96	711.96	365.22	9,345.71	9,345.71	9,163.36
N_2	71.03	71.03	36.43	36.43	36.43	36.43
Ar	141.76	141.76	72.72	72.72	72.72	72.72
CH_4	950.15	950.15	487.39	487.39	487.39	487.39
NH_3	36.61	36.61	18.79	18.79	18.79	18.79
H_2S	123.05	123.05	63.12	63.12	63.12	63.12
HCN	0.80	0.80	0.41	0.41	0.41	0.41
O_2	0.00	0.00	0.00	0.00	0.00	0.00
COS	14.49	14.49	7.43	7.43	7.43	7.43
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	16,220.23	16,220.23	8,320.59	21,632.03	21,632.03	21,632.03
Liquids (lb·mol/h)						
H_2O	0.00	0.00	0.00	0.00	0.00	0.00
Temperature (°F)	1,749.45	934.83	457.40	683.87	457.40	457.40
Pressure (psia)	465.00	465.00	457.00	457.00	457.00	457.00
Enthalpy of stream (Btu/h) (reference, 32°F)	240,514,724	125,753,677	32,874,692	295,099,586	239,029,336	242,258,675

TABLE 4.3 (Cont.)

Stream Data	Stream 1G	Stream 1H	Stream 1J	Stream 1K	Stream 1L	Stream 1M
Description of stream	Raw gases from heat exchanger 2	Raw gases from heat exchanger 3	Raw gases from heat exchanger 4	Raw gases to glycol system	Condensed water to shift system	Water to heat exchanger 2 for shift system
Gases (lb·mol/h)						
CO	45.59	45.59	45.59	45.59	0.00	0.00
CO_2	4,908.06	4,908.06	4,908.06	4,908.06	0.00	0.00
H_2	6,828.72	6,828.72	6,828.72	6,828.72	0.00	0.00
H_2O	9,163.36	9,163.36	9,163.36	26.29	0.00	0.00
N_2	36.43	36.43	36.43	36.43	0.00	0.00
Ar	72.72	72.72	72.72	72.72	0.00	0.00
CH_4	487.39	487.39	487.39	487.39	0.00	0.00
NH_3	18.79	18.79	18.79	18.79	0.00	0.00
H_2S	63.12	63.12	63.12	63.12	0.00	0.00
HCN	0.41	0.41	0.41	0.41	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00
COS	7.43	7.43	7.43	7.43	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	21,632.03	21,632.03	21,632.03	12,494.97	0.00	0.00
Liquids (lb·mol/h) H_2O	0.00	0.00	0.00	0.00	9,137.07	13,311.44
Temperature (°F)	457.00	448.00	100.00	100.00	100.00	100.00
Pressure (psia)	457.00	457.00	457.00	457.00	457.00	457.00
Enthalpy of stream (Btu/h) (reference, 32°F)	170,376,904	111,240,670	17,835,157	7,087,548	10,747,609	16,293,201

TABLE 4.3 (Cont.)

Stream Data	Stream 1N	Stream 1P	Stream 1Q	Stream 16B	Stream 16C	Stream 16D
Description of stream	Water from heat exchanger 2 for shift system	Water from heat exchanger 1 for shift system	Water to shift system	CO_2 lean gases from glycol system	CO_2 lean gases to gas-gas heat exchanger	CO_2 lean gases to gas turbines
Gases (lb·mol/h)						
CO	0.00	0.00	0.00	44.68	44.68	44.68
CO_2	0.00	0.00	0.00	43.10	43.10	43.10
H_2	0.00	0.00	0.00	6,773.42	6,773.42	6,773.42
H_2O	0.00	0.00	0.00	0.00	2,000.00	4,119.33
N_2	0.00	0.00	0.00	34.99	34.99	34.99
Ar	0.00	0.00	0.00	72.72	72.72	72.72
CH_4	0.00	0.00	0.00	439.87	439.87	439.87
NH_3	0.00	0.00	0.00	18.79	18.79	18.79
H_2S	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.41	0.41	0.41
O_2	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	1.19	1.19	1.19
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	0.00	0.00	0.00	7,429.18	9,429.18	11,548.52
Liquids (lb·mol/h) H_2O	13,311.44	13,311.44	13,311.44	0.00	0.00	0.00
Temperature (°F)	400.00	457.00	457.00	56.24	400.00	1,136.91 887.00
Pressure (psia)	457.00	150.00	457.00	232.00	232.00	232.00
Enthalpy of stream (Btu/h) (reference, 32°F)	88,174,973	144,245,158	185,542,089	1,262,022	62,846,191	537,171,354

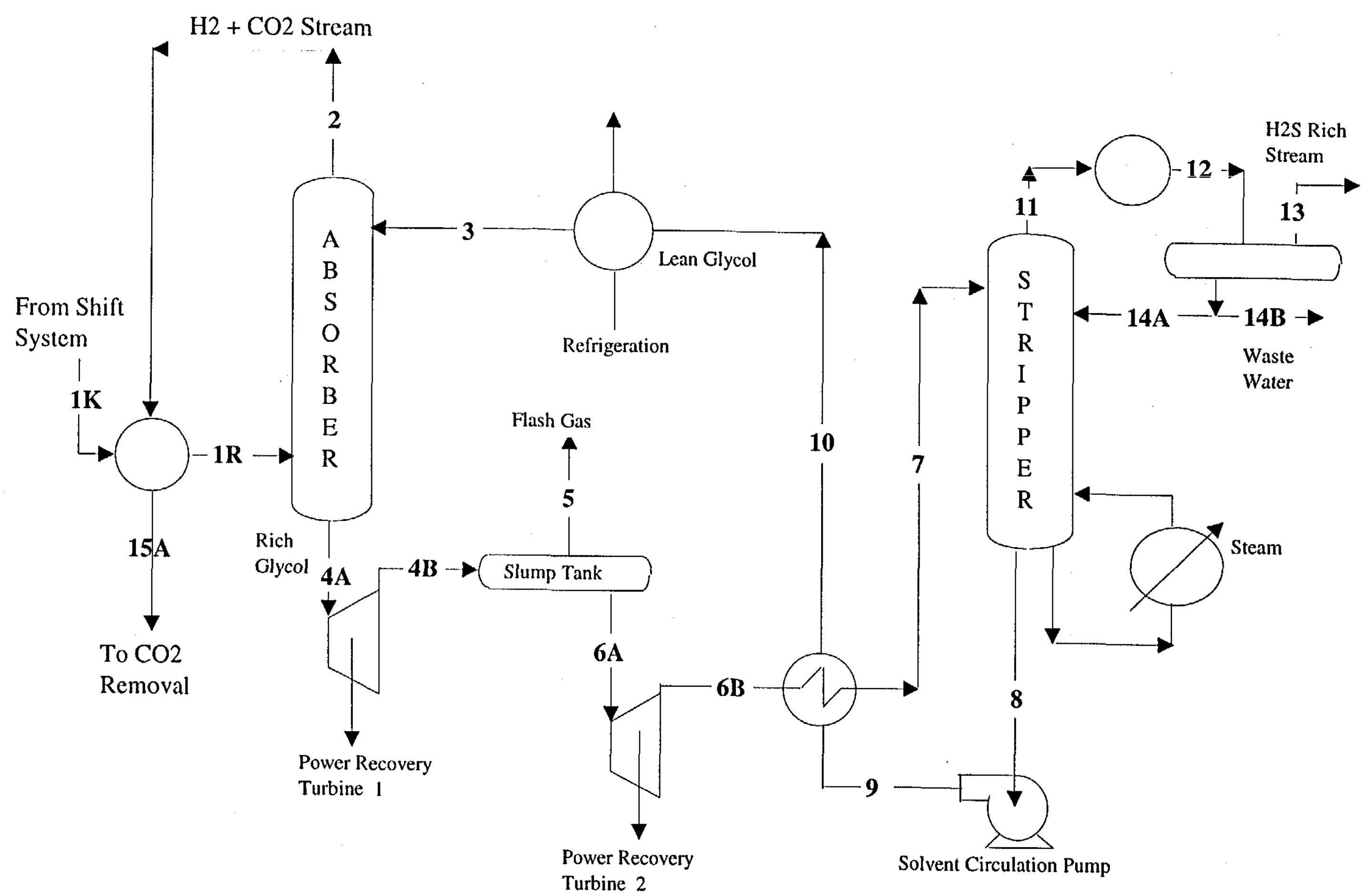

FIGURE 4.4 Flow Diagram of Glycol Process for H_2S Recovery in Case 1

TABLE 4.4 Stream Flows of Glycol Process for H_2S Removal in Case 1

Stream Data	Stream 1K	Stream 1R	Stream 2	Stream 3	Stream 4A	Stream 4B
Description of stream	Feed gas from shift system	Absorber feed	Sulfur-free gas from absorber	Lean glycol solvent	Rich glycol solvent from absorber	Rich glycol solvent after turbine 1
Gases (lb·mol/h)						
CO	45.59	45.59	45.13	0.00	0.46	0.46
CO_2	4,908.06	4,908.06	4,310.02	109.53	707.57	707.57
H_2	6,828.72	6,828.72	6,822.47	10.95	17.20	17.20
H_2O	26.29	26.29	0.00	1,022.29	1,048.58	1,048.58
N_2	36.43	36.43	35.71	0.00	0.73	0.73
Ar	72.72	72.72	72.72	0.00	0.00	0.00
CH_4	487.39	487.39	463.02	0.00	24.37	24.37
NH_3	18.79	18.79	18.79	0.00	0.00	0.00
H_2S	63.12	63.12	0.63	21.91	84.40	84.40
HCl	0.41	0.41	0.41	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00
COS	7.43	7.43	2.97	0.00	4.46	4.46
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	12,494.97	12,494.97	11,771.88	1,164.68	1,887.76	1,887.76
Liquids (lb·mol/h)						
Glycol solvent	0.00	0.00	0.00	2,190.62	2,190.62	2,190.62
Temperature (°F)	100.00	63.00	30.00	30.00	63.62	62.28
Pressure (psia)	451.00	451.00	446.00	451.00	446.00	100.00
Enthalpy (Btu/h) (reference, 32°F)	7,087,548	3,456,963	-180,609	-640,350	10,332,246	9,893,895

TABLE 4.4 (Cont.)

Stream Data	Stream 5	Stream 6A	Stream 6B	Stream 7	Stream 8	Stream 9
Description of stream	Flash gas	Rich glycol solvent to turbine 2	Rich glycol solvent from turbine 2	Rich glycol solvent after heat exchange	Lean glycol solvent from stripper	Lean glycol solvent after circulation pump
Gases (lb·mol/h)						
CO	0.41	0.05	0.05	0.05	0.00	0.00
CO_2	566.06	141.51	141.51	141.51	109.53	109.53
H_2	6.19	11.01	11.01	11.01	10.95	10.95
H_2O	10.49	1,038.10	1,038.10	1,038.10	1,022.29	1,022.29
N_2	0.66	0.07	0.07	0.07	0.00	0.00
Ar	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	20.71	3.66	3.66	3.66	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	4.22	80.18	80.18	80.18	21.91	21.91
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	2.23	2.23	2.23	2.23	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	610.96	1,276.80	1,276.80	1,276.80	1,164.68	1,164.68
Liquids (lb·mol/h)						
Glycol solvent	0.00	2,190.62	2,190.62	2,190.62	2,190.62	2,190.62
Temperature (°F)	42.44	42.44	42.10	190.00	212.00	215.21
Pressure (psia)	100.00	100.00	14.70	14.70	14.70	451.00
Enthalpy (Btu/h) (reference, 32°F)	589.41	3,354,029	3,245,651	50,770,201	57,640,407	58,667,747

TABLE 4.4 (Cont.)

Stream Data	Stream 10	Stream 11	Stream 12	Stream 13	Stream 14A	Stream 14B	Stream 15A
Description of stream	Lean glycol solvent after heat exchange	H_2S-rich gas from stripper	H_2S-rich gas after condenser	H_2S-rich gas	Recycle to stripper	Wastewater to disposal	Sulfur-free fuel gas after heat exchange
Gases (lb·mol/h)							
CO	0.00	0.05	0.05	0.05	0.00	0.00	45.13
CO_2	109.53	31.98	31.98	31.98	0.00	0.00	4,310.02
H_2	10.95	0.06	0.06	0.06	0.00	0.00	6,822.47
H_2O	1,022.29	1,038.10	1,038.10	4.92	1,022.29	10.88	0.00
N_2	0.00	0.07	0.07	0.07	0.00	0.00	35.71
Ar	0.00	0.00	0.00	0.00	0.00	0.00	72.72
CH_4	0.00	3.66	3.66	3.66	0.00	0.00	463.02
NH_3	0.00	0.00	0.00	0.00	0.00	0.00	18.79
H_2S	21.91	58.27	58.27	58.27	0.00	0.00	0.63
HCl	0.00	0.00	0.00	0.00	0.00	0.00	0.41
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	2.23	2.23	2.23	0.00	0.00	2.97
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	1,164.68	1,134.41	1,134.41	101.24	1,022.29	10.88	11,771.88
Liquids (lb·mol/h)							
Glycol solvent	2,190.62	0.00	0.00	0.00	0.00	0.00	0.00
Temperature (°F)	66.80	212.00	100.00	100.00	100.00	100.00	70.00
Pressure (psia)	14.70	14.70	14.70	14.70	14.70	14.70	446.00
Enthalpy (Btu/h) (reference, 32°F)	11,143,197	21,649,320	1,325,921	143,003	1,251,282	13,320	3,450,043

TABLE 4.5 Descriptions of Streams of Glycol Process for H_2S Removal in Case 1

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 1K: Synthesis gas from shift system		
Temperature (°F)	100	The synthesis gas is shifted to maximize the overall CO_2 recovery. After the shift, the gases are cooled to a temperature of 100°F.
Pressure (psia)	451	
Flow rate (lb·mol/h)	12,494.97	
CO_2 (mole fraction)	0.3928	
H_2S (mole fraction)	0.0051	
Stream 1R: Feed gas to absorber		
Temperature (°F)	63	The shifted gases are cooled against the sulfur-free gas from the absorber to a temperature of 63°F in order to decrease the solvent circulation rate.
Pressure (psia)	451	
Flow rate (lb·mol/h)	12,494.97	
CO_2 (mole fraction)	0.3928	
H_2S (mole fraction)	0.0051	
Stream 2: Sulfur-free gases from absorber		
Temperature (°F)	30	The composition of this stream corresponds to an H_2S-removal efficiency of 99%. Also, other gases like CO_2, COS, and H_2 are absorbed by the solvent. The temperature of this stream is close to the temperature of lean solvent entering the absorber at the top.
Pressure (psia)	446	
Flow rate (lb·mol/h)	11,771.88	
CO_2 (mole fraction)	0.3661	
H_2S (mole fraction)	0.0001	
Stream 3: Lean glycol solvent to absorber		
Temperature (°F)	30	Lean glycol solvent contains residual H_2S and CO_2. The solvent also contains 30% water. 100% excess solvent is used.
Pressure (psia)	451	
Flow rate (lb·mol/h)	3,355.30	
CO_2 (mole fraction)	0.0326	
H_2S (mole fraction)	0.0065	
Stream 4A: Rich glycol solvent from absorber		
Temperature (°F)	63.62	Flow rate reflects lean glycol solvent plus absorbed CO_2, H_2S, and other gases. The temperature rises because of the heat of absorption of CO_2 and H_2S.
Pressure (psia)	446	
Flow rate (lb·mol/h)	4,078.38	
CO_2 (mole fraction)	0.1735	
H_2S (mole fraction)	0.0207	

TABLE 4.5 (Cont.)

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 4B: Rich glycol solvent from turbine 1		
Temperature (°F)	62.68	This stream is exit stream from high-pressure power recovery turbine. Exit pressure has been selected to avoid release of H_2S and CO_2 while allowing some recovery of work of pressurization. The change in temperature over the turbine is estimated from change in enthalpy, which is taken to be equal to flow work.
Pressure (psia)	100	
Flow rate (lb·mol/h)	4,078.38	
CO_2 (mole fraction)	0.1735	
H_2S (mole fraction)	0.0207	
Stream 5: Flash gas		
Temperature (°F)	42.44	CO_2 and H_2S are released from the glycol solvent in the slump tank. This stream is not recycled to the absorber. The released gases contain mostly CO_2 (93%) and therefore can be disposed of.
Pressure (psia)	100	
Flow rate (lb·mol/h)	610.96	
CO_2 (mole fraction)	0.9265	
H_2S (mole fraction)	0.0007	
Stream 6A: Rich glycol solvent to low-pressure power recovery turbine		
Temperature (°F)	42.44	Change in composition simply reflects flashing of fuel gases to stream 5.
Pressure (psia)	100	
Flow rate (lb·mol/h)	3,467.42	
CO_2 (mole fraction)	0.0408	
H_2S (mole fraction)	0.0231	
Stream 6B: Rich glycol solvent from low-pressure power recovery turbine		
Temperature (°F)	42.10	This stream is exit stream from low-pressure turbine. The change in temperature is calculated as in 4B.
Pressure (psia)	14.7	
Flow rate (lb·mol/h)	3,467.42	
CO_2 (mole fraction)	0.0408	
H_2S (mole fraction)	0.0231	

TABLE 4.5 (Cont.)

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 7: Rich glycol solvent to stripper		
Temperature (°F)	190	Rich glycol solvent is heated from 42.1°F to 190°F in lean-rich solvent heat exchanger to decrease reboiler load.
Pressure (psia)	14.7	
Flow rate (lb·mol/h)	3,467.42	
CO_2 (mole fraction)	0.0408	
H_2S (mole fraction)	0.0231	
Stream 8: Lean glycol solvent from stripper		
Temperature (°F)	212	CO_2 and H_2S are stripped from the solvent by heat. Stripper is operated at a temperature of 212°F and a pressure of 14.7 psia.
Pressure (psia)	14.7	
Flow rate (lb·mol/h)	3,355.30	
CO_2 (mole fraction)	0.0326	
H_2S (mole fraction)	0.0065	
Stream 9: Lean glycol solvent from circulation pump		
Temperature (°F)	215.21	Lean glycol solvent from the stripper is at a pressure of 14.7 psia and is pressurized to absorber pressure of 451 psia by circulation pump. The slight increase in temperature is due to work of compression.
Pressure (psia)	451	
Flow rate (lb·mol/h)	3,355.30	
CO_2 (mole fraction)	0.0326	
H_2S (mole fraction)	0.0065	
Stream 10: Lean glycol solvent after lean-rich solvent heat exchanger		
Temperature (°F)	66.87	Lean solvent is cooled against rich solvent from the absorber to temperature of 67°F to decrease refrigeration load.
Pressure (psia)	451	
Flow rate (lb·mol/h)	3,355.30	
CO_2 (mole fraction)	0.0326	
H_2S (mole fraction)	0.0065	
Stream 11: H_2S-rich gas from stripper		
Temperature (°F)	212	The solubilities of gases decrease with temperature and therefore are released from the solvent. The composition of this stream represents amount of gases released and water evaporated.
Pressure (psia)	14.7	
Flow rate (lb·mol/h)	1,134.41	
CO_2 (mole fraction)	0.0282	
H_2S (mole fraction)	0.0514	

TABLE 4.5 (Cont.)

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 12: H_2S-rich gas after condenser		
Temperature (°F)	100	Mostly water is condensed in heat exchanger by using cooling water to a temperature of 100°F.
Pressure (psia)	14.7	
Flow rate (lb·mol/h)	1,134.41	
CO_2 (mole fraction)	0.0282	
H_2S (mole fraction)	0.0514	
Stream 13: H_2S-product stream		
Temperature (°F)	100	The gases are separated in the phase separator. The gases are sent to Claus plant for further treatment.
Pressure (psia)	14.7	
Flow rate (lb·mol/h)	101.24	
CO_2 (mole fraction)	0.3159	
H_2S (mole fraction)	0.5756	
Stream 14A: Recycle to stripper		
Temperature (°F)	100	To maintain low partial pressures of H_2S and CO_2, condensed water is recycled to the stripper. This also maintains the water balance in the solvent.
Pressure (psia)	14.7	
Flow rate (lb·mol/h)	1,022.29	
CO_2 (mole fraction)	0.0000	
H_2S (mole fraction)	0.0000	
Stream 14B: Wastewater for treatment		
Temperature (°F)	100	Excess water is removed through this stream.
Pressure (psia)	14.7	
Flow rate (lb·mol/h)	10.88	
CO_2 (mole fraction)	0.0000	
H_2S (mole fraction)	0.0000	
Stream 15A: Sulfur-free fuel gas after heat exchange		
Temperature (°F)	70	The fuel gas from the absorber is heated against the feed to the absorber. These gases are further treated in CO_2-removal section.
Pressure (psia)	446	
Flow rate (lb·mol/h)	11,771.88	
CO_2 (mole fraction)	0.3661	
H_2S (mole fraction)	0.0001	

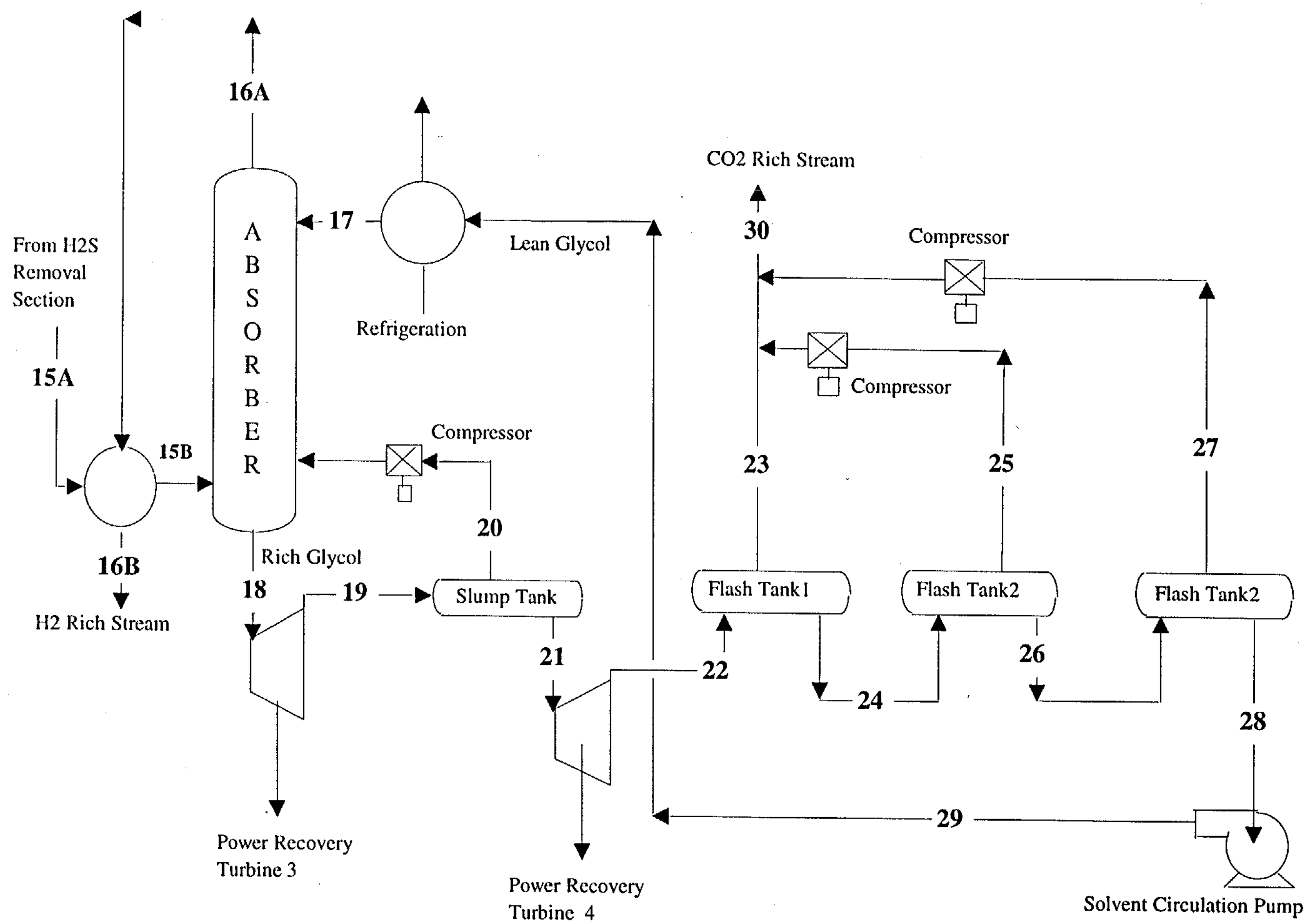

FIGURE 4.5 Flow Diagram of Glycol Process for CO_2 Recovery in Case 1

TABLE 4.6 Stream Flows of Glycol Process for CO_2 Removal in Case 1

Stream Data	Stream 15A	Stream 15B	Stream 16A	Stream 16B	Stream 17	Stream 18	Stream 15A
Description of stream	Sulfur-free feed gas from H_2S removal section	Absorber feed	Fuel gas from absorber	Fuel gas after heat exchanger	Lean glycol solvent	Rich glycol solvent from absorber	Sulfur-free fuel gas after heat exchange
Gases (lb·mol/h)							
CO	45.13	45.13	44.68	44.68	0.00	4.51	45.13
CO_2	4,310.02	4,310.02	43.10	43.10	118.16	4,474.57	4,310.02
H_2	6,822.47	6,822.47	6,773.42	6,773.42	59.08	1,081.28	6,822.47
H_2O	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N_2	35.71	35.71	34.99	34.99	0.00	7.14	35.71
Ar	72.72	72.72	72.72	72.72	0.00	0.00	72.72
CH_4	463.02	463.02	439.87	439.87	0.00	154.34	463.02
NH_3	18.79	18.79	18.79	18.79	0.00	0.00	18.79
H_2S	0.63	0.63	0.00	0.00	7.38	8.09	0.63
HCl	0.41	0.41	0.41	0.41	0.00	0.00	0.41
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
COS	2.97	2.97	1.19	1.19	0.00	3.57	2.97
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	11,771.88	11,771.88	7,429.18	7,429.18	184.62	5,733.51	11,771.88
Liquids (lb·mol/h)							
Glycol solvent	0.00	0.00	0.00	0.00	11,815.53	11,815.53	0.00
Temperature (°F)	70	55.00	30.00	56.24	30.00	61.97	70.00
Pressure (psia)	446	446.00	441.00	441.00	446.00	441.00	446.00
Enthalpy (Btu/h) (reference, 32°F)	3,450,043	2,083,999	-103,990	1,262,022	-3,245,210	50,053,110	3,450,043

TABLE 4.6 (Cont.)

Stream Data	Stream 19	Stream 20	Stream 21	Stream 22	Stream 23	Stream 24
Description of stream	Rich glycol solvent after turbine 3	Recycle to absorber	Rich glycol solvent to turbine 4	Rich glycol solvent after turbine 4	CO_2-rich gas from 1st flash	Rich glycol solvent to 2nd flash
Gases (lb·mol/h)						
CO	4.51	4.06	0.45	0.45	0.45	0.00
CO_2	4,474.57	89.49	4,385.08	4,385.08	3,288.81	1,096.27
H_2	1,081.28	973.16	108.13	108.13	32.44	75.69
H_2O	0.00	0.00	0.00	0.00	0.00	0.00
N_2	7.14	6.43	0.71	0.71	0.71	0.00
Ar	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	154.34	131.19	23.15	23.15	18.52	4.63
NH_3	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	8.09	0.08	8.01	8.01	0.08	7.93
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	3.57	1.78	1.78	1.78	0.45	1.34
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	5,733.51	1,206.19	4,527.32	4,527.32	3,341.46	1,185.86
Liquids (lb·mol/h)						
Glycol solvent	11,815.53	0.00	11,815.53	11,815.53	0.00	11,815.53
Temperature (°F)	60.99	60.38	60.38	59.76	37.26	37.26
Pressure (psia)	200.00	200.00	200.00	50.00	50.00	50.00
Enthalpy (Btu/h) (reference, 32°F)	48,408,324	246,412	47,144,244	46,116,263	155,492	8,587,255

TABLE 4.6 (Cont.)

Stream Data	Stream 25	Stream 26	Stream 27	Stream 28	Stream 29	Stream 30
Description of stream	CO_2-rich gas from 2nd flash	Rich glycol solvent to 3rd flash	CO_2-rich gas from 3rd flash	Lean glycol solvent	Lean glycol solvent after circulation pump	CO_2-rich product
Gases (lb·mol/h)						
CO	0.00	0.00	0.00	0.00	0.00	0.45
CO_2	767.39	328.88	210.47	118.41	118.41	4,266.66
H_2	15.14	60.55	1.37	59.18	59.18	48.94
H_2O	0.00	0.00	0.00	0.00	0.00	0.00
N_2	0.00	0.00	0.00	0.00	0.00	0.71
Ar	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	4.63	0.00	0.00	0.00	0.00	23.15
NH_3	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	0.40	7.53	0.14	7.39	7.39	0.62
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	1.34	0.00	0.00	0.00	0.00	1.78
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	788.89	396.97	211.98	184.99	184.99	4,342.33
Liquids (lb·mol/h)						
Glycol solvent	0.00	11,815.53	0.00	11,815.53	11,815.53	0.00
Temperature (°F)	31.92	31.92	30.44	30.44	33.90	81.53
Pressure (psia)	14.70	14.70	4.00	4.00	446.00	50.00
Enthalpy (Btu/h) (reference, 32°F)	-580	-135,628	-2,916	-2,525,525	3,075,331	1,922,357

TABLE 4.7 Descriptions of Streams of Glycol Process for CO_2 Removal in Case 1

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 15A: Sulfur-free gas from H_2S section		
Temperature (°F)	70	The synthesis gas is cleaned in two stages. First sulfur compounds are removed. Then they are fed to another absorption column for CO_2 recovery.
Pressure (psia)	446	
Flow rate (lb·mol/h)	11,771.88	
CO_2 (mole fraction)	0.3661	
H_2S (mole fraction)	0.0001	
Stream 15B: Feed gas to absorber		
Temperature (°F)	55	The sulfur-free synthesis gas is cooled against the cold fuel gas from top of the absorber to a temperature of 55°F.
Pressure (psia)	446	
Flow rate (lb·mol/h)	11,771.88	
CO_2 (mole fraction)	0.3661	
H_2S (mole fraction)	0.0001	
Stream 16A: Fuel gas from absorber		
Temperature (°F)	30	The composition of this stream corresponds to a CO_2-removal efficiency of 99%. Also, other gases like H_2S, COS, and H_2 are absorbed by the solvent. The temperature of this stream is close to the temperature of lean solvent entering the absorber at the top.
Pressure (psia)	441	
Flow rate (lb·mol/h)	7,429.18	
CO_2 (mole fraction)	0.0058	
H_2S (mole fraction)	0.0000	
Stream 16B: Fuel gas after heat exchanger		
Temperature (°F)	56.24	Fuel gas is heated against the sulfur-free gases from H_2S section.
Pressure (psia)	441	
Flow rate (lb·mol/h)	7,429.18	
CO_2 (mole fraction)	0.0058	
H_2S (mole fraction)	0.0000	
Stream 17: Lean glycol to the of absorber		
Temperature (°F)	30	Lean glycol solvent contains residual CO_2 and H_2S. 50% excess solvent is used. The solvent is cooled to 30°F by refrigeration.
Pressure (psia)	446	
Flow rate (lb·mol/h)	12,000.15	
CO_2 (mole fraction)	0.0098	
H_2S (mole fraction)	0.0006	

TABLE 4.7 (Cont.)

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 18: Rich glycol solvent from absorber		
Temperature (°F)	61.97	Flow rate reflects lean glycol solvent plus absorbed CO_2, H_2S, and other gases. The temperature rises because of the heat of absorption of CO_2 and H_2S.
Pressure (psia)	441	
Flow rate (lb·mol/h)	17,549.04	
CO_2 (mole fraction)	0.2550	
H_2S (mole fraction)	0.0005	
Stream 19: Rich glycol solvent from turbine 3		
Temperature (°F)	60.99	This stream is exit stream from high-pressure power recovery turbine. Exit pressure has been selected to avoid release of CO_2 and H_2S while allowing some recovery of work of pressurization. The change in temperature over the turbine is estimated from change in enthalpy, which is taken to be equal to flow work.
Pressure (psia)	200	
Flow rate (lb·mol/h)	17,549.04	
CO_2 (mole fraction)	0.2550	
H_2S (mole fraction)	0.0005	
Stream 20: Flash gas		
Temperature (°F)	60.38	CO_2 and H_2S are released from the glycol solvent in the slump tank. This stream is compressed and recycled to the absorber to decrease the losses of valuable gases like H_2 and CO.
Pressure (psia)	200	
Flow rate (lb·mol/h)	1,206.19	
CO_2 (mole fraction)	0.0742	
H_2S (mole fraction)	0.0001	
Stream 21: Rich glycol solvent to low-pressure power recovery turbine		
Temperature (°F)	60.38	Change in composition simply reflects flashing of fuel gases to stream 20.
Pressure (psia)	200	
Flow rate (lb·mol/h)	16,342.85	
CO_2 (mole fraction)	0.2683	
H_2S (mole fraction)	0.0005	
Stream 22: Rich glycol solvent from low-pressure power recovery turbine		
Temperature (°F)	59.76	This stream is exit from low-pressure turbine. The change in temperature is calculated as in stream 19.
Pressure (psia)	50	
Flow rate (lb·mol/h)	16,342.85	
CO_2 (mole fraction)	0.2683	
H_2S (mole fraction)	0.0005	

TABLE 4.7 (Cont.)

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 23: CO_2-rich flash gas from high-pressure flash tank		
Temperature (°F)	37.26	The CO_2 from the rich glycol solvent is released in stages. In the first stage, the gases are flashed to a pressure of 50 psia. The amount of CO_2 remaining in the solvent depends on pressure, and the CO_2 released is calculated by mass balance.
Pressure (psia)	50	
Flow rate (lb·mol/h)	3,341.46	
CO_2 (mole fraction)	0.9842	
H_2S (mole fraction)	0.0000	
Stream 24: Glycol solvent from high-pressure flash tank		
Temperature (°F)	37.26	-----
Pressure (psia)	50	
Flow rate (lb·mol/h)	13,001.39	
CO_2 (mole fraction)	0.0843	
H_2S (mole fraction)	0.0006	
Stream 25: CO_2-rich flash gas from intermediate-pressure flash tank		
Temperature (°F)	31.92	The amount of CO_2 in solvent and released as gas is calculated as in stream 23. Sufficient residence is provided for the gases to separate from solvent.
Pressure (psia)	17.70	
Flow rate (lb·mol/h)	788.89	
CO_2 (mole fraction)	0.9727	
H_2S (mole fraction)	0.0005	
Stream 26: Glycol solvent from intermediate-pressure flash tank		
Temperature (°F)	31.92	----
Pressure (psia)	14.7	
Flow rate (lb·mol/h)	12,212.50	
CO_2 (mole fraction)	0.0269	
H_2S (mole fraction)	0.0006	
Stream 27: CO_2-rich flash gas from low-pressure flash tank		
Temperature (°F)	30.44	Glycol solvent is flashed to a pressure of 4 psia to remove as much CO_2 as possible. The lower residual amount of CO_2 in lean glycol solvent reduces the circulation rate of solvent.
Pressure (psia)	4.0	
Flow rate (lb·mol/h)	211.98	
CO_2 (mole fraction)	0.9929	
H_2S (mole fraction)	0.0007	

TABLE 4.7 (Cont.)

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 28: Lean glycol solvent from low-pressure flash tank		
Temperature (°F)	30.44	----
Pressure (psia)	4.0	
Flow rate (lb·mol/h)	12,000.52	
CO_2 (mole fraction)	0.0098	
H_2S (mole fraction)	0.0006	
Stream 29: Lean glycol solvent after circulation pump		
Temperature (°F)	33.90	The lean solvent is pressurized to the absorber operating pressure by using a pump. The change in temperature is due to work of compression. The solvent is chilled before being sent to the absorber.
Pressure (psia)	446	
Flow rate (lb·mol/h)	12,000.52	
CO_2 (mole fraction)	0.0098	
H_2S (mole fraction)	0.0006	
Stream 30: CO_2-rich product gas		
Temperature (°F)	81.53	Flash gases from intermediate- and low-pressure flash tanks are compressed to the pressure of stream 23. Streams 23, 25, and 27 are combined for further compression for pipeline.
Pressure (psia)	50.0	
Flow rate (lb·mol/h)	4,342.33	
CO_2 (mole fraction)	0.9826	
H_2S (mole fraction)	0.0001	

TABLE 4.8 Power Output, Plant Power Use, and Net Power Output for Base Case and Case 1 Gas Turbine/Glycol Process

	Power (MW)	
Power Variable	Base Case	Glycol Case
Power output		
Gas turbine	298.8	284.8
Steam turbine	159.4	161.6
Internal power consumption		
CO_2 recovery		
CO_2 compression	0	-17.3
Solvent circulation	0	-5.8
Solvent refrigeration	0	-4.5
Power recovery turbine	0	3.4
Gasification system	-44.7	-44.7
Net power output	413.5	377.5
Energy penalty	0	36

4.5 Economics

Details of the direct capital investment estimates for the H_2S recovery system, the shift system, and the CO_2 recovery system are presented in Tables 4.9, 4.10, and 4.11, respectively. Total cost information, including indirect capital investment and operating and maintenance costs, is provided in Section 9.

TABLE 4.9 Sizing and Cost Estimation for Major Equipment Used for H_2S Removal in Glycol Process in Case 1

1.	**Heat Exchanger before the Absorption Column**		
	Q = Load (Btu/h)	3,630,585	
	Tha = Inlet temperature of hot fluid (°F)	100	
	Thb = Outlet temperature of hot fluid (°F)	63	
	Pressure of hot gases (psia)	451	
	Tca = Inlet temperature of cold fluid (°F)	30	
	Tcb = Outlet temperature of cold fluid (°F)	70	
	Delta T1	30	
	Delta T2	33	
	Log mean temperature difference (°F)	31	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	5	
	Heat transfer area (ft^2)	23,070	
	Operating Pressure (psia)	451	
	Pressure factor	1.175	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$185,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$812,765**
2.	**H_2S Absorption Column**		
	Diameter of tower (ft)	8	
	HETP (ft)	3	
	No. of theoretical stages	12	
	Absorber tower height (ft) (4 ft for inlet, outlet and gas, and liquid distributors)	40	
	Volume of packing (ft^3)	1,810	
	Pressure factor	2.6	
	Cost per foot of column height (mild steel construction)	$1,000	
	Materials correction factor	1	
	Module factor	4.16	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of absorber in 1995		**$505,513**
	Cost of packing per cubic foot (2-in. pall rings-metal)	$63.5	
	Total cost of packing		**$114,953**

TABLE 4.9 (Cont.)

3.	**Power Recovery Turbine 1**		
	Turbine size (hp)	173	
	Purchased cost in 1987	$120,000	
	Module factor	1	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of solvent pump in 1995		**$175,266**
4.	**Slump Tank**		
	Glycol solvent flow rate (lb/h)	613,374	
	Density of glycol solvent (lb/gal)	8.6	
	Residence time (s)	180	
	Slump tank volume (gal)	3,566	
	Pressure factor	1	
	Materials correction factor	1	
	Module factor	2.08	
	Purchased cost of slump tank in 1987 (mild steel construction)	$13,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of slump tank in 1995		**$31,595**
5.	**Power Recovery Turbine 2**		
	Turbine size (hp)	43	
	Purchased cost in 1987	$65,000	
	Module factor	1	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of solvent pump in 1995		**$75,948**
6.	**Solvent Circulation Pump**		
	Horsepower	403	
	Purchased cost of pump in 1987 (includes motor, coupling, base; cast iron, horizontal)	$30,000	
	Materials correction factor	1	
	Module factor	1.5	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of solvent pump in 1995		**$52,580**

TABLE 4.9 (Cont.)

7.	**Lean-Rich Solvent Heat Exchanger**		
	Q = Load (Btu/h)	47,724.550	
	Tha = Inlet temperature of hot fluid (°F)	215.21	
	Thb = Outlet temperature of hot fluid (°F)	67	
	Pressure of hot gases (psia)	450	
	Tca = Inlet temperature of cold fluid (°F)	42.10	
	Tcb = Outlet temperature of cold fluid (°F)	190.00	
	Delta T1	25.2077	
	Delta T2	25	
	Log mean temperature difference (°F)	25	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	150	
	Heat transfer area (ft^2)	12,697	
	Operating pressure (psia)	50	
	Pressure factor	1	
	Materials correction factor	1	
	Module factor	3.2	
	(includes all of the supporting equipment and connections and		
	Purchased cost of heat exchanger in 1987	$120,000	
	(mild steel construction; shell and tube floating head)		
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$448,680**
8.	**Stripping Column**		
	Diameter of tower (ft)	10	
	HETP (ft)	3	
	No. of theoretical stages	12	
	Absorber tower height	40	
	(4 ft for inlet, outlet and gas, and liquid distributors)		
	Volume of packing (ft^3)	2,829	
	Pressure factor	1	
	Materials correction factor (stainless steel 304)	1.7	
	Cost per foot of column height	$1,200	
	(mild steel construction)		
	Module factor	4.16	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of absorber in 1995		**$396,633**
	Cost of packing per cubic foot	$63.5	
	(2-in. pall rings-SS)		
	Materials correction factor	1	
	Total cost of packing		**$179,614**

TABLE 4.9 (Cont.)

9.	**Overhead Condenser**		
	Q = Load (Btu/h)	20,323,399	
	Tha = Inlet temperature of hot fluid (°F)	212.00	
	Thb = Outlet temperature of hot fluid (°F)	100	
	Pressure of hot gases (psia)	14.7	
	Tca = Inlet temperature of cold fluid (°F)	70.00	
	Tcb = Outlet temperature of cold fluid (°F)	180.00	
	Delta T1	32	
	Delta T2	30	
	Log mean temperature difference (°F)	31	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	40	
	Heat transfer area (ft^2)	16,396	
	Operating Pressure (psia)	14.7	
	Pressure factor	1	
	Materials correction factor 1 for SS	2.7	
	Materials correction factor 2 for SS	0.07	
	Materials correction factor	3.48	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$160,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$2,079,839**
10.	**Phase Separator**		
	Flow rate (lb/h)	22,291	
	Density of fluid (lb/gal)	0.04	
	Residence time (s)	120	
	phase separator volume (gal)	18,576	
	Pressure factor	1	
	Materials correction factor (stainless steel)	1.8	
	Module factor	2.08	
	Purchased cost of phase separator in 1987 (mild steel construction)	$44,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of phase separator in 1995		**$192,484**

TABLE 4.9 (Cont.)

11.	**Solvent Refrigeration**		
	Refrigeration (tons)	981.96	
	Purchased cost in 1987	$400,000	
	Temperature correction factor	1.25	
	Module factor	1.46	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of refrigeration in 1995		**$852,959**
	Total Direct Cost		**$5,918,829**
	Total Direct Cost for Three Trains		**$17,756,488**

TABLE 4.10 Sizing and Cost Estimation for Major Equipment Used for Shift System in Case 1

1.	**First-Stage Shift Reactor**		
	Catalyst volume (ft^3)	665	
	Reactor volume (ft^3) (1.2 times the catalyst volume)	798	
	Reactor volume (gal)	5,969	
	Pressure factor	2.8	
	Module factor	3.05	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Purchased cost of reactor in 1987	$7,000	
	Installed cost of reactor in 1995		**$69,849**
2.	**Second-Stage Shift Reactor**		
	Catalyst volume (ft^3)	285	
	Reactor volume (ft^3) (1.2 times the catalyst volume)	342	
	Reactor volume (gal)	2,558	
	Pressure factor	2.8	
	Module factor	3.05	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Purchased cost of reactor in 1987	$5,000	
	Installed cost of reactor in 1995		**$49,892**
3.	**Shift Catalyst**		
	Volume of catalyst in first stage (ft^3)	999	
	Volume of catalyst in second stage (ft^3)	339	
	Cost of high-temperature catalyst per cubic foot	$50	
	Cost of low-temperature catalyst per cubic foot	$250	
	Total cost of catalyst		**$134,647**

TABLE 4.10 (Cont.)

4.	**Heat Exchanger between First- and Second-Shift Stages**		
	Q = Load (Btu/h)	56,070,250	
	Tha = Inlet temperature of hot fluid (°F)	684	
	Thb = Outlet temperature of hot fluid (°F)	457	
	Pressure of hot gases (psia)	451	
	Tca = Inlet temperature of cold fluid (°F)	400	
	Tcb = Outlet temperature of cold fluid (°F)	457	
	Delta T1	226	
	Delta T2	57	
	Log mean temperature difference (°F)	123	
	Overall heat transfer coefficient ($Btu/h/ft^2/°F$)	40	
	Heat transfer area (ft^2)	11,383	
	Operating pressure (psia)	451	
	Pressure factor	1.175	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$120,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$527,199**
5.	**Heat Exchanger after Second-Stage Shift for Raising Steam**		
	Q = Load (Btu/h)	71,881,771	
	Tha = Inlet temperature of hot fluid (°F)	457	
	Thb = Outlet temperature of hot fluid (°F)	457	
	Pressure of hot gases (psia)	451	
	Tca = Inlet temperature of cold fluid (°F)	100	
	Tcb = Outlet temperature of cold fluid (°F)	400	
	Delta T1	57	
	Delta T2	357	
	Log mean temperature difference (°F)	164	
	Overall heat transfer coefficient ($Btu/h/ft^2/°F$)	40	
	Heat transfer area (ft^2)	10,955	
	Operating pressure (psia)	451	
	Pressure factor	1.175	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$118,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$518,412**

TABLE 4.10 (Cont.)

6. Heat Exchanger after Second-Stage for Heating Fuel Gas		
Q = Load (Btu/h)	59,136,171	
Tha = Inlet temperature of hot fluid (°F)	457	
Thb = Outlet temperature of hot fluid (°F)	457	
Pressure of hot gases (psia)	451	
Tca = Inlet temperature of cold fluid (°F)	56	
Tcb = Outlet temperature of cold fluid (°F)	400	
Delta T1	57	
Delta T2	401	
Log mean temperature difference (°F)	177	
Overall heat transfer coefficient (Btu/h/ft^2/°F)	5	
Heat transfer area (ft^2)	66,897	
Operating pressure (psia)	451	
Pressure factor	1.175	
Materials correction factor	1	
Module factor (includes all of the supporting equipment and connections and installation)	3.2	
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$400,000	
CE index for process equipment in 1987	320	
CE index for process equipment in 1995	373.9	
Installed cost of heat exchanger in 1995		**$1,757,330**
7. Heat Exchanger for Heating Clean Fuel Gas with Raw Gases from Gasifier		
Q = Load (Btu/h)	344,284,466	
Tha = Inlet temperature of hot fluid (°F)	1,750	
Thb = Outlet temperature of hot fluid (°F)	935	
Pressure of hot gases (psia)	451	
Tca = Inlet temperature of cold fluid (°F)	400	
Tcb = Outlet temperature of cold fluid (°F)	1,137	
Delta T1	613	
Delta T2	535	
Log mean temperature difference (°F)	573	
Overall heat transfer coefficient (Btu/h/ft^2/°F)	5	
Heat transfer area (ft^2)	120,154	
Operating pressure (psia)	451	
Pressure factor	1.175	
Materials correction factor	1	
Module factor (includes all of the supporting equipment and connections and installation)	3.2	
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$600,770	
CE index for process equipment in 1987	320	
CE index for process equipment in 1995	373.9	
Installed cost of heat exchanger in 1995		**$2,639,377**

TABLE 4.10 (Cont.)

8.	**Heat Exchanger for Cooling Shifted Synthesis Gas with Feedwater**		
	Q = Load (Btu/h)	93,405,576	
	Tha = Inlet temperature of hot fluid (°F)	449	
	Thb = Outlet temperature of hot fluid (°F)	100	
	Pressure of hot gases (psia)	451	
	Tca = Inlet temperature of cold fluid (°F)	70	
	Tcb = Outlet temperature of cold fluid (°F)	400	
	Delta T1	49	
	Delta T2	30	
	Log mean temperature difference (°F)	39	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	50	
	Heat transfer area (ft^2)	48,331	
	Operating pressure (psia)	451	
	Pressure factor	1.175	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$340,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$1,493,731**
	Total Direct Cost		**$7,190,437**
	Total Direct Cost for Three Trains		**$21,571,310**

TABLE 4.11 Sizing and Cost Estimation for Major Equipment Used for CO_2 Removal in Glycol Process in Case 1

1.	**Gas - Gas Heat Exchanger**		
	Q = Load (Btu/h)	1,366,044	
	Tha = Inlet temperature of hot fluid (°F)	70.00	
	Thb = Outlet temperature of hot fluid (°F)	55	
	Pressure of hot gases (psia)	450	
	Tca = Inlet temperature of cold fluid (°F)	30.00	
	Tcb = Outlet temperature of cold fluid (°F)	56.24	
	Delta T1	13.7558	
	Delta T2	25	
	Log mean temperature difference (°F)	19	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	5	
	Heat transfer area (ft^2)	14,516	
	Operating Pressure (psia)	50	
	Pressure factor	1.175	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$150,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$658,999**
2.	**CO_2 Absorption Column**		
	Diameter of tower (ft)	12	
	HETP (ft)	3	
	No. of theoretical stages	12	
	Absorber tower height (ft) (4 ft for inlet, outlet and gas, and liquid distributors)	40	
	Volume of packing (ft^3)	4,073	
	Pressure factor	1	
	Cost per foot of column height (mild steel construction)	$1,400	
	Materials correction factor	1	
	Module factor	4.16	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of absorber in 1995		**$272,199**
	Cost of packing per cubic foot (2-in. pall rings-metal)	$63.5	
	Total cost of packing		**$258,645**

TABLE 4.11 (Cont.)

3.	**Power Recovery Turbine 1**		
	Turbine size (hp)	649	
	Purchased cost in 1987	$200,000	
	Module factor	1	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of solvent pump in 1995		**$233,688**
4.	**Slump Tank**		
	Glycol solvent flow rate (lb/h)	3,308,349	
	Density of glycol solvent (lb/gal)	8.6	
	Residence time (s)	180	
	Slump tank volume (gal)	19,235	
	Pressure factor	1.38	
	Materials correction factor	1	
	Module factor	2.08	
	Purchased cost of slump tank in 1987 (mild steel construction)	$45,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of slump tank in 1995		**$150,925**
5.	**Recycle Compressor**		
	Inlet pressure (psia)	200	
	Outlet pressure (psia)	446.00	
	Compressor size (hp)	537	
	Purchased cost of reciprocating compressor in 1987 (includes electric motor drive and gear reducer)	$160,000	
	Size factor for compressor	1	
	Materials correction factor	1	
	Module factor	2.6	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of compressor in 1995		**$486,070**
6.	**Power Recovery Turbine 2**		
	Turbine size (hp)	404	
	Purchased cost in 1987	$170,000	
	Module factor	1	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of solvent pump in 1995		**$198,634**

TABLE 4.11 (Cont.)

7.	**Flash Tank 1**		
	Glycol flow rate (lb/h)	3,308,349	
	Density of glycol (lb/gal)	8.6	
	Residence time (s)	180	
	Flash tank volume (gal)	19,235	
	Pressure factor	1	
	Module factor	2.08	
	Purchased cost of flash tank 1987 (mild steel construction)	$45,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of flash tank in 1995		**$109,366**
8.	**Flash Tank 2**		
	Glycol flow rate (lb/h)	3,308,349	
	Density of glycol (lb/gal)	8.6	
	Residence time (s)	180	
	Flash tank volume (gal)	19,235	
	Pressure factor	1	
	Module factor	2.08	
	Purchased cost of flash tank 1987 (mild steel construction)	$45,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of flash tank in 1995		**$109,366**
9.	**Flash Tank 3**		
	Glycol flow rate (lb/h)	3,308,349	
	Density of glycol (lb/gal)	8.6	
	Residence time (s)	180	
	Flash tank volume (gal)	19,235	
	Pressure factor	1	
	Module factor	2.08	
	Purchased cost of flash tank 1987 (mild steel construction)	$45,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of flash tank in 1995		**$109,366**

TABLE 4.11 (Cont.)

10.	**Solvent Circulation Pump**		
	Horsepower	2,205	
	Purchased cost of pump in 1987	0.79	
	(includes motor, coupling, base; cast iron, horizontal)	$30,000	
	Materials correction factor		
	Module factor	1.5	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of solvent pump in 1995		**$254,161**
11.	**Compressor 1 for CO_2**		
	Inlet pressure (psia)	14.70	
	Outlet pressure (psia)	50.00	
	Compressor size (hp)	539.71	
	Purchased cost of reciprocating compressor in 1987 (includes electric motor drive and gear reducer)	$160,000	
	Size factor for compressor	1	
	Materials correction factor	1	
	Module factor	2.6	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of compressor in 1995		**$486,070**
12.	**Compressor 2 for CO_2**		
	Inlet pressure (psia)	4.00	
	Outlet pressure (psia)	50.00	
	Compressor size (hp)	155.52	
	Purchased cost of reciprocating compressor in 1987 (includes electric motor drive and gear reducer)	$60,000	
	Size factor for compressor	1	
	Materials correction factor	1	
	Module factor	2.6	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of compressor in 1995		**$182,276**
13.	**Refrigeration**		
	Refrigeration (tons)	526.71	
	Purchased cost in 1987	$260,000	
	Temperature correction factor	1.25	
	Module factor	1.46	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of refrigeration in 1995		**$554,424**

TABLE 4.11 (Cont.)

14.	**CO_2 Product Gas Compressors**		
	Compressor 1 (hp)	2,582.98	
	Compressor 2 (hp)	2,582.98	
	Compressor 3 (hp)	2,582.98	
	Purchased cost of centrifugal compressor 1 in 1987	$600,000	
	Purchased cost of centrifugal compressor 2 in 1987	$600,000	
	Purchased cost of centrifugal compressor 3 in 1987 (includes electric motor drive and gear reducer)	$600,000	
	Size factor for compressor	1	
	Module factor	2.6	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of Compressor 1 in 1995		**$1,822,763**
	Installed cost of Compressor 2 in 1995		**$1,822,763**
	Installed cost of Compressor 3 in 1995		**$1,822,763**
	Total Direct Cost	**$9,532,478**	
	Total Direct Cost for Three Trains		**$28,597,433**

5 Case 2 — Gas Turbine Topping Cycle and Membrane CO_2 Recovery

5.1 Design Basis

The overall system design with membrane recovery is essentially the same as that with glycol recovery as depicted in Figure 4.1, except a membrane separation unit replaces the glycol unit. The nominal CO_2-removal efficiency of the membrane system is 90%, although the calculated design efficiency is somewhat lower, primarily because of the methane content of the synthesis gas that remains with the hydrogen-rich retentate after separation. This methane is combusted and released as CO_2 with the gas turbine exhaust. Several configurations for the membrane system were evaluated, including various series and parallel arrangements. The arrangement that most economically approaches the 90% recovery target is depicted in Figure 5.1. This system treats the sulfur-free synthesis gas flow of 11,800 pound moles per hour. The use of a recycle stream is essential to achieving the net reduction in potential CO_2 emissions of 85% that is achieved with this design. In the glycol case, the absorber design assures removal of sufficient CO_2 to compensate for combustion of the methane and still achieve 90% recovery. Membrane performance is not sufficient to compensate for this methane combustion. The gasifier and power island equipment are of the same scale and type as those used in the reference case and the glycol recovery case. Reduced gas turbine power output is expected because of changes in the fuel gas, but any associated changes in turbine design are not incorporated in this analysis. The substantial energy use for operation of compressors, fans, and pumps associated with gas cleanup is treated as a reduction in net output. In other words, the gross plant capacity is not increased to compensate for these losses. Table 5.1 is a summary of principal material flows for the base case and for this design option.

5.2 Shift Reactor

The design of the shift reactor and its integration into the system are essentially the same as those used in the glycol recovery case depicted in Section 4.2 and Figure 4.3. The key to integrating the shift reaction is to use thermal energy available from cooling the syngas to preheat the humidified fuel gas before combustion in the turbine. A slight difference in the allocation of sensible heat from initial gas cooling is evident in a comparison of Table 5.2 with Table 4.2. Specifically, less heat is allocated to the turbine fuel gas stream in the membrane case than in the glycol case, reflecting the lower temperature of the treated fuel gas after the glycol process.

5.3 Membrane Process for CO_2 Recovery

The process flows for the glycol H_2S recovery are the same as those described in Section 4.3. Refer to that discussion for process calculations for the H_2S recovery system. In this case, the H_2S-free gas is treated in the membrane system rather than by a second glycol system for CO_2 recovery. The process flows for this membrane system and associated stream

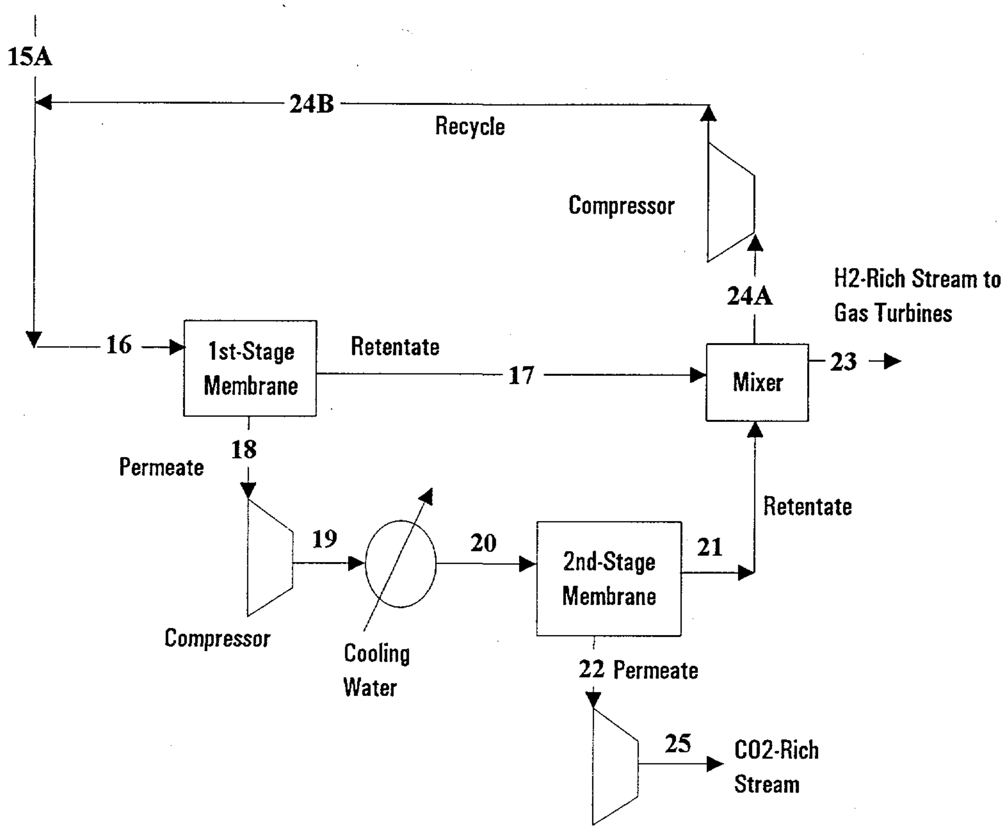

FIGURE 5.1 Flow Diagram of Membrane Process for CO_2 Recovery in Case 2

TABLE 5.1 Material Flows for Oxygen-Blown Base Case and Case 2

Material Flow (tons/d)	Base Case	Case 2
Coal (prepared)	3,845	3,845
Limestone	0	0
Air	0	0
Oxygen	2,347	2,347
Solid waste	492	492
Sulfur	78	78
CO_2 (gasifier only)	8,586	1,227
SO_2 (gasifier only)	6.92	6.92
Net power output (MW)	413.5	330

TABLE 5.2 Heat Recovery and Allocation (10^6 Btu/h) for Gas Turbine/Membrane Process in Case 2

Process	Enthalpy Change Available from Process	Allocation to Fuel Gas Preheating	Allocation for Raising Steam for Shift System	Allocation to Steam Cycle
Initial gas cooling to 460°F	513.89	327.22	123.89	62.78
Cooling after first-stage shift	168.21	0.00	168.21	0.00
Cooling after second-stage shift	673.27	171.84	215.65	285.78

calculations are summarized in Tables 5.3 and 5.4, respectively. The high level of recycle is needed to achieve the CO_2 recovery goal. The membrane technology selected for this study is the facilitated transport membrane, which incorporates an absorbent fluid layer held between two films. Such a membrane can have a high selectivity for H_2/CO_2 separation, although low permeability results in high cost. A more conventional membrane of single-layer polymeric or metallic material that is capable of effectively separating CO_2 from H_2 is not available. One scheme that has been proposed to circumvent this problem (Hendriks 1994) applies such conventional membranes directly to the synthesis gas without shift. The problem then is separation of CO from H_2.

The resulting CO-rich and H_2-rich streams are then used to fuel separate gas turbines. The exhaust from the CO turbine is a fairly pure CO_2 stream if oxygen is used as oxidant. The tradeoff is largely in the extra cost of air separation versus that of the more expensive membrane evaluated in this study.

5.4 Gas Turbine, Steam Cycle, and Plant Performance

A summary of power generation and internal power consumption when the membrane system is used for CO_2 recovery is presented in Table 5.5. The energy consumed by the CO_2-recovery system and the loss in gas turbine output, which is primarily a result of lost methane, result in an energy penalty of 20% relative to the base case generation. This result is compared in Table 5.6 with the glycol-based recovery system, which imposes an energy penalty of 9% relative to the base case.

5.5 Economics

Details of the capital investment estimates for the H_2S recovery system, the shift system, and the CO_2 recovery system are presented in Tables 5.7, 5.8, and 5.9, respectively.

TABLE 5.3 Stream Flows of Membrane Process for CO_2 Removal in Case 2

Stream Data	Stream 15A	Stream 16	Stream 17	Stream 18	Stream 19	Stream 20
Description of stream	Sulfur-free feed gas from H_2S removal section	Feed to 1st-stage membrane	Retentate from 1st-stage membrane	Permeate from 1st-stage membrane	Permeate of 1st stage after compressor	Permeate of 1st stage after heat exchange
Gases (lb·mol/h)						
CO	45.13	70.47	24.80	45.67	45.67	45.67
CO_2	4,310.02	5,007.88	598.56	4,409.32	4,409.32	4,409.32
H_2	6,822.47	17,840.18	16,725.17	1,115.01	1,115.01	1,115.01
H_2O	0.00	0.00	0.00	0.00	0.00	0.00
N_2	35.71	81.47	56.50	24.97	24.97	24.97
Ar	72.72	114.34	40.87	73.47	73.47	73.47
CH_4	463.02	941.95	542.39	399.56	399.56	399.56
NH_3	18.79	19.42	0.51	18.91	18.91	18.91
H_2S	0.63	0.65	0.02	0.64	0.64	0.64
HCl	0.41	0.43	0.01	0.41	0.41	0.41
O_2	0.00	0.00	0.00	0.00	0.00	0.00
COS	2.97	3.07	0.08	2.99	2.99	2.99
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	11,771.88	24,079.86	17,988.91	6,090.95	6,090.95	6,090.95
Liquids (lb·mol/h)	0.00	0.00	0.00	0.00	0.00	0.00
Temperature (°F)	70	86.62	86.62	86.62	538.71	212.00
Pressure (psia)	445	445.00	435.00	45.00	445.00	445.00
Enthalpy of stream (Btu/h) (reference, 32°F)	3,450,035	9,738,040	6,911,024	2,827,016	28,819,717	9,580,971

TABLE 5.3 (Cont.)

Stream Data	Stream 21	Stream 22	Stream 23	Stream 24A	Stream 24B
Description of stream	Retentate from 2nd-stage membrane	Permeate from 2nd-stage membrane	Fuel gas to gas turbines	Recycle to 1st-stage membrane	Compressed recycle to. 1st-stage membrane
Gases (lb·mol/h)					
CO	16.07	29.60	15.53	25.34	25.34
CO_2	527.01	3,882.31	427.72	697.85	697.85
H_2	1,045.32	69.69	6,752.79	11,017.70	11,017.70
H_2O	0.00	0.00	0.00	0.00	0.00
N_2	17.32	7.65	28.05	45.77	45.77
Ar	26.26	47.21	25.51	41.62	41.62
CH_4	230.07	169.49	293.54	478.93	478.93
NH_3	0.50	18.41	0.38	0.63	0.63
H_2S	0.02	0.62	0.01	0.02	0.02
HCl	0.01	0.40	0.01	0.01	0.01
O_2	0.00	0.00	0.00	0.00	0.00
COS	0.08	2.91	0.06	0.10	0.10
SO_2	0.00	0.00	0.00	0.00	0.00
Total gas flow	1,862.67	4,228.28	7,543.60	12,307.98	12,307.98
Liquids (lb·mol/h)	0.00	0.00	0.00	0.00	0.00
Temperature (°F)	212.00	212.00	99.64	99.64	103.98
Pressure (psia)	435.00	45.00	435.00	435.00	445.00
Enthalpy of stream (Btu/h) (reference, 32°F)	2,617,896	6,963,075	3,620,990	5,907,930	6,288,005

TABLE 5.4 Descriptions of Streams of Membrane Process for CO_2 Removal in Case 2

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 15A: Sulfur-free gas from H_2S section		
Temperature (°F)	70	The synthesis gas is cleaned in two stages. First sulfur compounds are removed. Then they are fed to the membrane system for CO_2 recovery.
Pressure (psia)	445	
Flow rate (lb·mol/h)	11,771.88	
CO_2 (mole fraction)	0.3661	
H_2S (mole fraction)	0.0001	
Stream 16: Feed gas to 1st-stage membrane system		
Temperature (°F)	86.62	The sulfur-free gas is mixed with the recycle from the 2nd-stage retentate and fed to the 1st-stage membranes.
Pressure (psia)	445	
Flow rate (lb·mol/h)	24,079.86	
CO_2 (mole fraction)	0.2080	
H_2S (mole fraction)	0.0000	
Stream 17: Retentate from 1st-stage membrane system		
Temperature (°F)	86.62	The composition of this stream depends on the permeability and selectivity of the membranes. The membrane system is a facilitated membrane that has a higher selectivity and permeability for CO_2 than H_2.
Pressure (psia)	435	
Flow rate (lb·mol/h)	17,988.91	
CO_2 (mole fraction)	0.0333	
H_2S (mole fraction)	0.0000	
Stream 18: Permeate from 1st-stage membrane system		
Temperature (°F)	86.62	The composition of this stream is calculated by mass balance around the membrane.
Pressure (psia)	45	
Flow rate (lb·mol/h)	6,090.95	
CO_2 (mole fraction)	0.7329	
H_2S (mole fraction)	0.0001	
Stream 19: Gases from compressor		
Temperature (°F)	538	The permeate from 1st-stage membrane systems is at a pressure of 45 psia. These gases are again compressed to a pressure of 445 psia for the 2nd-stage membrane system.
Pressure (psia)	445	
Flow rate (lb·mol/h)	6,090.95	
CO_2 (mole fraction)	0.7329	
H_2S (mole fraction)	0.0001	

TABLE 5.4 (Cont.)

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 20: Gases from heat exchanger		
Temperature (°F)	212	The temperature of the gases rises because of the compression. Therefore, this stream is cooled to a temperature of 212°F, suitable for the membrane system.
Pressure (psia)	445	
Flow rate (lb·mol/h)	6,090.95	
CO_2 (mole fraction)	0.7329	
H_2S (mole fraction)	0.0001	
Stream 21: Retentate of 2nd-stage membrane system		
Temperature (°F)	212	The composition of this stream is calculated on the basis of the selectivity and permeability of gases, as is done for stream 17.
Pressure (psia)	435	
Flow rate (lb·mol/h)	1,862.67	
CO_2 (mole fraction)	0.2829	
H_2S (mole fraction)	0.0000	
Stream 22: Permeate of 2nd-stage membrane system		
Temperature (°F)	212	The composition of this stream is calculated on the basis of the mass balance around the membrane. This is the rich-CO_2 stream for disposal.
Pressure (psia)	45	
Flow rate (lb·mol/h)	4,228.28	
CO_2 (mole fraction)	0.9182	
H_2S (mole fraction)	0.0001	
Stream 23: Fuel gas to gas turbines		
Temperature (°F)	99.64	H_2-rich retentate from the 1st stage (stream 17) and that from the 2nd stage (stream 21) are mixed, and part of mixture is taken as fuel gas for gas turbines.
Pressure (psia)	435	
Flow rate (lb·mol/h)	7,543.60	
CO_2 (mole fraction)	0.0567	
H_2S (mole fraction)	0.0000	
Stream 24A: Recycle to 1st-stage membrane system		
Temperature (°F)	99.64	Part of the retentate from stream 17 and part from stream 21 are recycled back to the 1st-stage membrane systems to increase the CO_2-removal efficiency.
Pressure (psia)	435	
Flow rate (lb·mol/h)	12,307.98	
CO_2 (mole fraction)	0.0567	
H_2S (mole fraction)	0.0000	
Stream 24B: Recycle to 1st-stage membrane after compression		
Temperature (°F)	103.98	The recycle from the retentate is at a pressure of 435 psia and is compressed to the inlet pressure of the 1st membrane.
Pressure (psia)	445	
Flow rate (lb·mol/h)	12,307.98	
CO_2 (mole fraction)	0.0567	
H_2S (mole fraction)	0.0000	

TABLE 5.5 Turbine Output, Plant Power Use, and Net Power Output for Base Case and Case 2 Gas Turbine/Membrane Process

Power Variable	Power (MW) Base Case	Membrane Case
Power output		
Gas turbine	298.8	262.8
Steam turbine	159.4	154.8
Internal power consumption		
CO_2 recovery		
CO_2 compression	0.0	-20.0
Solvent circulation	0.0	-0.9
Solvent refrigeration	0.0	-3.0
Others	0.0	-19.0
Gasification system	-44.7	-44.7
Net power output	413.5	330.0
Energy penalty	0.0	83.5

TABLE 5.6 Overall Power Recovery and Production for Three Gas Turbine Cases

Power Variable	Power (MW) Base Case	Glycol Case 1	Membrane Case 2
Power output			
Gas turbine	298.8	284.8	262.8
Steam turbine	159.4	161.6	154.8
Internal power consumption			
CO_2 recovery	0.0	-24.2	-42.9
Gasification system	-44.7	-44.7	-44.7
Net power output	413.5	377.5	330.0
Energy penalty	0.0	36.0	83.5

TABLE 5.7 Sizing and Cost Estimation for Major Equipment Used for H_2S Removal in Glycol Process in Case 2

1.	**Heat Exchanger before the Absorption Column**		
	Q = Load (Btu/h)	3,630,585	
	Tha = Inlet temperature of hot fluid (°F)	100	
	Thb = Outlet temperature of hot fluid (°F)	63	
	Pressure of hot gases (psia)	451	
	Tca = Inlet temperature of cold fluid (°F)	30	
	Tcb = Outlet temperature of cold fluid (°F)	70	
	Delta T1	30	
	Delta T2	33	
	Log mean temperature difference (°F)	31	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	5	
	Heat transfer area (ft^2)	23,070	
	Operating pressure (psia)	451	
	Pressure factor	1.175	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$185,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$812,765**
2.	**H_2S Absorption Column**		
	Diameter of tower (ft)	8	
	HETP (ft)	3	
	No. of theoretical stages	12	
	Absorber tower height (ft) (4 ft for inlet, outlet and gas, and liquid distributors)	40	
	Volume of packing (ft^3)	1,810	
	Pressure factor	2.6	
	Cost per foot of column height per foot (mild steel construction)	$1,000	
	Materials correction factor	1	
	Module factor	4.16	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of absorber in 1995		**$505,513**
	Cost of packing per cubic foot (2-in. pall rings-metal)	$63.5	
	Total cost of packing		**$114,953**

TABLE 5.7 (Cont.)

3.	**Power Recovery Turbine 1**		
	Turbine size (hp)	173	
	Purchased cost in 1987	$120,000	
	Module factor	1	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of solvent pump in 1995		**$175,266**
4.	**Slump Tank**		
	Glycol solvent flow rate (lb/h)	613,374	
	Density of glycol solvent (lb/gal)	8.6	
	Residence time (s)	180	
	Slump tank volume (gal)	3,566	
	Pressure factor	1	
	Materials correction factor	1	
	Module factor	2.08	
	Purchased cost of slump tank in 1987 (mild steel construction)	$13,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of slump tank in 1995		**$31,595**
5.	**Power Recovery Turbine 2**		
	Turbine size (hp)	43	
	Purchased cost in 1987	$65,000	
	Module factor	1	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of solvent pump in 1995		**$75,948**
6.	**Solvent Circulation Pump**		
	Horsepower	403	
	Purchased cost of pump in 1987 (includes motor, coupling, base; cast iron, horizontal)	$30,000	
	Materials correction factor	1	
	Module factor	1.5	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of solvent pump in 1995		**$52,580**

TABLE 5.7 (Cont.)

7.	**Lean-Rich Solvent Heat Exchanger**		
	Q = Load (Btu/h)	47,524,550	
	Tha = Inlet temperature of hot fluid (°F)	215.21	
	Thb = Outlet temperature of hot fluid (°F)	67	
	Pressure of hot gases (psia)	450	
	Tca = Inlet temperature of cold fluid (°F)	42.10	
	Tcb = Outlet temperature of cold fluid (°F)	190.00	
	Delta T1	25.2077	
	Delta T2	25	
	Log mean temperature difference (°F)	25	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	150	
	Heat transfer area (ft^2)	12,697	
	Operating pressure (psia)	50	
	Pressure factor	1	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$120,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$448,680**
8.	**Stripping Column**		
	Diameter of tower (ft)	10	
	HETP (ft)	3	
	No. of theoretical stages	12	
	Absorber tower height (4 ft for inlet, outlet and gas, and liquid distributors)	40	
	Volume of packing (ft^3)	2,829	
	Pressure factor	1	
	Materials correction factor (stainless steel 304)	1.7	
	Cost per ft of column height (mild steel construction)	$1,200	
	Module factor	4.16	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of absorber in 1995		**$396,633**
	Cost of packing per cubic foot (2-in. pall rings-SS)	$63.5	
	Materials correction factor	1	
	Total cost of packing		**$179,614**

TABLE 5.7 (Cont.)

9.	**Overhead Condenser**		
	Q = Load (Btu/h)	20,323,399	
	Tha = Inlet temperature of hot fluid (°F)	212.00	
	Thb = Outlet temperature of hot fluid (°F)	100	
	Pressure of hot gases (psia)	14.7	
	Tca = Inlet temperature of cold fluid (°F)	70.00	
	Tcb = Outlet temperature of cold fluid (°F)	180.00	
	Delta T1	32	
	Delta T2	30	
	Log mean temperature difference (°F)	31	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	40	
	Heat transfer area (ft^2)	16,396	
	Operating Pressure (psia)	14.7	
	Pressure factor	1	
	Materials correction factor 1 for SS	2.7	
	Materials correction factor 2 for SS	0.07	
	Materials correction factor	3.48	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating Head)	$160,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$2,079,839**
10.	**Phase Separator**		
	Flow rate (lb/h)	22,291	
	Density of fluid (lb/gal)	0.04	
	Residence time (s)	120	
	Phase separator volume (gal)	18,576	
	Pressure factor	1	
	Materials correction factor (stainless steel)	1.8	
	Module factor	2.08	
	Purchased cost of phase separator in 1987 (mild steel construction)	$44,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of phase separator in 1995		**$192,484**

TABLE 5.7 (Cont.)

11.	**Solvent Refrigeration**		
	Refrigeration (tons)	981.96	
	Purchased cost in 1987	$400,000	
	Temperature correction factor	1.25	
	Module factor	1.46	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of refrigeration in 1995		**$852,959**
	Total Direct Cost		**$5,918,829**
	Total Direct Cost for Three Trains		**$17,756,488**

TABLE 5.8 Sizing and Cost Estimation for Major Equipment Used for Shift System in Case 2

1.	**First-Stage Shift Reactor**		
	Catalyst volume (ft^3)	665	
	Reactor volume (ft^3) (1.2 times the catalyst volume)	798	
	Reactor volume (gal)	5,969	
	Pressure factor	2.8	
	Module factor	3.05	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Purchased cost of reactor in 1987	$7,000	
	Installed cost of reactor in 1995		**$69,849**
2.	**Second-Stage Shift Reactor**		
	Catalyst volume (ft^3)	285	
	Reactor volume (ft^3) (1.2 times the catalyst volume)	342	
	Reactor volume (gal)	2,558	
	Pressure factor	2.8	
	Module factor	3.05	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Purchased cost of reactor in 1987	$5,000	
	Installed cost of reactor in 1995		**$49,849**
3.	**Cost of Shift Catalyst**		
	Volume of catalyst in first stage (ft^3)	999	
	Volume of catalyst in second stage (ft^3)	339	
	Cost of high-temperature catalyst per cubic foot	$50	
	Cost of low-temperature catalyst per cubic foot	$250	
	Total cost of catalyst		**$134,647**

TABLE 5.8 (Cont.)

4.	**Heat Exchanger between First- and Second-Shift Stages**		
	Q = Load (Btu/h)	56,070,250	
	Tha = Inlet temperature of hot fluid (°F)	684	
	Thb = Outlet temperature of hot fluid (°F)	457	
	Pressure of hot gases (psia)	451	
	Tca = Inlet temperature of cold fluid (°F)	400	
	Tcb = Outlet temperature of cold fluid (°F)	457	
	Delta T1	226	
	Delta T2	57	
	Log mean temperature difference (°F)	123	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	40	
	Heat transfer area (ft^2)	11,383	
	Operating pressure (psia)	451	
	Pressure factor	1.175	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$120,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$527,199**
5.	**Heat Exchanger after Second-Stage Shift for Raising Steam**		
	Q = Load (Btu/h)	71,881,771	
	Tha = Inlet temperature of hot fluid (°F)	457	
	Thb = Outlet temperature of hot fluid (°F)	457	
	Pressure of hot gases (psia)	451	
	Tca = Inlet temperature of cold fluid (°F)	100	
	Tcb = Outlet temperature of cold fluid (°F)	400	
	Delta T1	57	
	Delta T2	357	
	Log mean temperature difference (°F)	164	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	40	
	Heat transfer area (ft^2)	10,955	
	Operating pressure (psia)	451	
	Pressure factor	1.175	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$118,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$518,412**

TABLE 5.8 (Cont.)

6.	**Heat Exchanger after Second-Stage for Heating Fuel Gas**		
	Q = Load (Btu/h)	57,280,972	
	Tha = Inlet temperature of hot fluid (°F)	457	
	Thb = Outlet temperature of hot fluid (°F)	457	
	Pressure of hot gases (psia)	451	
	Tca = Inlet temperature of cold fluid (°F)	100	
	Tcb = Outlet temperature of cold fluid (°F)	400	
	Delta T1	57	
	Delta T2	356	
	Log mean temperature difference (°F)	164	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	5	
	Heat transfer area (ft^2)	70,015	
	Operating pressure (psia)	451	
	Pressure factor	1.175	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$400,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$1,757,330**
7.	**Heat Exchanger for Heating Clean Fuel Gas with Raw Gases from Gasifier**		
	Q = Load (Btu/h)	327,214,827	
	Tha = Inlet temperature of hot fluid (°F)	1,750	
	Thb = Outlet temperature of hot fluid (°F)	977	
	Pressure of hot gases (psia)	465	
	Tca = Inlet temperature of cold fluid (°F)	400	
	Tcb = Outlet temperature of cold fluid (°F)	1,100	
	Delta T1	650	
	Delta T2	577	
	Log mean temperature difference (°F)	613	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	5	
	Heat transfer area (ft^2)	106,782	
	Operating pressure (psia)	465	
	Pressure factor	1.175	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$500,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$2,196,663**

TABLE 5.8 (Cont.)

8.	**Heat Exchanger for Cooling Shifted Synthesis Gas with Feedwater**		
	Q = Load (Btu/h)	95,260,677	
	Tha = Inlet temperature of hot fluid (°F)	456	
	Thb = Outlet temperature of hot fluid (°F)	100	
	Pressure of hot gases (psia)	457	
	Tca = Inlet temperature of cold fluid (°F)	70	
	Tcb = Outlet temperature of cold fluid (°F)	400	
	Delta T1	56	
	Delta T2	30	
	Log mean temperature difference (°F)	42	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	50	
	Heat transfer area (ft^2)	45,899	
	Operating pressure (psia)	457	
	Pressure factor	1.175	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$320,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$1,405,864**
	Total Direct Cost		**$6,659,856**
	Total Direct Cost for Three Trains		**$19,979,567**

TABLE 5.9 Sizing and Cost Estimation for Major Equipment Used for CO_2 Removal in Membrane Process in Case 2

1.	**First-Stage Membranes**		
	Membrane area (ft^2)	1,639,589	
	Unit cost of membrane	$13.00	
	Total cost		**$21,314,656**
2.	**Second-Stage Membranes**		
	Membrane area (ft^2)	414,731	
	Unit cost of membrane	$13.00	
	Total cost		**$5,391,500**
3.	**Compressor between First and Second Stages**		
	Inlet pressure (psia)	45.00	
	Outlet pressure (psia)	445.00	
	Compressor size (hp)	10,208	
	Purchased cost of reciprocating compressor in 1987 (includes electric motor drive and gear reducer)	$1,600,000	
	Size factor for compressor	1	
	Materials correction factor	1	
	Module factor	2.6	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of compressor in 1995		**$4,860,700**
4.	**Recycle Compressor**		
	Inlet pressure (psia)	435.00	
	Outlet pressure (psia)	445.00	
	Compressor size (hp)	149	
	Purchased cost of reciprocating compressor in 1987 (includes electric motor drive and gear reducer)	$60,000	
	Size factor for compressor	1	
	Materials correction factor	1	
	Module factor	2.6	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of compressor in 1995		**$182,276**

TABLE 5.9 (Cont.)

5.	**Heat Exchanger After Compressor**		
	Q = Load (Btu/h)	19,238,746	
	Tha = Inlet temperature of hot fluid (°F)	538.71	
	Thb = Outlet temperature of hot fluid (°F)	212	
	Pressure of hot gases (psia)	450	
	Tca = Inlet temperature of cold fluid (°F)	70.00	
	Tcb = Outlet temperature of cold fluid (°F)	150.00	
	Delta T1	388.71	
	Delta T2	142	
	Log mean temperature difference (°F)	245	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	40	
	Heat transfer area (ft^2)	1,963	
	Operating pressure (psia)	445	
	Pressure factor	1.08	
	Materials correction factor	1	
	Module factor	3.2	
	(includes all of the supporting equipment and connections and installation)		
	Purchased cost of heat exchanger in 1987	$36,000	
	(mild steel construction; shell and tube floating head)		
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$145,372**
6.	**CO_2 Product Gas Compressors**		
	Compressor 1 (hp)	2,583	
	Compressor 2 (hp)	2,583	
	Compressor 3 (hp)	2,583	
	Purchased cost of centrifugal compressor 1 in 1987	$540,000	
	Purchased cost of centrifugal compressor 2 in 1987	$540,000	
	Purchased cost of centrifugal compressor 3 in 1987	$540,000	
	(includes electric motor drive and gear reducer)		
	Size factor for compressor	1	
	Module factor	2.6	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of Compressor 1 in 1995		**$1,640,486**
	Installed cost of Compressor 2 in 1995		**$1,640,486**
	Installed cost of Compressor 3 in 1995		**$1,640,486**
	Total Direct Cost		**$36,815,962**
	Total Direct Cost for Three Trains		**$110,447,887**

6 Case 3 — Fuel Cell Topping Cycle and Glycol CO_2 Recovery

Because fuel cells require a hydrogen-rich fuel stream, the fuel cell system employs a reformer to convert hydrocarbon fuels to hydrogen-rich fuels. For medium-Btu coal gas, a shift reaction is required to create a hydrogen-rich fuel. Because of the high operating temperature of the molten carbonate fuel cell, a reforming or a shift reaction can take place within the cell, eliminating the need for separate reactors for these processes. The associated economies recommend a fuel cell as the topping cycle for IGCC with CO_2 recovery. Material and energy balances have been developed in this section for the application of an internal reforming molten carbonate fuel cell as the topping cycle for an IGCC plant. The CO_2 from the fuel cell exhaust is recovered in a glycol process. This situation is quite different from use of a gas turbine topping cycle, in which CO_2 recovery must precede use of the fuel in the turbine to avoid dilution with air, which would increase the cost of CO_2 recovery.

6.1 Design Basis

Figure 6.1 provides an overview of the of the IGCC system, including the gasifier, gas treatment, the fuel cell, and the steam cycle. The overall design of the fuel cell is determined by the gasifier capacity and synthesis gas composition. These are assumed to be the same as in the base case, which has no CO_2 recovery. The fuel cell has very low tolerance for contaminants, including particulates and sulfur compounds. To achieve the required level of H_2S removal, a chilled methanol system has been employed rather than the glycol system used in the gas turbine cases. The chilled methanol system is designed to reduce the sulfur species (H_2S and COS) concentration to less than 1 part per million volume (ppmv). The reactions in the fuel cell anode shift the synthesis gas to a hydrogen-rich gas with a high concentration of CO_2 and reduce the resultant hydrogen with carbonate ion. Oxidation of the carbonate at the anode releases CO_2 and two moles of electrons per mole of H_2 converted. The CO_2-rich anode exhaust is treated in a glycol recovery system to separate most of the CO_2. Thermal energy released by cooling this anode exhaust provides heat for the steam bottoming cycle. An expansion turbine is used on the cathode exhaust to extract energy.

Table 6.1 is a summary of principal material flows for the base case and for this design option. The CO_2 reduction accomplished at the power plant is 89% and is accompanied by a 25% reduction in net electrical output. A full accounting of the net CO_2 reduction would include CO_2 released in the generation of replacement power, mining, coal and reagent preparation, and materials transport.

6.2 Chilled Methanol Process for H_2S Recovery

Because of the extremely low tolerance of the fuel cell for H_2S, a chilled methanol process has been employed rather than the more economical glycol process preferred for the base

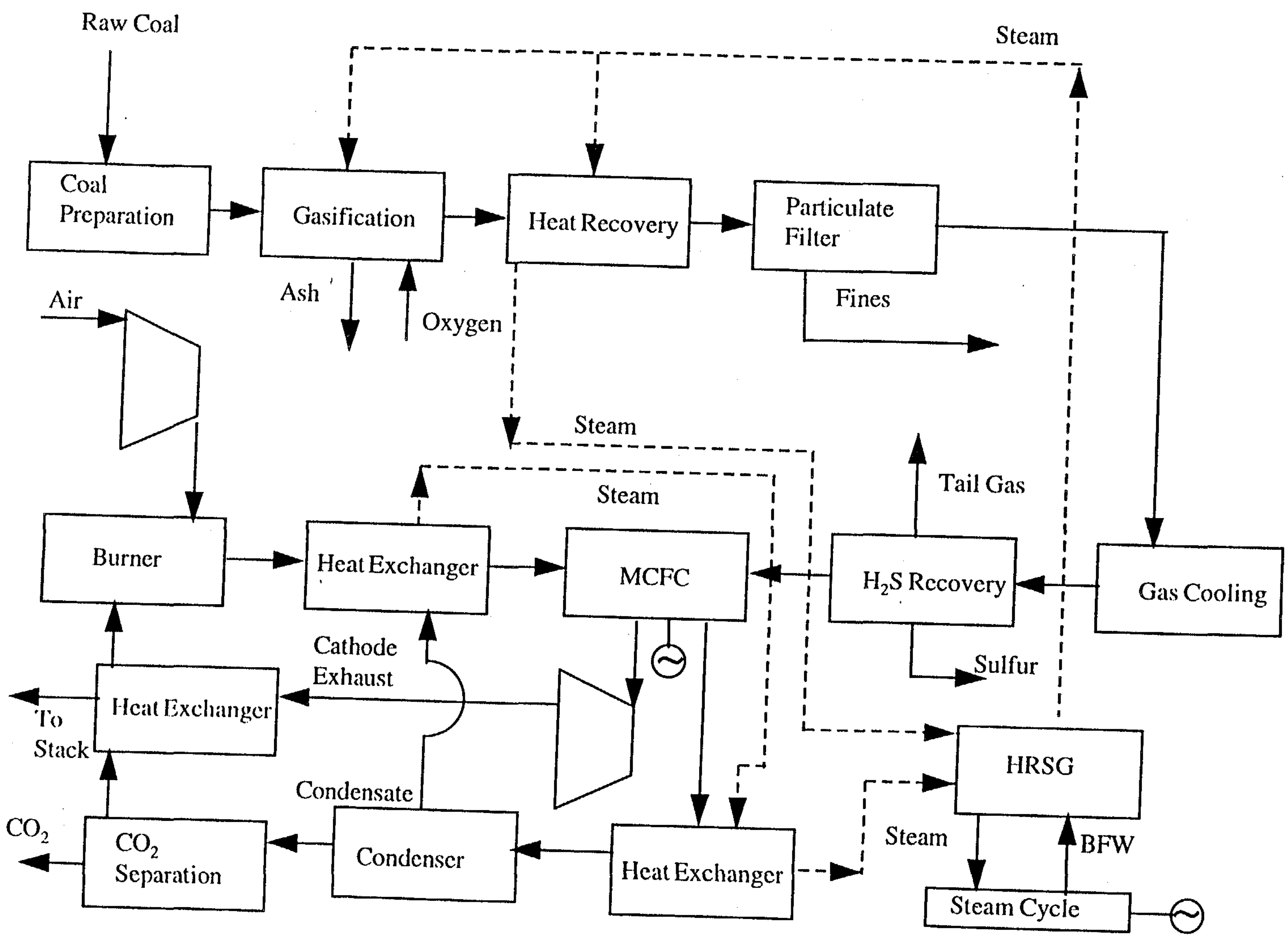

FIGURE 6.1 Block Diagram of the IGCC System with CO_2 Recovery Used in Cases 3 and 4

TABLE 6.1 Material Flows for Oxygen-Blown Base Case and Case 3

Material Flow (tons/d)	Base Case	Case 3
Coal (prepared)	3,845	3,845
Oxygen	2,347	2,347
Solid waste	492	492
Sulfur	78	78
CO_2 (power plant only)	9210	993
SO_2 (power plant only)	1.08	6.92
Net power output (MW)	458.4	340.11

case and gas turbine options. The chilled methanol process is depicted in Figure 6.2. The feed gas is cooled by heat exchange with the cleaned fuel gas. Because it is cooled to well below the point at which water would condense and freeze, methanol is added to the feed gas to act as an antifreeze. Condensate is removed in a phase separater and sent to a distillation unit to recover the methanol. The rich methanol from the absorber is flashed in three stages to release the H_2S and is finally stripped with steam heating. The lean methanol from the stripper is cooled by heat exchange with the methanol feed to the stripper and by refrigeration prior to reinjection into the absorber tower. Table 6.2 provides the details of stream composition, flows, and conditions for the H_2S recovery system. Comparing the feed stream, 1A, with the product stream, 2B, the reduction in H_2S in lb·mol/h is 99.99% and the H_2S content of the fuel gas is about 0.7 ppmv. A description of the streams and assumptions used in the stream calculations is provided in Table 6.3.

6.3 Molten Carbonate Fuel Cell System

Figure 6.3 shows the molten carbonate fuel cell in the context of supporting systems. The sulfur-free gas from the methanol system is brought to fuel cell operating pressure in a power recovery turbine. The gas is then heated by steam injection and fed to the fuel cell, where the shift reaction converts CO to CO_2 and reforming converts CH_4 to H_2 and CO_2. The anode exhaust is rich in CO_2. The sensible heat of this stream is used to raise steam for the steam cycle. After further cooling by heat exchange with steam cycle condensate, the anode exhaust is sent to CO_2 recovery following water removal in a condenser. The CO_2-lean gas has residual CO and H_2, which is burned in air before this stream is used as the cathode feed. The cathode exhaust is sent through a power recovery turbine and heat exchangers before being exhausted as stack gas. The line list corresponding to Figure 6.3 is provided in Table 6.4. A description of the streams and key assumptions are provided in Table 6.5.

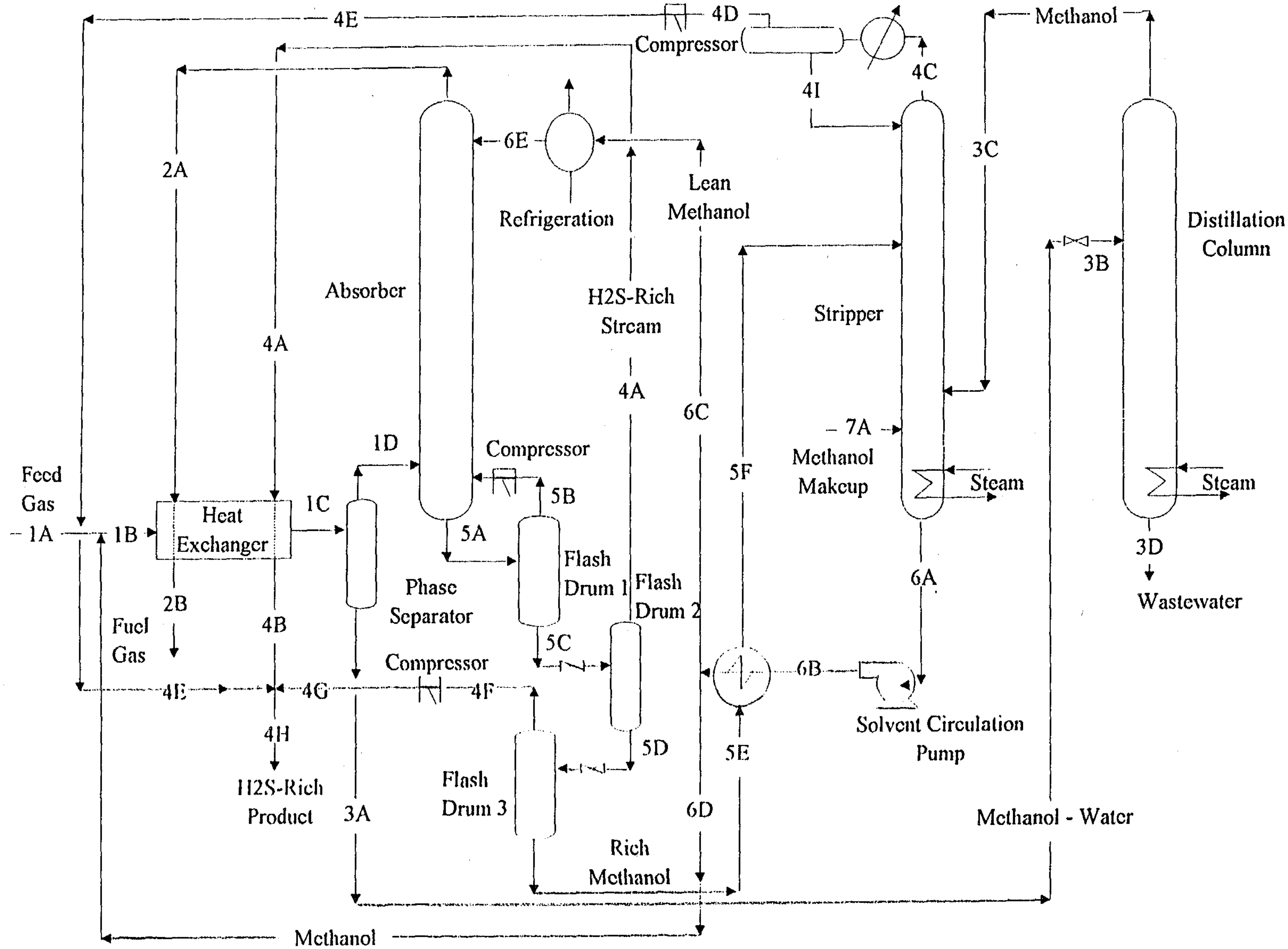

FIGURE 6.2 Flow Diagram of Chilled Methanol Process for H_2S Recovery in Case 3

TABLE 6.2 Stream Flows of Chilled Methanol Process for H_2S Removal in Case 3

Stream Data	Stream 1A	Stream 1B	Stream 1C	Stream 1D	Stream 2A	Stream 2B
Description of stream	Feed gas from KRW gasifier	Gases to heat exchanger	Gases from heat exchanger	Gases from phase separator	Sulfur-free gas from absorber	Sulfur-free gas to fuel cell
Gases (lb·mol/h)						
CO	4,559.29	4,559.29	4,559.29	4,559.29	4,530.65	4,530.65
CO_2	389.37	389.37	389.37	389.37	267.08	267.08
H_2	2,315.44	2,315.44	2,315.44	2,315.44	2,311.51	2,311.51
H_2O	19.41	19.41	19.41	0.00	0.00	0.00
N_2	36.44	36.44	36.44	36.44	36.44	36.44
Ar	72.73	72.73	72.73	72.73	72.73	72.73
CH_4	487.31	487.31	487.31	487.31	480.56	480.56
NH_3	0.00	0.00	0.00	0.00	0.01	0.01
H_2S	59.01	59.01	59.01	59.01	5.90E-03	5.90E-03
HCN	0.00	0.00	0.00	0.00	7.42E-04	7.42E-04
O_2	0.00	0.00	0.00	0.00	0.00	0.00
COS	7.42	7.42	7.42	7.42	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	7,946.42	7,946.42	7,946.42	7,927.01	7,698.98	7,698.98
Liquids (lb·mol/h) Methanol	0.00	15.07	15.07	0.00	0.03	0.03
Temperature (°F)	105.00	104.55	-34.18	-34.18	-70.00	80.00
Pressure (psia)	456.00	456.00	456.00	456.00	450.00	450.00
Enthalpy of stream (Btu/h) (reference, 32°F)	4,444,715	4,433,245	-3,707,336	-3,666,146	-5,440,448	2,596,543

TABLE 6.2 (Cont.)

Stream Data	Stream 3A	Stream 3B	Stream 3C	Stream 3D	Stream 4A	Stream 4B
Description of stream	Bottoms from phase separator	Feed to distillation column	Overhead from distillation column	Wastewater from distillation column	H_2S-rich gas from flash drum 2	H_2S-rich gas from heat exchanger
Gases (lb·mol/h)						
CO	0.00	0.00	0.00	0.00	21.48	21.48
CO_2	0.00	0.00	0.00	0.00	61.14	61.14
H_2	0.00	0.00	0.00	0.00	3.14	3.14
H_2O	19.41	19.41	0.79	18.62	0.00	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	4.05	4.05
NH_3	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	0.00	0.00	0.00	0.00	17.83	17.83
HCN	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	2.35	2.35
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	19.41	19.41	0.79	18.62	109.99	109.99
Liquids (lb·mol/h) Methanol	15.07	15.07	15.07	0.00	0.00	0.00
Temperature (°F)	-34.18	-34.18	150.00	280.00	-29.88	84.50
Pressure (psia)	456.00	50.00	50.00	50.00	150.00	150.00
Enthalpy of stream (Btu/h) (reference, 32°F)	-41,190	-41,190	270,074	83,134	-55,403	48,168

TABLE 6.2 (Cont.)

Stream Data	Stream 4C	Stream 4D	Stream 4E	Stream 4F	Stream 4G	Stream 4H
Description of stream	Overhead from stripper	H_2S-rich gas from phase separator	H_2S-rich gas after compressor	H_2S-rich gas from flash drum 3	H_2S-rich gas after compressor	H_2S-rich product
Gases (lb·mol/h)						
CO	1.79	1.79	1.79	5.37	5.37	28.64
CO_2	30.57	30.57	30.57	30.57	30.57	122.29
H_2	0.16	0.16	0.16	0.63	0.63	3.93
H_2O	0.79	0.79	0.79	0.00	0.00	0.79
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	1.35	1.35	1.35	1.35	1.35	6.75
NH_3	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	28.70	28.70	28.70	12.48	12.48	59.01
HCN	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	3.42	3.42	3.42	1.64	1.64	7.42
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	66.78	66.78	66.78	52.04	52.04	228.83
Liquids (lb·mol/h) Methanol	224.97	21.45	21.45	0.00	0.00	21.45
Temperature (°F)	135.00	100.00	619.94	-33.64	286.22	318.59
Pressure (psia)	14.70	14.70	150.00	20.00	150.00	95.00
Enthalpy of stream (Btu/h) (reference, 32°F)	3,874,837	410,887	803,850	-28,354	117,525	969,544

TABLE 6.2 (Cont.)

Stream Data	Stream 4I	Stream 5A	Stream 5B	Stream 5C	Stream 5D	Stream 5E
Description of stream	Methanol reflux to stripper	Rich methanol from absorber	Gases from flash drum 1	Feed to flash drum 2	Feed to flash drum 3	Rich methanol from flash drum 3
Gases (lb·mol/h)						
CO	0.00	286.38	257.74	28.64	7.16	1.79
CO_2	0.00	152.86	30.57	122.29	61.14	30.57
H_2	0.00	39.27	35.34	3.93	0.79	0.16
H_2O	0.00	0.00	0.00	0.00	0.00	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	67.52	60.77	6.75	2.70	1.35
NH_3	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	0.00	66.02	6.60	59.42	41.60	29.12
HCN	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	8.70	0.87	7.83	5.48	3.84
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	0.00	620.75	391.89	228.86	118.87	66.83
Liquids (lb·mol/h) Methanol	203.53	4,114.03	0.00	4,114.03	4,114.03	4,114.03
Temperature (°F)	100.00	-23.04	-23.04	-23.04	-29.88	-33.64
Pressure (psia)	95.00	450.00	300.00	300.00	150.00	20.00
Enthalpy of stream (Btu/h) (reference, 32°F)	250,640	-4,358,335	-153,557	-4,204,777	-4,671,853	-4,927,451

TABLE 6.2 (Cont.)

Stream Data	Stream 5F	Stream 6A	Stream 6B	Stream 6C	Stream 6D	Stream 6E	Stream 7A
Description of stream	Rich methanol to stripper	Lean methanol to circulation pump	Lean methanol from solvent circulation pump	Lean methanol to refrigeration	Methanol for feed gas injection	Lean methanol to absorber	Methanol makeup to stripper
Gases (lb·mol/h)							
CO	1.79	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	30.57	0.00	0.00	0.00	0.00	0.00	0.00
H_2	0.16	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N_2	0.00	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	0.00
CH_4	1.35	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	0.00
H_2S	29.12	0.41	0.41	0.41	0.00	0.41	0.00
HCN	0.00	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	0.00
COS	3.84	0.41	0.41	0.41	0.00	0.41	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	66.83	0.82	0.82	0.82	0.00	0.82	0.00
Liquids (lb·mol/h) Methanol	4,114.03	4,129.14	4,129.14	4,114.06	15.07	4,114.06	21.48
Temperature (°F)	128.66	149.00	152.90	-10.00	-10.00	-70.00	70.00
Pressure (psia)	20.00	14.70	456.00	456.00	456.00	456.00	14.70
Enthalpy of stream (Btu/h) (reference, 32°F)	7,257,493	8,749,976	9,043,938	-3,129,543	-11,463	-7,600,310	14,782

TABLE 6.3 Descriptions of Streams of Chilled Methanol Process for H_2S Removal in Case 3

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 1A: Gas feed from KRW process		
Temperature (°F)	105	This stream is coming from KRW process.
Pressure (psia)	456	This stream will be cooled against cold fuel gas
Flow rate (lb·mol/h)	7,946	from absorber and cold H_2S-rich gas from flash
H_2S (mol fraction)	0.0074	drum 2.
Stream 2A: Fuel gas from top of absorber		
Temperature (°F)	-70	Chilled methanol enters the top of the column at a
Pressure (psia)	450	temperature of -70°F. Gases leaving the column
Flow rate (lb·mol/h)	7,699	are in equilibrium with methanol; hence, they
H_2S (ppm)	0.767	are at a temperature of -70°F. Gas composition corresponds to 99.99% removal of H_2S .
Stream 3A: Methanol-water mixture from phase separator		
Temperature (°F)	-34.18	Methanol is added to feed gas prior to absorption
Pressure (psia)	456	column to prevent icing of water in feed gas.
Flow rate (lb·mol/h)	34.49	Condensed water and methanol are separated
H_2S (mol fraction)	0	from gas in phase separator.
Stream 3B: Methanol-water mixture to distillation column		
Temperature (°F)	-34.18	Methanol is separated from the methanol-water
Pressure (psia)	50	mixture in distillation column.
Flow rate (lb·mol/h)	34.49	
H_2S (mol fraction)	0	
Stream 3C: Methanol from distillation column to stripper		
Temperature (°F)	150	Methanol from distillation column is sent to
Pressure (psia)	50	stripper.
Flow rate (lb·mol/h)	15.86	
H_2S (mol fraction)	0	
Stream 3D: Wastewater from distillation column		
Temperature (°F)	280	Water from distillation column is removed from
Pressure (psia)	50	bottom of column for disposal.
Flow rate (lb·mol/h)	18.62	
H_2S (mol fraction)	0	
Stream 4A: H_2S-rich gas from flash drum 2		
Temperature (°F)	-29.88	Rich methanol from flash drum 1 is flashed to
Pressure (psia)	150	pressure of 150 psia to desorb major portion
Flow rate (lb·mol/h)	109.99	of H_2S from solvent.
H_2S (mol fraction)	0.1621	

TABLE 6.3 (Cont.)

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 4C: H_2S-rich gas from stripper		
Temperature (°F)	135	The final removal of H_2S is achieved in stripper by heat. Because of low vapor pressure of methanol, substantial amounts of methanol will be vaporized along with value of H_2S.
Pressure (psia)	14.7	
Flow rate (lb·mol/h)	291.76	
H_2S (mol fraction)	0.0984	
Stream 4D: H_2S-rich gas from phase separator		
Temperature (°F)	100	Methanol is condensed from H_2S -methanol mixture, and H_2S is separated in phase separator.
Pressure (psia)	14.7	
Flow rate (lb·mol/h)	88.24	
H_2S (mol fraction)	0.3253	
Stream 4F: H_2S-rich gas from flash drum 3		
Temperature (°F)	-33.64	Rich methanol solution from flash drum 2 is further flashed to pressure of 20 psia in flash drum 3 to desorb H_2S from solvent.
Pressure (psia)	20	
Flow rate (lb·mol/h)	52.04	
H_2S (mol fraction)	0.2398	
Stream 4H: Final H_2S-rich product		
Temperature (°F)	318.59	The H_2S-rich streams from stripper and flash drum 3 are compressed to pressure of 95 psia and then combined with H_2S-rich stream from flash drum 2. This stream is further processed in a Claus plant for sulfur recovery.
Pressure (psia)	95	
Flow rate (lb·mol/h)	250.27	
H_2S (mol fraction)	0.2358	
Stream 5A: Rich methanol from the absorber		
Temperature (°F)	-23.04	Rich methanol, which contains H_2S and other soluble gases, is withdrawn from bottom of tower. Temperature of solvent rises because of heat of absorption of H_2S into methanol.
Pressure (psia)	450	
Flow rate (lb·mol/h)	4,734.79	
H_2S (mol fraction)	0.0139	
Stream 5B: Recycle to absorption tower		
Temperature (°F)	-23.04	Rich methanol is flashed to pressure of 300 psia to desorb gases like H_2 and CH_4, and the desorbed gases are recycled to absorption tower.
Pressure (psia)	300	
Flow rate (lb·mol/h)	391.90	
H_2S (mol fraction)	0.0168	
Stream 6A: Lean methanol from stripper		
Temperature (°F)	149	Lean methanol from stripper bottom is to be circulated to absorption tower. The H_2S content in lean methanol is 0.0001 moles of H_2S per mole of methanol.
Pressure (psia)	14.7	
Flow rate (lb·mol/h)	4,129.96	
H_2S (mol fraction)	0.0001	

TABLE 6.3 (Cont.)

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 6B: Lean methanol from circulation pump		
Temperature (°F)	152.9	Lean methanol from stripper is at pressure of 14.7 psia and is pressurized to absorption tower operating pressure of 456 psia by using circulation pump.
Pressure (psia)	456	
Flow rate (lb·mol/h)	4,129.96	
H_2S (mol fraction)	0.0001	
Stream 6C: Lean methanol from heat exchanger		
Temperature (°F)	-10	Lean methanol from circulating pump is cooled against cold rich methanol from flash drum 3 to temperature of −10°F. Small portion of methanol is injected into feed gas prior to absorption to prevent icing of water.
Pressure (psia)	456	
Flow rate (lb·mol/h)	4,114.89	
H_2S (mol fraction)	0.0001	
Stream 6E: Lean methanol to stripper		
Temperature (°F)	-70	Lean methanol from heat exchanger is further cooled to temperature of −70°F by refrigeration.
Pressure (psia)	456	
Flow rate (lb·mol/h)	4,114.89	
H_2S (mol fraction)	0.0001	
Stream 7A: Methanol makeup		
Temperature (°F)	70	Methanol has low vapor pressure; hence, it is lost in stripper along with H_2S. Also, some methanol is lost in distillation column along with wastewater.
Pressure (psia)	14.7	
Flow rate (lb·mol/h)	21.48	
H_2S (mol fraction)	0.0	

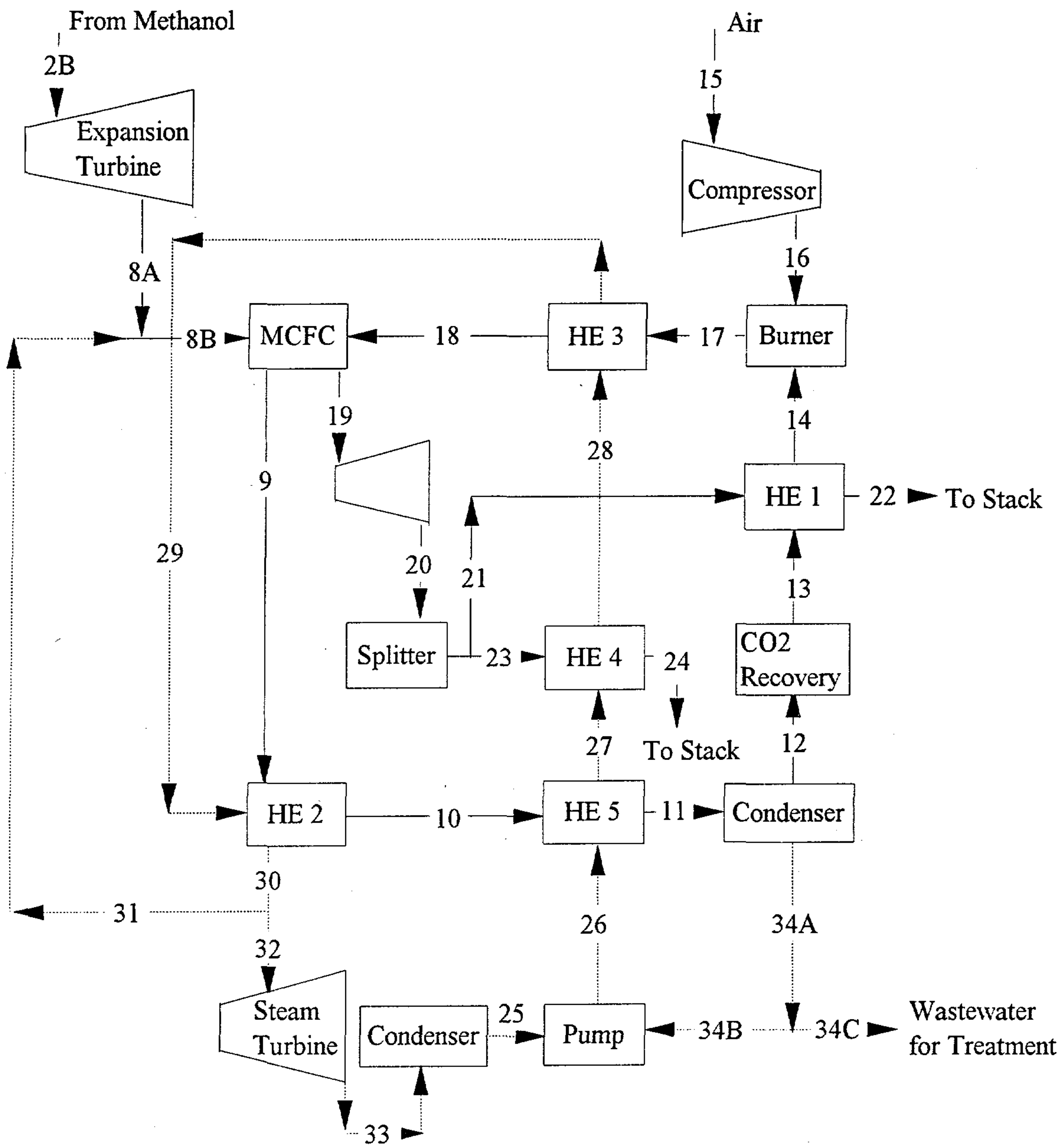

FIGURE 6.3 Flow Diagram of Fuel Cell System and Associated Heat Recovery in Case 3

TABLE 6.4 Stream Flows of Molten Carbonate Fuel Cell System in Case 3

Stream Data	Stream 2B	Stream 8A	Stream 8B	Stream 9	Stream 10	Stream 11
Description of stream	Feed gas from methanol	Fuel gas from expansion turbine	Fuel gas to fuel cell	Fuel cell anode exhaust	Gases from heat exchanger 2	Gases from heat exchanger 5
Gases (lb·mol/h)						
CO	4,530.65	4,530.65	4,530.65	1,812.26	1,812.26	1,812.26
CO_2	267.08	267.08	267.08	8,724.66	8,724.66	8,724.66
H_2	2,311.51	2,311.51	2,311.51	1,693.50	1,693.50	1,693.50
H_2O	0.00	0.00	12,000.00	13,579.13	13,579.13	13,579.13
N_2	36.44	36.44	36.44	36.44	36.44	36.44
Ar	72.73	72.73	72.73	72.73	72.73	72.73
CH_4	480.56	480.56	480.56	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	0.8 ppm	0.8 ppm	0.8 ppm	0.00	0.00	0.00
HCN	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	0.1 ppm	0.1 ppm	0.1 ppm	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	7,698.98	7,698.98	19,698.97	25,918.72	25,918.72	25,918.72
Liquids (lb·mol/h) H_2O	0.00	0.00	0.00	0.00	0.00	0.00
Temperature (°F)	80.00	-28.67	502.23	1,300.00	450.00	150.00
Pressure (psia)	450.00	150.00	150.00	150.00	150.00	150.00
Enthalpy of stream (Btu/h) (reference, 32°F)	2,596,057	-3,248,936	302,507,570	591,791,666	351,871,758	41,118,662

TABLE 6.4 (Cont.)

Stream Data	Stream 12	Stream 13	Stream 14	Stream 15	Stream 16	Stream 17
Description of stream	Gases to glycol process	Gases from glycol process	Gases from heat exchanger 1	Air to compressor	Air from compressor	Gases from burner
Gases (lb·mol/h)						
CO	1,812.26	1,794.14	1,794.14	0.00	0.00	0.00
CO_2	8,724.66	3,926.10	3,926.10	0.00	0.00	5,720.23
H_2	1,693.50	1,688.08	1,688.08	0.00	0.00	0.00
H_2O	78.96	0.00	0.00	0.00	0.00	1,688.08
N_2	36.44	35.71	35.71	44,202.24	44,202.24	44,237.95
Ar	72.73	72.73	72.73	543.05	543.05	615.78
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_2S	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	11,822.71	11,822.71	10,081.60
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	12,418.55	7,516.76	7,516.76	56,568.00	56,568.00	62,343.64
Liquids (lb·mol/h) H_2O	0.00	0.00	0.00	0.00	0.00	0.00
Temperature (°F)	70.00	56.14	600.00	81.00	713.05	1,411.87
Pressure (psia)	150.00	145.00	145.00	14.70	150.00	150.00
Enthalpy of stream (Btu/h) (reference, 32°F)	3,933,703	1,441,106	36,784,141	19,042,871	275,242,646	706,519,291

TABLE 6.4 (Cont.)

Stream Data	Stream 18	Stream 19	Stream 20	Stream 21	Stream 22	Stream 23
Description of stream	Gases from heat exchanger 3	Fuel cell cathode exhaust	Gases from expansion turbine	Gases from splitter to heat exchanger 1	Gases from heat exchanger 1	Gases from splitter to heat exchanger 4
Gases (lb·mol/h)						
CO	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	5,720.23	461.60	461.60	73.82	73.82	387.78
H_2	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	1,688.08	1,688.08	1,688.08	269.95	269.95	1,418.13
N_2	44,237.95	44,237.95	44,237.95	7,074.39	7,074.39	37,163.55
Ar	615.78	615.78	615.78	98.47	98.47	517.31
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_2S	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	10,081.60	7,452.29	7,452.29	1,191.75	1,191.75	6,260.54
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	62,343.64	54,455.70	54,455.70	8,708.38	8,708.38	45,747.31
Liquids (lb·mol/h) H_2O	0.00	0.00	0.00	0.00	0.00	0.00
Temperature (°F)	980.33	1,300.00	667.23	667.23	100.00	667.23
Pressure (psia)	150.00	150.00	14.70	14.70	14.70	14.70
Enthalpy of stream (Btu/h) (reference, 32°F)	482,306,421	546,008,492	280,696,393	44,888,086	9,545,050	235,808,307

TABLE 6.4 (Cont.)

Stream Data	Stream 24	Stream 25	Stream 26	Stream 27	Stream 28	Stream 29
Description of stream	Gases from heat exchanger 4	Water from condenser	Water from pump	Steam from heat exchanger 5	Steam from heat exchanger 4	Steam from heat exchanger 3
Gases (lb·mol/h)						
CO	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	387.78	0.00	0.00	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	1,418.13	0.00	0.00	9,917.77	15,668.16	30,055.55
N_2	37,163.55	0.00	0.00	0.00	0.00	0.00
Ar	517.31	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_2S	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	6,260.54	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	45,747.31	0.00	0.00	9,917.77	15,668.16	30,055.55
Liquids (lb·mol/h) H_2O	0.00	24,215.42	36,215.42	26,297.65	20,547.25	6,159.87
Temperature (°F)	400.00	121.36	121.36	356.77	356.77	356.77
Pressure (psia)	14.70	1.76	146.96	146.96	146.96	146.96
Enthalpy of stream (Btu/h) (reference, 32°F)	146,194,205	38,996,150	58,258,290	368,957,384	458,571,485	682,784,353

TABLE 6.4 (Cont.)

Stream Data	Stream 30	Stream 31	Stream 32	Stream 33	Stream 34A	Stream 34B	Stream 34C
Description of stream	Steam from heat exchanger 2	Steam for heating feed to fuel cell	Steam to steam turbine	Steam turbine exhaust	Water from condenser	Makeup water to pump	Wastewater for treatment
Gases (lb·mol/h)							
CO	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	36,215.42	12,000.00	24,215.42	22,813.57	0.00	0.00	0.00
N_2	0.00	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	0.00
CH_4	0.00	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	0.00
H_2S	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	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	0.00
COS	0.00	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	0.00
Total gas flow	36,215.42	12,000.00	24,215.42	22,813.57	0.00	0.00	0.00
Liquids (lb·mol/h) H_2O	0.00	0.00	0.00	1,401.85	13,500.17	12,000.00	1,500.17
Temperature (°F)	775.00	775.00	775.00	121.36	70.00	70.00	70.00
Pressure (psia)	146.96	146.96	146.96	1.76	150.00	150.00	150.00
Enthalpy of stream (Btu/h) (reference, 32°F)	922,758,266	305,756,514	617,001,752	459,798,748	9,234,116	8,208,000	1,026,116

TABLE 6.5 Descriptions of Streams of Fuel Cell System in Case 3

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 2B: Sulfur-free gas from H_2S section		
Temperature (°F)	80	The synthesis gas is cleaned in two stages. Sulfur compounds are removed before the gas is fed to the fuel cell.
Pressure (psia)	450	
Flow rate (lb·mol/h)	7,698.98	
CO_2 (mole fraction)	0.0347	
CO (mole fraction)	0.5885	
Stream 8A: Expanded gases from expansion turbine		
Temperature (°F)	-28.67	Sulfur-free gases are expanded through an expansion turbine for power recovery to a pressure suitable for fuel cell operation.
Pressure (psia)	150	
Flow rate (lb·mol/h)	7,698.98	
CO_2 (mole fraction)	0.0347	
CO (mole fraction)	0.5885	
Stream 8B: Feed to fuel cell anode		
Temperature (°F)	502.23	The expanded gases are heated by direct steam injection to temperature of 502.23°F. Direct injection of steam will increase the conversion of CO and also prevent the deposition of carbon on fuel cell anode.
Pressure (psia)	150	
Flow rate (lb·mol/h)	19,698.97	
CO_2 (mole fraction)	0.0136	
CO (mole fraction)	0.2300	
Stream 9: Fuel cell anode exhaust		
Temperature (°F)	1300	The composition of the gases corresponds to 100% conversion of CH_4 and 60% conversion of H_2 and CO. The temperature of gases is determined by energy balance.
Pressure (psia)	150	
Flow rate (lb·mol/h)	25,918.72	
CO_2 (mole fraction)	0.3366	
CO (mole fraction)	0.0699	
Stream 10: CO_2-rich gases from heat exchanger 2		
Temperature (°F)	450	The hot anode exhaust gases are cooled to a temperature of 450°F in heat exchanger 2 to raise high steam for bottoming cycle.
Pressure (psia)	150	
Flow rate (lb·mol/h)	25,918.72	
CO_2 (mole fraction)	0.3366	
CO (mole fraction)	0.0699	

TABLE 6.5 (Cont.)

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 11: CO_2-rich gases from heat exchanger 5		
Temperature (°F)	150	The anode exhaust gases are further cooled in heat exchanger 5 to a temperature of 150°F. The heat is utilized for preheating water for steam cycle. The amount of water vapor in the gases corresponds to the water's vapor pressure.
Pressure (psia)	150	
Flow rate (lb·mol/h)	25,918.98	
CO_2 (mole fraction)	0.3366	
CO (mole fraction)	0.0699	
Stream 12: Feed gas to CO_2 recovery		
Temperature (°F)	70	CO_2-rich gases are cooled in a condenser to knock out the water vapor from the gases.
Pressure (psia)	150	
Flow rate (lb·mol/h)	12,418.55	
CO_2 (mole fraction)	0.7026	
CO (mole fraction)	0.1459	
Stream 13: CO_2-lean gases from CO_2 recovery section		
Temperature (°F)	56.14	Fuel cell cathode takes CO_2 as its feed; therefore, the CO_2-lean gases along with unconverted CO and H_2 are fed back to the fuel cell system.
Pressure (psia)	145	
Flow rate (lb·mol/h)	7,516.76	
CO_2 (mole fraction)	0.5223	
CO (mole fraction)	0.2387	
Stream 14: CO_2-lean gases from heat exchanger 1		
Temperature (°F)	600	The CO_2-lean gases from CO_2 recovery section are heated with part of the cathode exhaust gases to a temperature of 600°F.
Pressure (psia)	145	
Flow rate (lb·mol/h)	7,516.76	
CO_2 (mole fraction)	0.5223	
CO (mole fraction)	0.2387	
Stream 15: Air to air compressor		
Temperature (°F)	81	The cathode reaction involves both O_2 and CO_2. The O_2 is supplied by air. Also air is supplied to burn unconverted CO and H_2.
Pressure (psia)	14.7	
Flow rate (lb·mol/h)	56,568	
CO_2 (mole fraction)	0.0000	
CO (mole fraction)	0.0000	

TABLE 6.5 (Cont.)

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 16: Compressed air from air compressor		
Temperature (°F)	713.05	The air is compressed to the operating pressure of the fuel cell.
Pressure (psia)	150	
Flow rate (lb·mol/h)	56,568	
CO_2 (mole fraction)	0.0000	
CO (mole fraction)	0.0000	
Stream 17: Gases from combustion chamber		
Temperature (°F)	1,411.87	The composition of gases is based on the composition of gases from CO_2 recovery and air from compressor. The temperature is adiabatic temperature.
Pressure (psia)	150	
Flow rate (lb·mol/h)	62,343.64	
CO_2 (mole fraction)	0.0918	
CO (mole fraction)	0.0000	
Stream 18: Fuel cell cathode feed		
Temperature (°F)	980.33	The gases from the combustion chamber are cooled to a suitable temperature of the fuel cell in heat exchanger 3.
Pressure (psia)	150	
Flow rate (lb·mol/h)	62,343.64	
CO_2 (mole fraction)	0.0918	
CO (mole fraction)	0.0000	
Stream 19: Fuel cell cathode exhaust		
Temperature (°F)	1,300	Part of the CO_2 in the cathode feed is consumed by cathode reaction. The temperature of gases is by energy balance.
Pressure (psia)	150	
Flow rate (lb·mol/h)	54,455.70	
CO_2 (mole fraction)	0.0085	
CO (mole fraction)	0.0000	
Stream 20: Cathode exhaust from expansion turbine		
Temperature (°F)	667.23	High-temperature cathode exhaust gases are expanded in expansion turbine to recover power.
Pressure (psia)	14.70	
Flow rate (lb·mol/h)	54,455.70	
CO_2 (mole fraction)	0.0085	
CO (mole fraction)	0.0000	

TABLE 6.5 (Cont.)

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 21: Cathode exhaust		
Temperature (°F)	667.23	The gases from the expansion turbine are at 667°F. Part of this gas stream is used in heating the gases from the CO_2 recovery system.
Pressure (psia)	14.70	
Flow rate (lb·mol/h)	8,708.38	
CO_2 (mole fraction)	0.0085	
CO (mole fraction)	0.0000	
Stream 22: Exhaust to stack		
Temperature (°F)	100	-----
Pressure (psia)	14.70	
Flow rate (lb·mol/h)	8,708.38	
CO_2 (mole fraction)	0.0085	
CO (mole fraction)	0.0000	
Stream 23: Cathode exhaust		
Temperature (°F)	667.23	The second portion of the cathode exhaust is utilized in raising the temperature of water for the steam cycle.
Pressure (psia)	14.70	
Flow rate (lb·mol/h)	45,747.31	
CO_2 (mole fraction)	0.0085	
CO (mole fraction)	0.0000	
Stream 24: Exhaust to stack		
Temperature (°F)	400	-----
Pressure (psia)	14.70	
Flow rate (lb·mol/h)	45,747.31	
CO_2 (mole fraction)	0.0085	
CO (mole fraction)	0.0000	
Stream 25: Water from steam condenser		
Temperature (°F)	121.36	Water from the steam condenser is for steam cycle.
Pressure (psia)	1.76	
Flow rate (lb·mol/h)	38,996,150	
Quality	0	
Stream 26: Water from pump		
Temperature (°F)	121.36	Water from the pump is for steam cycle.
Pressure (psia)	146.96	
Flow rate (lb·mol/h)	38,996,150	
Quality	0	

TABLE 6.5 (Cont.)

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 27: Steam from heat exchanger 5		
Temperature (°F)	356.77	Steam is from heat exchanger 5.
Pressure (psia)	146.96	
Flow rate (lb·mol/h)	38,996,150	
Quality	0.2738	
Stream 28: Steam from heat exchanger 4		
Temperature (°F)	356.77	Steam is from heat exchanger 4.
Pressure (psia)	146.96	
Flow rate (lb·mol/h)	38,996,150	
Quality	0.4326	
Stream 29: Steam from heat exchanger 3		
Temperature (°F)	356.77	Steam is from heat exchanger 3.
Pressure (psia)	146.96	
Flow rate (lb·mol/h)	38,996,150	
Quality	0.8299	
Stream 30: Steam from heat exchanger 2		
Temperature (°F)	775	Superheated steam is from heat exchanger 2.
Pressure (psia)	146.96	
Flow rate (lb·mol/h)	38,996,150	
Quality	1	
Stream 31: Steam for heating fuel cell feed		
Temperature (°F)	775	Superheated steam is used for heating the fuel cell feed.
Pressure (psia)	146.96	
Flow rate (lb·mol/h)	12,000	
Quality	1	
Stream 32: Superheated steam to steam turbine		
Temperature (°F)	775	Superheated steam goes to steam turbine for power recovery.
Pressure (psia)	146.96	
Flow rate (lb·mol/h)	4,215.42	
Quality	1	

TABLE 6.5 (Cont.)

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 33: Expanded steam from steam turbine		
Temperature (°F)	121.36	-----
Pressure (psia)	1.76	
Flow rate (lb·mol/h)	24,215.42	
Quality	0.9421	
Stream 34A: Condensate from anode exhaust condenser		
Temperature (°F)	70	-----
Pressure (psia)	150	
Flow rate (lb·mol/h)	13,500.17	
Quality	0	
Stream 34B: Makeup water to steam cycle pump		
Temperature (°F)	70	-----
Pressure (psia)	150	
Flow rate (lb·mol/h)	12,000	
Quality	0	
Stream 34C: Wastewater for treatment		
Temperature (°F)	70	-----
Pressure (psia)	150	
Flow rate (lb·mol/h)	1,500.17	
Quality	0	

6.4 Glycol Process for CO_2 Recovery

Figure 6.4 is an overall flow diagram of a glycol-based CO_2 recovery system. It is similar to the glycol system described in Section 4. In this system, the CO_2 is absorbed under pressure in a low-temperature glycol absorber. The pressure is released through a hydraulic turbine and in a series of flash tanks. The first tank in that series, the slump tank, allows for recovery of hydrogen from the rich absorbent. The subsequent tanks release CO_2 for disposal. The use of a series of tanks reduces the compression requirement. Table 6.6 is a line list corresponding to Figure 6.4. Stream descriptions and associated assumptions are provided in Table 6.7.

6.5 Fuel Cell, Steam Cycle, and Plant Performance

Use of the fuel cell topping cycle with methanol-based H_2S recovery and glycol-based CO_2 recovery results in a net plant output of 340 MW, 18% less than in the base case plant without CO_2 recovery. Table 6.8 lists the topping cycle output, steam cycle output, and internal plant consumption for the base case (no CO_2 recovery) and for the current case, Case 3. The most significant losses are the consumption of power for CO_2 compression and reduced steam cycle output.

6.6 Economics

Details of the capital investment estimates for the H_2S recovery system, the fuel cell system, and the CO_2 recovery system are presented in Tables 6.9, 6.10, and 6.11, respectively. A summary of capital costs, including indirect capital investment, operating, and maintenance costs, is provided in Section 9.

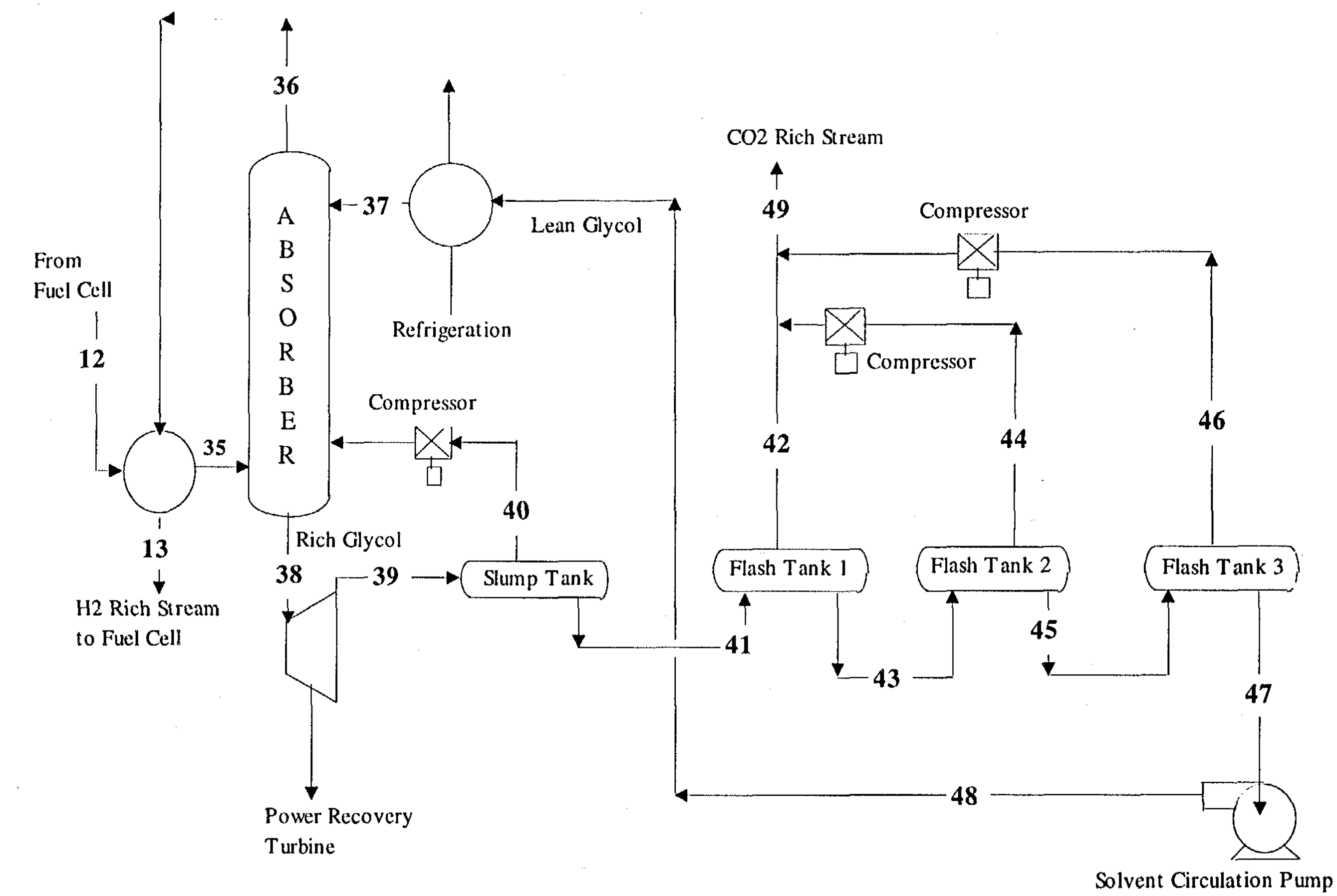

FIGURE 6.4 Flow Diagram of Glycol Process for CO_2 Recovery and Chilled Methanol Process for H_2S Recovery with Fuel Cell Topping Cycle in Case 3

TABLE 6.6 Stream Flows of Glycol System for CO_2 Removal in Case 3

Stream Data	Stream 12	Stream 13	Stream 35	Stream 36	Stream 37	Stream 38
Description of stream	Feed gas from fuel cell system	H_2-rich gas after heat exchanger	Absorber feed	Clean fuel gas from absorber	Lean glycol solvent to absorber	Rich glycol from absorber
Gases (lb·mol/h)						
CO	1,812.16	1,794.14	1,812.16	1,794.14	0.00	181.23
CO_2	8,724.66	3,926.10	8,724.66	3,926.10	210.13	5,110.91
	1,693.50	1,688.08	1,693.50	1,688.08	10.41	158.18
H_2O	78.96	0.00	78.96	0.00	0.00	79.76
N_2	36.44	35.71	36.44	35.71	0.00	7.29
Ar	72.73	72.73	72.73	72.73	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_2S	0.01	0.00	0.01	0.00	0.00	0.01
HCN	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	12,418.46	7,516.76	12,418.46	7,516.76	220.54	5,537.38
Liquids (lb·mol/h) Glycol solvent	0.00	0.00	0.00	0.00	20,802.84	20,802.84
Temperature (°F)	70.00	56.14	55.00	30.00	30.00	50.36
Pressure (psia)	150.00	145.00	150.00	145.00	150.00	145.00
Enthalpy of stream (Btu/h) (reference, 32°F)	3,933,489	1,441,041	2,373,591	-119,000	-5,712,163	53,306,378

TABLE 6.6 (Cont.)

Stream Data	Stream 39	Stream 40	Stream 41	Stream 42	Stream 43	Stream 44
Description of stream	Rich glycol solvent after turbine 1	Gases from slump tank	Rich glycol to flash tank 1	CO_2-rich gas from flash tank 1	Rich glycol to flash tank 2	CO_2-rich gas from flash tank 2
Gases (lb·mol/h)						
CO	181.23	163.10	18.12	18.12	0.00	0.00
CO_2	5,110.91	102.22	5,008.69	3,756.52	1,252.17	876.52
H_2	158.18	142.36	15.82	1.58	14.24	2.14
H_2O	79.76	0.80	78.96	78.96	0.00	0.00
N_2	7.29	6.56	0.73	0.73	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_2S	0.01	0.00	0.01	0.00	0.01	0.00
HCN	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	5,537.38	415.04	5,122.33	3,855.91	1,266.42	878.66
Liquids (lb·mol/h) Glycol solvent	20,802.84	0.00	20,802.84	0.00	20,802.84	0.00
Temperature (°F)	49.97	49.57	49.57	34.81	34.81	31.33
Pressure (psia)	50.00	50.00	50.00	25.00	25.00	14.70
Enthalpy of stream (Btu/h) (reference, 32°F)	52,160,172	53,779	50,942,277	95,822	8,065,266	-5,201

TABLE 6.6 (Cont.)

Stream Data	Stream 45	Stream 46	Stream 47	Stream 48	Stream 49
Description of stream	Rich glycol to flash tank 3	CO_2-rich gas from flash tank 3	Lean glycol to solvent circulation pump	Lean glycol from pump	Rich CO_2 gas product
Gases (lb·mol/h)					
CO	0.00	0.00	0.00	0.00	18.12
CO_2	375.65	162.68	212.98	212.98	4,795.72
H_2	12.10	1.54	10.56	10.56	5.26
H_2O	0.00	0.00	0.00	0.00	78.96
N_2	0.00	0.00	0.00	0.00	0.73
Ar	0.00	0.00	0.00	0.00	0.00
CH_4	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00
H_2S	0.01	0.00	0.01	0.01	0.00
HCN	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00
Total gas flow	387.76	164.22	223.55	223.55	4,898.79
Liquids (lb·mol/h) Glycol solvent	20,802.84	0.00	20,802.84	20,802.84	0.00
Temperature (°F)	31.33	30.68	30.68	31.83	32.44
Pressure (psia)	14.70	4.00	4.00	150.00	25.00
Enthalpy of stream (Btu/h) (reference, 32°F)	-1,911,810	-1,911	-3,762,523	-497,860	1,156,091

TABLE 6.7 Descriptions of Streams of Glycol Process for CO_2 Removal in Case 3

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 12: CO_2-rich gas from fuel cell system		
Temperature (°F)	70	The synthesis gas is cleaned in two stages. First, sulfur compounds are removed with chilled methanol. Then they are fed to another absorption column for CO_2 recovery.
Pressure (psia)	150	
Flow rate (lb·mol/h)	12,418.46	
CO_2 (mole fraction)	0.7026	
Stream 35: Feed gas to absorber		
Temperature (°F)	55	The CO_2-rich gas is cooled against the cold fuel gas from the top of the absorber to a temperature of 55°F.
Pressure (psia)	150	
Flow rate (lb·mol/h)	12,418.46	
CO_2 (mole fraction)	0.7026	
Stream 36: Fuel gas from absorber		
Temperature (°F)	30	The composition of this stream corresponds to a CO_2-removal efficiency of 55%. Also, other gases like H_2S, COS, and H_2 are absorbed by the solvent. The temperature of this stream is close to the temperature of lean solvent entering the absorber at the top.
Pressure (psia)	145	
Flow rate (lb·mol/h)	7,516.76	
CO_2 (mole fraction)	0.5223	
Stream 13: Fuel gas after heat exchanger		
Temperature (°F)	56.14	Fuel gas is heated against the CO_2-rich gases from the fuel cell section.
Pressure (psia)	145	
Flow rate (lb·mol/h)	7,516.76	
CO_2 (mole fraction)	0.5223	
Stream 37: Lean glycol to the of absorber		
Temperature (°F)	30	Lean glycol solvent contains residual CO_2. 50% excess solvent is used. The solvent is cooled to 30°F by refrigeration.
Pressure (psia)	150	
Flow rate (lb·mol/h)	21,023.38	
CO_2 (mole fraction)	0.01	

TABLE 6.7 (Cont.)

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 38: Rich glycol solvent from absorber		
Temperature (°F)	50.36	Flow rate reflects lean glycol solvent plus absorbed CO_2, H_2S, and other gases. The temperature increases because of the heat of absorption of CO_2 and H_2S.
Pressure (psia)	145	
Flow rate (lb·mol/h)	26,340.22	
CO_2 (mole fraction)	0.1940	
Stream 39: Rich glycol solvent from turbine 1		
Temperature (°F)	49.97	This stream is the exit stream from power recovery turbine. Exit pressure has been selected to avoid release of CO_2 and H_2S while allowing some recovery of work of pressurization. The change in temperature over the turbine is estimated from change in enthalpy, which is taken to be equal to flow work.
Pressure (psia)	50	
Flow rate (lb·mol/h)	26,340.22	
CO_2 (mole fraction)	0.1940	
Stream 40: Flash gas		
Temperature (°F)	49.57	CO_2 and H_2S are released from the glycol solvent in the slump tank. This stream is compressed and recycled to the absorber to decrease the losses of valuable gases like H_2 and CO.
Pressure (psia)	50	
Flow rate (lb·mol/h)	415.04	
CO_2 (mole fraction)	0.2463	
Stream 41: Rich glycol to high-pressure flash tank 1		
Temperature (°F)	49.57	The CO_2 from the rich glycol solvent is released in stages.
Pressure (psia)	50	
Flow rate (lb·mol/h)	25,925.17	
CO_2 (mole fraction)	0.1932	
Stream 42: CO_2-rich flash gas from high-pressure flash tank		
Temperature (°F)	34.81	In first stage, the gases are flashed to a pressure of 25 psia. The amount of CO_2 remaining in the solvent depends on pressure, and the CO_2 released is calculated by mass balance.
Pressure (psia)	25	
Flow rate (lb·mol/h)	3,855.91	
CO_2 (mole fraction)	0.9742	

TABLE 6.7 (Cont.)

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 43: Glycol solvent from high-pressure flash tank		
Temperature (°F)	34.81	-----
Pressure (psia)	25	
Flow rate (lb·mol/h)	22,069.26	
CO_2 (mole fraction)	0.0567	
Stream 44: CO_2-rich flash gas from intermediate-pressure flash tank		
Temperature (°F)	31.33	The amount of CO_2 in solvent and released as gas is calculated as in stream 42. Sufficient residence is provided for the gases to separate from solvent. This determines tank volume.
Pressure (psia)	14.70	
Flow rate (lb·mol/h)	878.66	
CO_2 (mole fraction)	0.9976	
Stream 45: Glycol solvent from intermediate-pressure flash tank		
Temperature (°F)	31.33	-----
Pressure (psia)	14.7	
Flow rate (lb·mol/h)	21,190.60	
CO_2 (mole fraction)	0.0177	
Stream 46: CO_2-rich flash gas from low-pressure flash tank		
Temperature (°F)	30.68	Glycol solvent is flashed to a pressure of 4 psia to remove as much CO_2 as possible. The lower residual amount of CO_2 in lean glycol solvent reduces the circulation rate of solvent.
Pressure (psia)	4.0	
Flow rate (lb·mol/h)	164.22	
CO_2 (mole fraction)	0.9906	
Stream 47: Lean glycol solvent from low-pressure flash tank		
Temperature (°F)	30.68	-----
Pressure (psia)	4.0	
Flow rate (lb·mol/h)	21,026.38	
CO_2 (mole fraction)	.0101	

TABLE 6.7 (Cont.)

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 48: Lean glycol solvent after circulation pump		
Temperature (°F)	31.83	The lean solvent is pressurized to the absorber operating pressure by using a pump. The change in temperature results from work of compression.The solvent is chilled before being sent to the absorber.
Pressure (psia)	150	
Flow rate (lb·mol/h)	21,026.38	
CO_2 (mole fraction)	0.0101	
Stream 49: CO_2-rich product gas		
Temperature (°F)	32.44	Flash gases from intermediate- and low-pressure flash tanks are compressed to the pressure of stream 42. Streams 42, 44, and 46 are combined for further compression for pipeline.
Pressure (psia)	25.0	
Flow rate (lb·mol/h)	4,898.79	
CO_2 (mole fraction)	0.9790	

TABLE 6.8 Power Output, Plant Power Use, and Net Power Output for Base Case and Case 3 Fuel Cell/ Glycol Process

	Power (MW)	
Power Variable	Base Case	Fuel Cell Case
Power output		
Gas turbine or fuel cell	298.8	246.7
Steam turbine	159.4	171.8
Internal power consumption		
CO_2 recovery		
CO_2 compression	0	-24.9
Solvent circulation	0	-2.9
Solvent refrigeration	0	-1.3
Others	0	-0.4
Gasification system[a]	-44.7	-48.9
Net power output	413.5	340.1
Energy penalty	0	73.4

[a] Includes H_2S recovery system energy use.

TABLE 6.9 Sizing and Cost Estimation for Major Equipment Used for H_2S Removal in Chilled Methanol Process in Case 3

1.	**Gas-Gas Heat Exchanger for Raw Gas Cooling**		
	a) with H_2S-Rich Gas		
	Q = Load (Btu/h)	103,571	
	Tha = Inlet temperature of hot fluid (°F)	104.55	
	Thb = Outlet temperature of hot fluid (°F)	-10	
	Pressure of hot gases (psia)	456	
	Tca = Inlet temperature of cold fluid (°F)	-29.9	
	Tcb = Outlet temperature of cold fluid (°F)	84.50	
	Delta T1	20.045	
	Delta T2	20	
	Log mean temperature difference (°F)	20	
	Overall heat transfer coefficient ($Btu/h/ft^2/°F$)	5	
	Heat transfer area (ft^2)	1,038	
	Operating pressure (psia)	275	
	Pressure factor	1.165	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction, shell and tube floating head)	$23,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$100,187**
	b) with H_2S-Lean Fuel Gas		
	Q = Load (Btu/h)	8,036,992	
	Tha = Inlet temperature of hot fluid (°F)	104.55	
	Thb = Outlet temperature of hot fluid (°F)	-34	
	Pressure of hot gases (psia)	456	
	Tca = Inlet temperature of cold fluid (°F)	-70	
	Tcb = Outlet temperature of cold fluid (°F)	80	
	Delta T1	24.545	
	Delta T2	36	
	Log mean temperature difference (°F)	30	
	Overall heat transfer coefficient ($Btu/h/ft^2/°F$)	5	
	Heat transfer area (ft^2)	53,888	
	Operating pressure (psia)	275	
	Pressure factor	1.165	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction, shell and tube floating head)	$350,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$1,524,577**

TABLE 6.9 (Cont.)

2.	**H_2S Absorption Column**		
	Diameter of tower (ft)	7	
	HETP (ft)	3	
	Number of theoretical stages	15	
	Absorber tower height (ft)	49	
	(4 ft for inlet, outlet and gas, and liquid distributions)		
	Volume of packing (ft^3)	1,733	
	Pressure factor	2.6	
	Cost per foot of column height	$950	
	(mild steel construction)		
	Materials correction factor	1	
	Module factor	4.16	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of absorber in 1995		**$588,291**
	Cost of packing per cubic foot	$63.5	
	(2 in. pall rings-metal)		
	Total cost of packing		**$110,014**
3.	**H_2S Stripping Column**		
	Diameter of tower (ft)	2.5	
	HETP (ft)	3	
	Number of theoretical stages	17	
	Absorber tower height (ft)	55	
	(4 ft for inlet, outlet and gas, and liquid distributions)		
	Volume of packing (ft^3)	250	
	Pressure factor	1	
	Cost per ft of column height	$500	
	(mild steel construction)		
	Module factor	4.16	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of absorber in 1995		**$133,669**
	Cost of packing per cubic foot	$63.5	
	(2-in. pall rings-metal)		
	Total cost of packing		**$15,903**

TABLE 6.9 (Cont.)

4.	**Flash Drum 1**		
	Methanol flow rate (lb/h)	123,450	
	Density of methanol (lb/gal)	6.55	
	Residence time (s)	180	
	Slump tank volume (gal)	942	
	Pressure factor	1	
	Module factor	2.08	
	Purchased cost of flash drum in 1987 (mild steel construction)	$5,200	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of flash drum in 1995		**$12,638**
5.	**Recycle Compressor**		
	Inlet pressure (psia)	300	
	Outlet pressure (psia)	456	
	Compressor size (hp)	72	
	Purchased cost of centrifugal compressor in 1987 (includes electric motor drive and gear reducer)	$32,000	
	Size factor	1	
	Module factor	2.6	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of compressor in 1995		**$97,214**
6.	**Flash Drum 2**		
	Methanol flow rate (lb/h)	123,450	
	Density of methanol (lb/gal)	6.55	
	Residence time (s)	180	
	Slump tank volume (gal)	942	
	Pressure factor	1	
	Module factor	2.08	
	Purchased cost of flash drum in 1987 (mild steel construction)	$5,200	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of flash drum in 1995		**$12,638**

TABLE 6.9 (Cont.)

7.	**Flash Drum 3**		
	Methanol flow rate (lb/h)	123,450	
	Density of methanol (lb/gal)	6.55	
	Residence time (s)	180	
	Slump tank volume (gal)	942	
	Pressure factor	1	
	Module factor	2.08	
	Purchased cost of flash drum in 1987 (mild steel construction)	$5,200	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of flash drum in 1995		**$12,638**
8.	**Flash Gas Compressor 1**		
	Inlet pressure (psia)	20.00	
	Outlet pressure (psia)	150.00	
	Compressor size (hp)	57	
	Purchased cost of centrifugal compressor in 1987 (includes electric motor drive and gear reducer)	$27,000	
	Size factor	1	
	Module factor	2.6	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of compressor in 1995		**$82,024**
9.	**Flash Gas Compressor 2**		
	Inlet pressure (psia)	14.70	
	Outlet pressure (psia)	150.00	
	Compressor size (hp)	154	
	Purchased cost of centrifugal compressor in 1987 (includes electric motor drive and gear reducer)	$60,000	
	Size factor	1	
	Module factor	2.6	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of compressor in 1995		**$182,276**
10.	**Solvent Circulation Pump**		
	Horse power	115	
	Size exponent	1	
	Purchased cost of pump in 1987 (includes motor, coupling, base; cast iron, horizontal)	$12,000	
	Module factor	1.5	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of solvent pump in 1995		**$21,032**

TABLE 6.9 (Cont.)

11.	**Lean-Rich Solvent Heat Exchanger**		
	Q = Load (Btu/h)	12,184,945	
	Tha = Inlet temperature of hot fluid (°F)	153	
	Thb = Outlet temperature of hot fluid (°F)	-10	
	Pressure of hot gases (psia)	20	
	Tca = Inlet temperature of cold fluid (°F)	-34	
	Tcb = Outlet temperature of cold fluid (°F)	129	
	Delta T1	24	
	Delta T2	24	
	Log mean temperature difference (°F)	24	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	150	
	Heat transfer area (ft^2)		
	Operating pressure (psia)	3,391	
	Pressure factor	456	
	Materials correction factor	1.175	
	Module factor	3.2	
	(includes all of the supporting equipment and connections and installation)		
	Purchased cost of heat exchanger in 1987	$54,000	
	(mild steel construction, shell and tube floating head)		
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$237,240**
12.	**Solvent Refrigeration**		
	Refrigeration (tons)	2,235	
	Purchased cost in 1987	$750,000	
	Temperature correction factor	3.5	
	Module factor	1.46	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of solvent refrigeration in 1995		**$4,478,037**
	Total Direct Cost		**$7,608,378**
	Total Direct Cost for Three Trains		**$22,825,134**

TABLE 6.10 Sizing and Cost Estimation for Major Equipment Used for Fuel Cell System in Case 3

1.	**Fuel Gas Expansion Turbine**		
	Turbine size (hp)	2,296	
	Purchased cost in 1979	$1,607,439	
	Module factor	1.00	
	CE index for process equipment in 1979	$256	
	CE index for process equipment in 1995	373.9	
	Installed cost of turbine in 1995		**$2,347,740**
2.	**Heat Exchanger 1**		
	Q = Load (Btu/h)	35,343,035	
	Tha = Inlet temperature of hot fluid (°F)	667.23	
	Thb = Outlet temperature of hot fluid (°F)	100	
	Pressure of hot gases (psia)	15	
	Tca = Inlet temperature of cold fluid (°F)	32.4	
	Tcb = Outlet temperature of cold fluid (°F)	600.00	
	Delta T1	67.2315	
	Delta T2	68	
	Log mean temperature difference (°F)	67	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	5	
	Heat transfer area (ft^2)	104,882	
	Operating pressure (psia)	145.00	
	Pressure factor	1.16	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction, shell and tube floating head)	$524,411	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$2,274,498**

TABLE 6.10 (Cont.)

3.	**Heat Exchanger 2**		
	Q = Load (Btu/h)	239,973,908	
	Tha = Inlet temperature of hot fluid (°F)	1300.00	
	Thb = Outlet temperature of hot fluid (°F)	450	
	Pressure of hot gases (psia)	150	
	Tca = Inlet temperature of cold fluid (°F)	356.8	
	Tcb = Outlet temperature of cold fluid (°F)	775.00	
	Delta T1	525	
	Delta T2	93	
	Log mean temperature difference (°F)	250	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	30	
	Heat transfer area (ft^2)	32,019	
	Operating pressure (psia)	146.96	
	Pressure factor	1.165	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction, shell and tube floating head)	$250,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$1,088,984**
4.	**Heat Exchanger 3**		
	Q = Load (Btu/h)	224,212,870	
	Tha = Inlet temperature of hot fluid (°F)	1411.87	
	Thb = Outlet temperature of hot fluid (°F)	980	
	Pressure of hot gases (psia)	150	
	Tca = Inlet temperature of cold fluid (°F)	356.8	
	Tcb = Outlet temperature of cold fluid (°F)	356.77	
	Delta T1	1055.102336	
	Delta T2	624	
	Log mean temperature difference (°F)	820	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	30	
	Heat transfer area (ft^2)	9,109	
	Operating pressure (psia)	146.96	
	Pressure factor	1.165	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction, shell and tube floating head)	$100,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$435,594**

TABLE 6.10 (Cont.)

5.	**Heat Exchanger 4**		
	Q = Load (Btu/h)	89,614,102	
	Tha = Inlet temperature of hot fluid (°F)	667.23	
	Thb = Outlet temperature of hot fluid (°F)	400	
	Pressure of hot gases (psia)	15	
	Tca = Inlet temperature of cold fluid (°F)	356.8	
	Tcb = Outlet temperature of cold fluid (°F)	356.79	
	Delta T1	310.4448363	
	Delta T2	43	
	Log mean temperature difference (°F)	136	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	30	
	Heat transfer area (ft^2)	22,038	
	Operating pressure (psia)	146.96	
	Pressure factor	1.165	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction, shell and tube floating head)	$180,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$784,068**
6.	**Heat Exchanger 5**		
	Q = Load (Btu/h)	310,699,095	
	Tha = Inlet temperature of hot fluid (°F)	450.00	
	Thb = Outlet temperature of hot fluid (°F)	150	
	Pressure of hot gases (psia)	150	
	Tca = Inlet temperature of cold fluid (°F)	121.4	
	Tcb = Outlet temperature of cold fluid (°F)	356.77	
	Delta T1	93.23133627	
	Delta T2	29	
	Log mean temperature difference (°F)	55	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	30	
	Heat transfer area (ft^2)	189,274	
	Operating pressure (psia)	146.96	
	Pressure factor	1.165	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction, shell and tube floating head)	$946,369	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$4,122,323**

TABLE 6.10 (Cont.)

7.	**Refrigeration**		
	Refrigeration (tons)	2,324	
	Purchased cost in 1987	700,000	
	Temperature correction factor		
	Module factor	1.46	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of solvent refrigeration in 1995		**$1,194,143**
8.	**Cathode Exhaust Gas Expansion Turbine**		
	Turbine size (hp)	104,234	
	Purchased cost in 1987	$10,435,839	
	Module factor	1.00	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of turbine in 1995		**$12,193,625**
9.	**Air Compressor for Fuel Cell**		
	Inlet pressure (psia)	14.70	
	Outlet pressure (psia)	150.00	
	Compressor size (MW)	225	
	Purchased cost in 1987	$24,446,768	
	Module factor	1.00	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost air compressor in 1995		**$28,564,520**
10.	**Steam Turbine**		
	Turbine output (MW)	172	

The cost of steam turbine is already included in the base case.

TABLE 6.10 (Cont.)

11.	**Condenser**		
	Q = Load (Btu/h)	420,802,598	
	Tha = Inlet temperature of hot fluid (°F)	121.36	
	Thb = Outlet temperature of hot fluid (°F)	121	
	Pressure of hot gases (psia)	2	
	Tca = Inlet temperature of cold fluid (°F)	70.0	
	Tcb = Outlet temperature of cold fluid (°F)	100.00	
	Delta T1	21.35924367	
	Delta T2	51	
	Log mean temperature difference (°F)	34	
	Overall heat transfer coefficient ($Btu/h/ft^2/°F$)	500	
	Heat transfer area (ft^2)	24,613	
	Operating pressure (psia)	146.96	
	Pressure factor	1.165	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction, shell and tube floating head)	$200,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$871,187**
12.	**Pump**		
	Horsepower	110	
	Size exponent	1	
	Purchased cost of pump in 1987 (includes motor, coupling, base; cast iron, horizontal)	$12,000	
	Module factor	1.5	
	CE index in 1987	320	
	CE index in 1995	373.9	
	Installed cost of solvent pump in 1995		**$21,032**
13.	**Fuel Cell Stack**		
	Fuel cell power output (kW)	77,952	
	Unit cost per kilowatt	$180	
	Total cost		**$14,031,388**

TABLE 6.10 (Cont.)

14.	**Fuel Cell Invertor**		
	Unit cost per kilowatt	$100	
	Total cost		**$7,795,216**
15.	**Fuel Cell Controls**		
	Unit cost per kilowatt	$140	
	Total cost		**$10,913,302**
16.	**Fuel Cell and Components Assembly**		
	Unit cost per kilowatt	$110	
	Total cost		**$8,574,737**
	Total Direct Cost		**$95,212,358**
	Total Direct Cost for Three Trains		**$285,637,074**

TABLE 6.11 Sizing and Cost Estimation for Major Equipment Used for CO_2 Removal in Glycol Process in Case 3

1.	**Gas-Gas Heat Exchanger**		
	Q = Load (Btu/h)	1,559,898	
	Tha = Inlet temperature of hot fluid (°F)	70.00	
	Thb = Outlet temperature of hot fluid (°F)	55	
	Pressure of hot gases (psia)	150.00	
	Tca = Inlet temperature of cold fluid (°F)	30.00	
	Tcb = Outlet temperature of cold fluid (°F)	56.14	
	Delta T1	13.8564	
	Delta T2	25	
	Log mean temperature difference (°F)	19	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	5	
	Heat transfer area (ft^2)	16,521	
	Operating pressure (psia)	150.00	
	Pressure factor	1.16	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction, shell and tube floating head)	$160,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1993	360.4	
	Installed cost of heat exchanger in 1993		**$693,958**
2.	**CO_2 Absorption Column**		
	Diameter of tower (ft)	16	
	HETP (ft)	3	
	Number of theoretical stages	12	
	Absorber tower height (ft) (4 ft for inlet, outlet and gas, and liquid distributions)	40	
	Volume of packing (ft^3)	7,241	
	Pressure factor	1	
	Cost per foot of column height (mild steel construction)	$1,400	
	Materials correction factor	1	
	Module factor	4.16	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1993	360.4	
	Installed cost of absorber in 1993		**$272,199**
	Cost of packing per cubic foot (2-in. pall rings-metal)	$63.5	
	Total cost of packing		**$459,813**

TABLE 6.11 (Cont.)

3.	**Power Recovery Turbine 1**		
	Turbine size (hp)	451	
	Purchased cost in 1979	$180,000	
	Module factor	1	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1993	360.4	
	Installed cost of solvent pump in 1993		**$210,319**
4.	**Slump Tank**		
	Glycol flow rate (lb/h)	5,824,796	
	Density of glycol (lb/gal)	8.6	
	Residence time (s)	180	
	Slump tank volume (gal)	33,865	
	Pressure factor	1.38	
	Materials correction factor	1	
	Module factor	2.08	
	Purchased cost of slump tank in 1987 (mild steel construction)	$65,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1993	360.4	
	Installed cost of slump tank in 1993		**$218,002**
5.	**Recycle Compressor**		
	Inlet pressure (psia)	50	
	Outlet pressure (psia)	150.00	
	Compressor size (hp)	259	
	Purchased cost of centrifugal compressor in 1987 (includes electric motor drive and gear reducer)	$95,000	
	Size factor	1	
	Materials correction factor	1	
	Module factor	2.6	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1993	360.4	
	Installed cost of compressor in 1993		**$288,604**

TABLE 6.11 (Cont.)

6.	**Flash Tank 1**		
	Glycol flow rate (lb/h)	5,824,796	
	Density of glycol (lb/gal)	8.6	
	Residence time (s)	180	
	Slump tank volume (gal)	33,865	
	Pressure factor	1	
	Materials correction factor	1	
	Module factor	2.08	
	Purchased cost of slump tank in 1987 (mild steel construction)	$65,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1993	360.4	
	Installed cost of slump tank in 1993		**$157,973**
7.	**Flash Tank 2**		
	Glycol flow rate (lb/h)	5,824,796	
	Density of glycol (lb/gal)	8.6	
	Residence time (s)	180	
	Slump tank volume (gal)	33,865	
	Pressure factor	1	
	Materials correction factor	1	
	Module factor	2.08	
	Purchased cost of slump tank in 1987 (mild steel construction)	$65,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1993	360.4	
	Installed cost of slump tank in 1993		**$157,973**
8.	**Flash Tank 3**		
	Glycol flow rate (lb/h)	5,824,796	
	Density of glycol (lb/gal)	8.6	
	Residence time (s)	180	
	Slump tank volume (gal)	33,865	
	Pressure factor	1	
	Materials correction factor	1	
	Module factor	2.08	
	Purchased cost of slump tank in 1987 (mild steel construction)	$65,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1993	360.4	
	Installed cost of slump tank in 1993		**$157,973**

TABLE 6.11 (Cont.)

9.	**Solvent Circulation Pump**		
	Horsepower	1,282	
	Size exponent	0.79	
	Purchased cost of 300-hp pump in 1987 (includes motor, coupling, base; cast iron, horizontal)	$30,000	
	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		**$165,631**
10.	**Compressor 1 for CO_2**		
	Inlet pressure (psia)	14.70	
	Outlet pressure (psia)	50.00	
	Compressor size (hp)	600.41	
	Purchased cost of centrifugal compressor in 1987 (includes electric motor drive and gear reducer)	$85,000	
	Size factor	1	
	Materials correction factor	1	
	Module factor	2.6	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1993	360.4	
	Installed cost of compressor in 1993		**$258,225**
11.	**Compressor 2 for CO_2**		
	Inlet pressure (psia)	4.00	
	Outlet pressure (psia)	50.00	
	Compressor size (hp)	120.54	
	Purchased cost of centrifugal compressor in 1987 (includes electric motor drive and gear reducer)	$50,000	
	Size factor	1	
	Materials correction factor	1	
	Module factor	2.6	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1993	360.4	
	Installed cost of compressor in 1993		**$151,897**
12.	**Solvent Refrigeration**		
	Refrigeration (tons)	434.53	
	Purchased cost in 1987	$230,000	
	Temperature correction factor	1.25	
	Module factor	1.46	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1993	360.4	
	Installed cost of solvent refrigeration in 1993		**$490,452**

TABLE 6.11 (Cont.)

13.	**CO_2 Product Gas Compressors**		
	Compressor 1 (hp)	2,913.98	
	Compressor 2 (hp)	2,913.98	
	Compressor 3 (hp)	2,913.98	
	Purchased cost of centrifugal compressor 1 in 1987	$750,000	
	Purchased cost of centrifugal compressor 2 in 1987	$750,000	
	Purchased cost of centrifugal compressor 3 in 1987 (includes electric motor drive and gear reducer)	$750,000	
	Size factor	1	
	Module factor	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		**2,278,453**
	Installed cost of compressor 2 in 1993		**2,278,453**
	Installed cost of compressor 3 in 1993		**2,278,453**
	Total Direct Cost		**$10,518,378**
	Total Direct Cost for Three Trains		**$31,555,133**

7 Case 4 — Fuel Cell Topping Cycle and Membrane CO_2 Recovery

Material and energy balances have been developed in this section for the application of an internal reforming molten carbonate fuel cell as the topping cycle for an IGCC plant. The CO_2 from the fuel cell exhaust is recovered by membrane separation. The analysis is very similar to that presented in Section 6, except the glycol-based absorption system is replaced with a membrane system.

7.1 Design Basis

Figure 7.1 provides an overview of the of the IGCC system, including the gasifier, gas treatment, the fuel cell, and the steam cycle. This system is identical to that represented in Figure 6.1 and is reproduced here for convenience. The overall design of the fuel cell is determined by the gasifier capacity and synthesis gas composition. These are assumed to be the same as in the base case, which has no CO_2 recovery. The fuel cell has very low tolerance for contaminants, including particulates and sulfur compounds. To achieve the required level of H_2S removal, a chilled methanol system has been employed rather than the glycol system used in the gas turbine cases. The chilled methanol system is designed to reduce the sulfur species (H_2S and COS) concentration to less than 1 ppmv. The reactions in the fuel cell anode shift the synthesis gas to a hydrogen-rich gas with a high concentration of CO_2 and reduce the resultant hydrogen with carbonate ion. Oxidation of the carbonate at the anode releases CO_2 and two electrons. The CO_2-rich anode exhaust is treated in a membrane recovery system to separate most of the CO_2. Thermal energy released by cooling this anode exhaust provides heat for the steam bottoming cycle. An expansion turbine is used on the cathode exhaust to extract energy.

Table 7.1 is a summary of principal material flows for the base case and for this design option. The CO_2 reduction accomplished at the power plant is 89% and is accompanied by a 24% reduction in net electrical output from the base case, which uses a gas turbine and no CO_2 recovery. A full accounting of the net CO_2 reduction would include CO_2 released in the generation of replacement power; mining, coal, and reagent preparation; and materials transport.

7.2 Chilled Methanol Process for H_2S Recovery

The design of the chilled methanol system is the same as that described for Case 3. It is required to provide adequate H_2S removal to meet fuel cell requirements. See Figure 6.2 and Tables 6.2 and 6.3 for details.

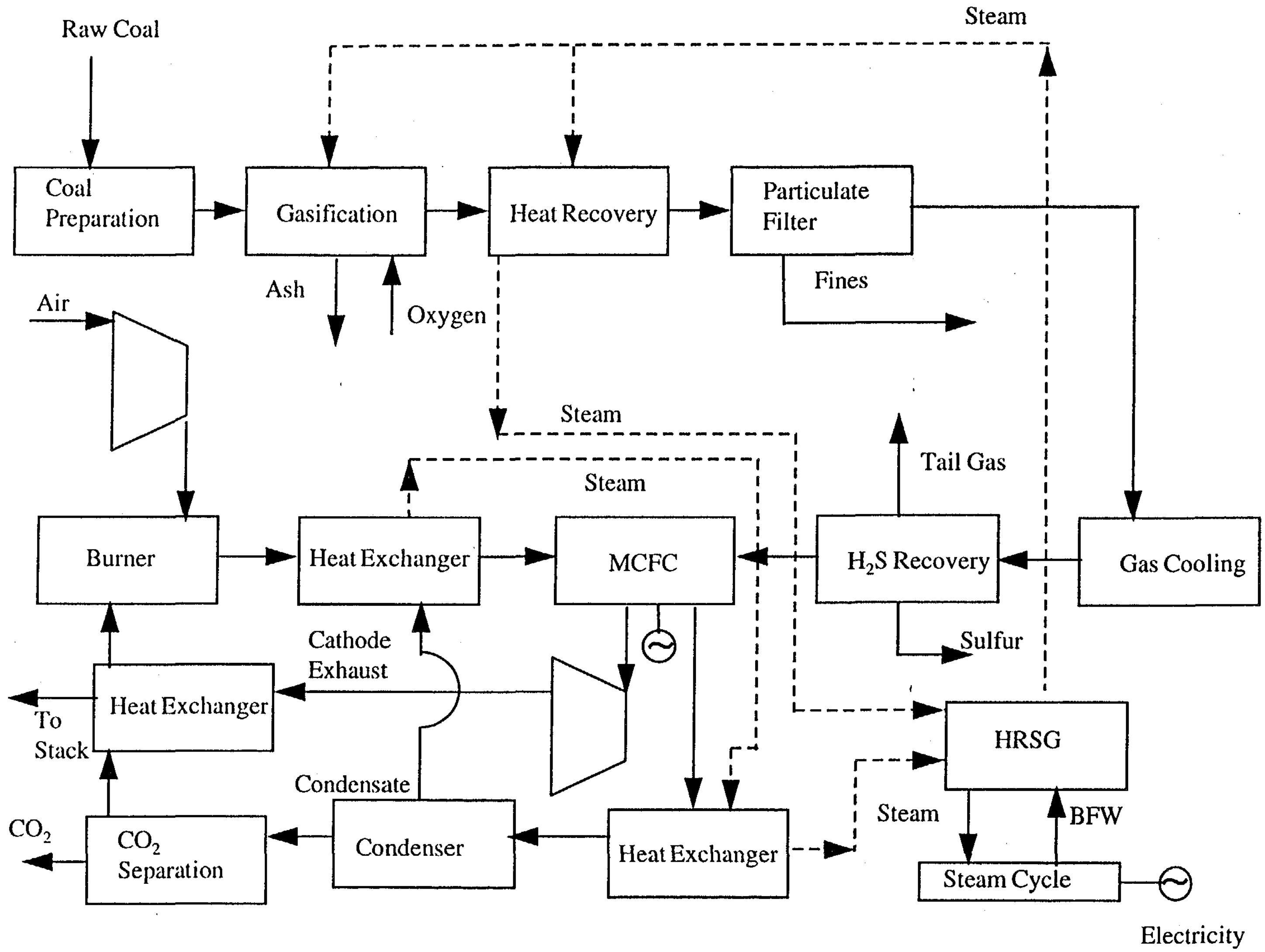

FIGURE 7.1 Block Diagram of the IGCC System with CO_2 Recovery Used in Cases 3 and 4

TABLE 7.1 Material Flows for Oxygen-Blown Base Case and Case 4

Material Flow (tons/d)	Base Case	Case 4
Coal (prepared)	3,845	3,845
Oxygen	2,347	2,347
Solid waste	492	492
Sulfur	78	78
CO_2 (power plant only)	9210	993
SO_2 (power plant only)	1.08	6.92
Net power output (MW)	413.5	313.77

7.3 Molten Carbonate Fuel Cell System

The molten carbonate fuel cell system with the membrane for CO_2 recovery is virtually identical to that described in section 6.3. A slight difference in stream composition following the membrane system reflects the performance difference between the two CO_2-recovery systems. Table 7.2 is an alternate line list for Figure 6.3 detailing these differences.

7.4 Membrane System for CO_2 Recovery

Figure 7.2 is an overall flow diagram of a membrane CO_2-recovery system. It is similar to the membrane system described in Section 5. Table 7.3 is a line list corresponding to Figure 7.2. Stream descriptions and associated assumptions are provided in Table 7.4.

7.5 Fuel Cell, Steam Cycle, and Plant Performance

Use of the fuel cell topping cycle with methanol-based H_2S recovery and membrane CO_2 recovery results in a net plant output of 314 MW, 24% less than in the base case plant without CO_2 recovery. Table 7.5 lists the topping cycle output, steam cycle output, and internal plant consumption for the base case (no CO_2 recovery) and for the current case, Case 4. The most significant losses are the consumption of power for CO_2 compression and power required for permeate compression between membrane stages.

7.6 Economics

Details of the capital investment estimates for the H_2S recovery system, the fuel cell system, and the CO_2 recovery system are presented in Tables 6.9, 7.6, and 7.7, respectively. A summary of capital costs, including indirect capital investment, operating, and maintenance costs, is provided in Section 9.

TABLE 7.2 Stream Flows of Molten Carbonate Fuel Cell System in Case 4

Stream Data	Stream 12	Stream 13	Stream 14	Stream 15	Stream 16	Stream 17
Description of stream	Gases to membrane process	Gases from membrane process	Gases from heat exchanger 1	Air to compressor	Air from compressor	Gases from burner
Gases (lb·mol/h)						
CO	1,812.26	1,539.74	1,539.74	0.00	0.00	0.00
CO_2	8,724.66	4,180.59	4,180.59	0.00	0.00	5,720.33
H_2	1,693.50	1,692.69	1,692.69	0.00	0.00	0.00
H_2O	135.79	24.61	24.61	0.00	0.00	1,717.29
N_2	36.44	35.82	35.82	44,039.07	44,039.07	44,074.89
Ar	72.73	62.12	62.12	541.05	541.05	603.17
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_2S	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	11,779.07	11,779.07	10,162.86
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	12,475.38	7,535.57	7,535.57	56,359.19	56,359.19	62,278.54
Liquids (lb·mol/h)						
H_2O	0.00	0.00	0.00	0.00	0.00	0.00
Temperature (°F)	150.00	154.88	600.00	81.00	713.05	1,355.19
Pressure (psia)	150.00	140.00	140.00	14.70	150.00	150.00
Enthalpy of stream (Btu/h) (reference, 32°F)	12,463,567	7,542,911	37,780,990	18,972,575	274,226,598	676,482,506

TABLE 7.2 (Cont.)

Stream Data	Stream 18	Stream 19	Stream 20	Stream 21	Stream 22	Stream 23
Description of stream	Gases from heat exchanger 3	Fuel cell cathode exhaust	Gases from expansion turbine	Gases from splitter to heat exchanger 1	Gases from heat exchanger 1	Gases from splitter to heat exchanger 4
Gases (lb·mol/h)						
CO	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	5,720.33	461.70	461.70	75.84	75.84	385.85
H_2	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	1,717.29	1,717.29	1,717.29	282.10	282.10	1,435.19
N_2	44,074.89	44,074.89	44,074.89	7,240.25	7,240.25	36,834.64
Ar	603.17	603.17	603.17	99.08	99.08	504.08
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_2S	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	10,162.86	7,533.54	7,533.54	1,237.55	1,237.55	6,295.99
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	62,278.54	54,390.59	54,390.59	8,934.82	8,934.82	45,455.75
Liquids (lb·mol/h)						
H_2O	0.00	0.00	0.00	0.00	0.00	0.00
Temperature (°F)	980.33	1,300.00	667.24	667.24	200.00	667.24
Pressure (psia)	150.00	150.00	14.70	14.70	14.70	14.70
Enthalpy of stream (Btu/h) (reference, 32°F)	482,489,027	546,065,398	281,008,417	46,161,714	15,923,634	234,846,704

TABLE 7.2 (Cont.)

Stream Data	Stream 24	Stream 25	Stream 26	Stream 27	Stream 28	Stream 29
Description of stream	Gases from heat exchanger 4	Water from condenser	Water from pump	Steam from heat exchanger 5	Steam from heat exchanger 4	Steam from heat exchanger 3
Gases (lb·mol/h)						
CO	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	385.85	0.00	0.00	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	1,435.19	0.00	0.00	10,270.88	15,985.92	28,434.17
N_2	36,834.64	0.00	0.00	0.00	0.00	0.00
Ar	504.08	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_2S	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	6,295.99	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	45,455.75	0.00	0.00	10,270.88	15,985.92	28,434.17
Liquids (lb·mol/h)						
H_2O	0.00	22,923.59	34,923.59	24,652.71	18,937.68	6,489.42
Temperature (°F)	400.00	121.36	121.36	356.77	356.79	356.77
Pressure (psia)	14.70	1.76	146.96	146.96	146.96	146.96
Enthalpy of stream (Btu/h) (reference, 32°F)	145,783,678	36,915,818	56,173,167	366,812,595	455,875,619	649,869,098

TABLE 7.2 (Cont.)

Stream Data	Stream 30	Stream 31	Stream 32	Stream 33	Stream 34A	Stream 34B	Stream 34C
Description of stream	Steam from heat exchanger 2	Steam for heating feed to fuel cell	Steam to steam turbine	Steam turbine exhaust	Makeup water to pump	Makeup water to pump	Makeup water to pump
Gases (lb·mol/h)							
CO	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	34,923.59	12,000.00	22,923.59	21,596.52	0.00	0.00	0.00
N_2	0.00	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	0.00
CH_4	0.00	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	0.00
H_2S	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	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	0.00
COS	0.00	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	0.00
Total gas flow	34,923.59	12,000.00	22,923.59	21,596.52	0.00	0.00	0.00
Liquids (lb·mol/h)							
H_2O	0.00	0.00	0.00	1,327.07	13,443.34	12,000.00	1,443.34
Temperature (°F)	775.00	775.00	775.00	121.36	150.00	150.00	150.00
Pressure (psia)	146.96	146.96	146.96	1.76	146.96	146.96	146.96
Enthalpy of stream (Btu/h) (reference, 32°F)	889,843,009	305,756,514	584,086,495	435,269,723	28,553,650	25,488,000	3,065,650

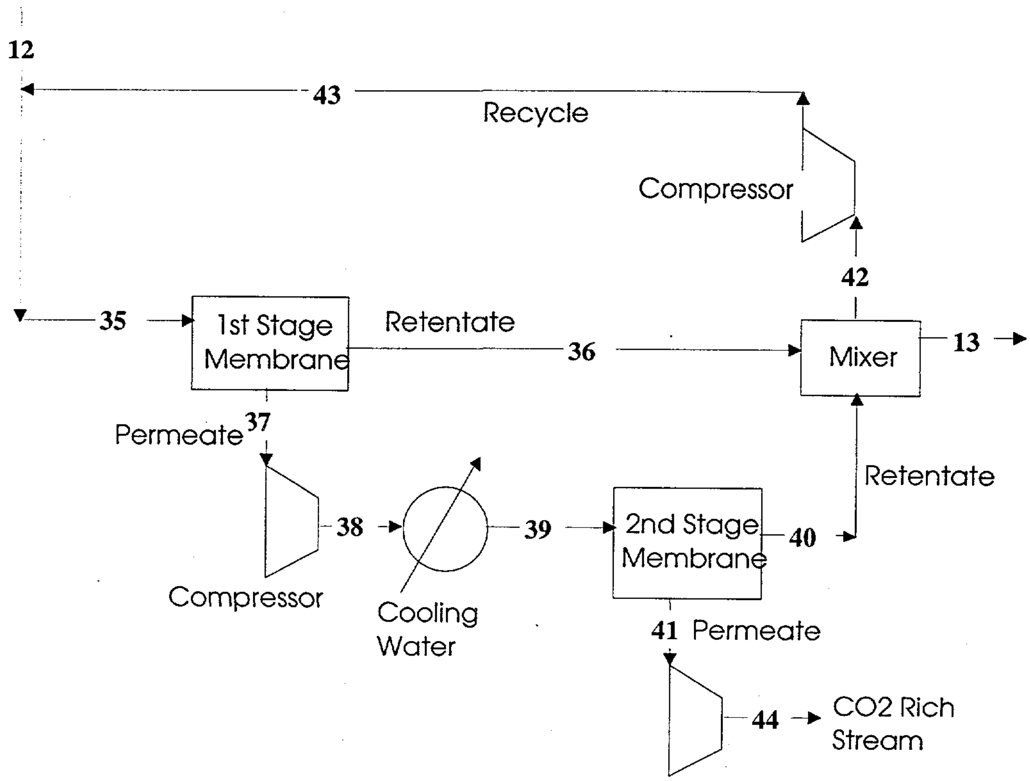

FIGURE 7.2 Flow Diagram of Membrane Process for CO_2 Removal in Case 4

TABLE 7.3 Stream Flows of Membrane Process for CO_2 Removal in Case 4

Stream Data	Stream 12	Stream 13	Stream 35	Stream 36	Stream 37	Stream 38
Description of stream	Feed gas from fuel cell system	H_2-rich gas to fuel cell	To 1st-stage membrane	1st-stage retentate	1st-stage permeate	Gases from compressor
Gases (lb·mol/h)						
CO	1,812.26	1,539.74	2,042.34	1,296.30	746.04	746.04
CO_2	8,724.66	4,180.59	9,349.34	2,831.36	6,517.98	6,517.98
H_2	1,693.50	1,692.69	1,946.43	1,906.70	39.72	39.72
H_2O	135.79	24.61	139.47	14.94	124.53	124.53
N_2	36.44	35.82	41.79	36.72	5.07	5.07
Ar	72.73	62.12	82.01	52.51	29.50	29.50
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_2S	0.00	0.00	0.00	0.00	0.00	0.00
HCN	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	12,475.38	7,535.57	13,601.38	6,138.53	7,462.84	7,462.84
Liquids (lb·mol/h)	0.00	0.00	0.00	0.00	0.00	0.00
Temperature (°F)	150.00	154.88	128.70	128.70	128.70	472.66
Pressure (psia)	150.00	140.00	150.00	140.00	25.00	150.00
Enthalpy of stream (Btu/h) (reference, 32°F)	12,462,913	7,542,574	11,053,736	4,694,179	6,359,557	31,178,763

TABLE 7.3 (Cont.)

Stream Data	Stream 39	Stream 40	Stream 41	Stream 42	Stream 43
Description of stream	Gases to 2nd-stage membrane	2nd-stage retentate	CO_2-rich product gas	Recycle gases to compressor	Recycle gases from compressor
Gases (lb·mol/h)					
CO	746.04	473.52	272.52	230.08	230.08
CO_2	6,517.98	1,973.91	4,544.07	624.69	624.69
H_2	39.72	38.91	0.81	252.93	252.93
H_2O	124.53	13.34	111.18	3.68	3.68
N_2	5.07	4.46	0.62	5.35	5.35
Ar	29.50	18.89	10.61	9.28	9.28
CH_4	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00
H_2S	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00
Total gas flow	7,462.84	2,523.03	4,939.81	1,126.01	1,126.01
Liquids (lb·mol/h)	0.00	0.00	0.00	0.00	0.00
Temperature (°F)	212.00	212.00	212.00	154.88	167.34
Pressure (psia)	150.00	140.00	25.00	140.00	150.00
Enthalpy of stream (Btu/h) (reference, 32°F)	12,062,267	3,975,443	8,086,823	1,127,051	1,243,809

TABLE 7.4 Descriptions of Streams of Membrane Process for CO_2 Removal in Case 4

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 12: CO_2-rich gas from fuel cell section		
Temperature (°F)	150	The synthesis gas is cleaned in two stages. First, sulfur compounds are removed by chilled methanol. This is the sulfur-free system.
Pressure (psia)	150	
Flow rate (lb·mol/h)	12,475.38	
CO_2 (mole fraction)	0.6994	
Stream 35: Feed gas to 1st-stage membrane system		
Temperature (°F)	128.70	The sulfur-free gas is mixed with the recycle from the 2nd-stage retentate and fed to the 1st-stage membranes.
Pressure (psia)	150	
Flow rate (lb·mol/h)	13,601.38	
CO_2 (mole fraction)	0.6874	
Stream 36: Retentate from 1st-stage membrane system		
Temperature (°F)	128.70	The composition of this stream depends on the permeability and selectivity of the membranes. The membrane system is a facilitated membrane that has a higher selectivity and permeability for CO_2 than for H_2. The ratio of permeate to retentate CO_2 selectivity is 2.3 times for a pressure drop of 125 psia.
Pressure (psia)	140	
Flow rate (lb·mol/h)	6,138.53	
CO_2 (mole fraction)	0.4612	
Stream 37: Permeate from 1st-stage membrane system		
Temperature (°F)	128.70	The composition of this stream is calculated by mass balance around the membrane.
Pressure (psia)	25	
Flow rate (lb·mol/h)	7,462.84	
CO_2 (mole fraction)	0.8734	
Stream 38: Gases from compressor		
Temperature (°F)	472.66	The permeate from the 1st-stage membrane is at a pressure of 25 psia. These gases are again compressed to a pressure of 150 psia for the 2nd-stage membrane.
Pressure (psia)	150	
Flow rate (lb·mol/h)	7,462.84	
CO_2 (mole fraction)	0.8734	

TABLE 7.4 (Cont.)

Stream and Characteristics	Data	Comments on Stream Calculations
Stream 39: Gases from heat exchanger		
Temperature (°F)	212	The temperature of the gases rises because of the compression. Therefore, this stream is cooled to a temperature of 212°F, suitable for the membrane system.
Pressure (psia)	150	
Flow rate (lb·mol/h)	7,462.84	
CO_2 (mole fraction)	0.8734	
Stream 40: Retentate of 2nd-stage membrane system		
Temperature (°F)	212	The composition of this stream is calculated on the basis of the selectivity and permeability of gases, as is done for stream 36. The ratio of permeate to retentate CO_2 selectivity is 2.3 for a pressure drop of 125 psia.
Pressure (psia)	140	
Flow rate (lb·mol/h)	2,523.03	
CO_2 (mole fraction)	0.7824	
Stream 41: Permeate of 2nd-stage membrane system		
Temperature (°F)	212	The composition of this stream is calculated on the basis of the mass balance around the membrane. This is the CO_2-rich stream for disposal.
Pressure (psia)	25	
Flow rate (lb·mol/h)	4,939.81	
CO_2 (mole fraction)	0.9199	
Stream 13: Fuel gas to gas turbines		
Temperature (°F)	154.88	H_2-rich retentate from 1st stage (stream 36) and that from 2nd stage (stream 40) are mixed. Part of mixture is taken as fuel gas for gas turbines.
Pressure (psia)	140	
Flow rate (lb·mol/h)	7,535.57	
CO_2 (mole fraction)	0.5548	
Stream 42: Recycle to 1st-stage membrane system		
Temperature (°F)	154.88	Part of the retentate from stream 36 and part from stream 40 are recycled back to the 1st-stage membrane systems to increase the CO_2 removal efficiency.
Pressure (psia)	140	
Flow rate (lb·mol/h)	1,126.01	
CO_2 (mole fraction)	0.5548	
Stream 43: Recycle to 1st-stage membrane after compression		
Temperature (°F)	167.34	The recycle from the retentate is at a pressure of 150 psia and is compressed to the inlet pressure of the 1st membrane.
Pressure (psia)	150	
Flow rate (lb·mol/h)	1,126.01	
CO_2 (mole fraction)	0.5548	

TABLE 7.5 Power Output, Plant Power Use, and Net Power Output for Base Case and Case 4 Fuel Cell/ Membrane Process

	Power (MW)	
Power Variable	Base Case	Fuel Cell Case
Power output		
Gas turbine or fuel cell	298.8	247.4
Steam turbine	159.4	165.8
Internal power consumption		
CO_2 recovery		
CO_2 compression	0	28.7
Solvent circulation	0	0
Solvent refrigeration	0	0
Others	0	-21.8
Gasification system[a]	-44.7	-48.9
Net power output	413.5	313.8
Energy penalty	0	99.7

[a] Includes H_2S recovery system energy use.

TABLE 7.6 Sizing and Cost Estimation for Major Equipment Used for Fuel Cell System in Case 4

1.	**Fuel Gas Expansion Turbine**		
	Turbine size (hp)	2,296	
	Purchased cost in 1979	\$1,607,439	
	Module factor	1.00	
	CE index for process equipment in 1979	\$256	
	CE index for process equipment in 1995	373.9	
	Installed cost of turbine in 1995		**\$2,347,740**
2.	**Heat Exchanger 1**		
	Q = Load (Btu/h)	30,238,080	
	Tha = Inlet temperature of hot fluid (°F)	667.24	
	Thb = Outlet temperature of hot fluid (°F)	200	
	Pressure of hot gases (psia)	15	
	Tca = Inlet temperature of cold fluid (°F)	154.9	
	Tcb = Outlet temperature of cold fluid (°F)	600.00	
	Delta T1	67.2395	
	Delta T2	45	
	Log mean temperature difference (°F)	55	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	5	
	Heat transfer area (ft^2)	109,077	
	Operating pressure (psia)	150.00	
	Pressure factor	1.16	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction, shell and tube floating head)	\$545,385	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**\$2,365,464**

TABLE 7.6 (Cont.)

3.	**Heat Exchanger 2**		
	Q = Load (Btu/h)	239,973,908	
	Tha = Inlet temperature of hot fluid (°F)	1300.00	
	Thb = Outlet temperature of hot fluid (°F)	450	
	Pressure of hot gases (psia)	150	
	Tca = Inlet temperature of cold fluid (°F)	356.8	
	Tcb = Outlet temperature of cold fluid (°F)	775.00	
	Delta T1	525	
	Delta T2	93	
	Log mean temperature difference (°F)	250	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	30	
	Heat transfer area (ft^2)	32,019	
	Operating pressure (psia)	146.96	
	Pressure factor	1.165	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction, shell and tube floating head)	$250,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$1,088,984**
4.	**Heat Exchanger 3**		
	Q = Load (Btu/h)	193,993,479	
	Tha = Inlet temperature of hot fluid (°F)	1355.19	
	Thb = Outlet temperature of hot fluid (°F)	980	
	Pressure of hot gases (psia)	150	
	Tca = Inlet temperature of cold fluid (°F)	356.8	
	Tcb = Outlet temperature of cold fluid (°F)	356.77	
	Delta T1	998.4229363	
	Delta T2	624	
	Log mean temperature difference (°F)	796	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	30	
	Heat transfer area (ft^2)	8,120	
	Operating pressure (psia)	146.96	
	Pressure factor	1.165	
	Materials correction factor		
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction, shell and tube floating head)	$95,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$413,814**

TABLE 7.6 (Cont.)

5.	**Heat Exchanger 4**		
	Q = Load (Btu/h)	89,063,026	
	Tha = Inlet temperature of hot fluid (°F)	667.24	
	Thb = Outlet temperature of hot fluid (°F)	400	
	Pressure of hot gases (psia)	15	
	Tca = Inlet temperature of cold fluid (°F)	356.8	
	Tcb = Outlet temperature of cold fluid (°F)	356.79	
	Delta T1	310.4528363	
	Delta T2	43	
	Log mean temperature difference (°F)	136	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	30	
	Heat transfer area (ft^2)	21,903	
	Operating pressure (psia)	146.96	
	Pressure factor	1.165	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction, shell and tube floating head)	$180,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$784,068**
6.	**Heat Exchanger 5**		
	Q = Load (Btu/h)	310,639,429	
	Tha = Inlet temperature of hot fluid (°F)	450.00	
	Thb = Outlet temperature of hot fluid (°F)	150	
	Pressure of hot gases (psia)	150	
	Tca = Inlet temperature of cold fluid (°F)	121.4	
	Tcb = Outlet temperature of cold fluid (°F)	356.77	
	Delta T1	93.23133627	
	Delta T2	29	
	Log mean temperature difference (°F)	55	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	30	
	Heat transfer area (ft^2)	189,208	
	Operating pressure (psia)	146.96	
	Pressure factor	1.165	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction, shell and tube floating head)	$946,038	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$4,120,882**

TABLE 7.6 (Cont.)

7.	**Cathode Gas Expansion Turbine**		
	Turbine size (hp)	104,190	
	Purchased cost in 1987	$10,432,285	
	(assumes that the cost of expansion turbine is same as that of a compressor of similar size)		
	Module factor	1.00	
	CE index for process equipment in 1987	$320	
	CE index for process equipment in 1995	$374	
	Installed cost of turbine in 1995		**$12,189,473**
8.	**Air Compressor for Fuel Cell**		
	Inlet pressure (psia)	14.70	
	Outlet pressure (psia)	$150	
	Compressor size (MW)	224.43	
	Purchased cost in 1987	$24,374,545	
	Module factor	1.00	
	CE index for process equipment in 1987	320	
	CE index for process equipment in	1995373.9	
	Installed cost of air compressor in 1995		**$28,480,133**
9.	**Steam Turbine**		
	Turbine output (MW)	165.77	
	The cost of steam turbine is already included in base case.		
10.	**Condenser**		
	Q = Load (Btu/h)	398,353,905	
	Tha = Inlet temperature of hot fluid (°F)	121.36	
	Thb = Outlet temperature of hot fluid (°F)	121	
	Pressure of hot gases (psia)	2	
	Tca = Inlet temperature of cold fluid (°F)	70.0	
	Tcb = Outlet temperature of cold fluid (°F)	100.00	
	Delta T1	21.35924367	
	Delta T2	51	
	Log mean temperature difference (°F)	34	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	500	
	Heat transfer area (ft^2)	23,300	
	Operating pressure (psia)	146.96	
	Pressure factor	1.165	
	Materials correction factor	1	
	Module factor	3.2	
	(includes all of the supporting equipment and connections and installation)		
	Purchased cost of heat exchanger in 1987	$190,000	
	(mild steel construction, shell and tube floating head)		
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$827,628**

TABLE 7.6 (Cont.)

11.	**Pump**		
	Horsepower	106	
	Size exponent	1	
	Purchased cost in 1987	$12,000	
	(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 1995	373.9	
	Installed cost of pump in 1995		**$21,035**
12.	**Fuel Cell Stack**		
	Fuel cell power output (kW)	77,989	
	Unit cost per kilowatt	$180	
	Total cost		**$14,038,020**
13.	**Fuel Cell Invertor**		
	Unit cost per kilowatt	$100	
	Total cost		**$7,798,900**
14.	**Fuel Cell Controls**		
	Unit cost per kilowatt	$140	
	Total cost		**$10,918,460**
15.	**Fuel Cell and Component Assembly**		
	Unit cost per kilowatt	$110	
	Total cost		**$8,578,790**
	Total Direct Cost		**$93,973,387**
	Total Direct Cost for Three Trains		**$281,920,162**

TABLE 7.7 Sizing and Cost Estimation for Major Equipment Used for CO_2 Removal in Membrane Process in Case 4

1.	**First-Stage Membranes**		
	Membrane area (ft^2)	2,346,506	
	Unit cost of membrane	$13.00	
	Total cost		**$30,504,579**
2.	**Second-Stage Membranes**		
	Membrane area (ft^2)	1,287,497	
	Unit cost of membrane	$13.00	
	Total cost		**$16,737,465**
3.	**Compressor between First and Second Stages**		
	Inlet pressure (psia)	25.00	
	Outlet pressure (psia)	150.00	
	Compressor size (hp)	9,747	
	Purchased cost of reciprocating compressor in 1987 (includes electric motor drive and gear reducer)	$1,600,000	
	Size factor for compressor	1	
	Materials correction factor	1	
	Module factor	2.6	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of compressor in 1995		**$4,860,700**
4.	**Recycle Compressor**		
	Inlet pressure (psia)	140.00	
	Outlet pressure (psia)	150.00	
	Compressor size (hp)	46	
	Purchased cost of reciprocating compressor in 1987 (includes electric motor drive and gear reducer)	$38,000	
	Size factor for compressor	1	
	Materials correction factor	1	
	Module factor	2.6	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of compressor in 1995		**$115,442**

TABLE 7.7 (Cont.)

5.	**Heat Exchanger after Compressor**		
	Q = Load (Btu/h)	19,116,496	
	Tha = Inlet temperature of hot fluid (°F)	472.66	
	Thb = Outlet temperature of hot fluid (°F)	212	
	Pressure of hot gases (psia)	150	
	Tca = Inlet temperature of cold fluid (°F)	70.00	
	Tcb = Outlet temperature of cold fluid (°F)	150.00	
	Delta T1	322.66	
	Delta T2	142	
	Log mean temperature difference (°F)	220	
	Overall heat transfer coefficient (Btu/h/ft^2/°F)	30	
	Heat transfer area (ft^2)	2,895	
	Operating pressure (psia)	445	
	Pressure factor	1.08	
	Materials correction factor	1	
	Module factor (includes all of the supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$50,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$201,906**
6.	**CO_2 Product Gas Compressors**		
	Compressor 1 (hp)	4,276	
	Compressor 2 (hp)	4,276	
	Compressor 3 (hp)	4,276	
	Purchased cost of centrifugal compressor 1 in 1987	$900,000	
	Purchased cost of centrifugal compressor 2 in 1987	$900,000	
	Purchased cost of centrifugal compressor 3 in 1987 (includes electric motor drive and gear reducer)	$900,000	
	Size factor for compressor	1	
	Module factor	2.6	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of Compressor 1 in 1995		**$2,734,144**
	Installed cost of Compressor 2 in 1995		**$2,734,144**
	Installed cost of Compressor 3 in 1995		**$2,734,144**
	Total Direct Cost		**$162,204,286**
	Total Direct Cost for Three Trains		**$486,612,859**

8 CO_2 Pipeline Transport and Sequestering

8.1 Pipeline Transport of CO_2

Once the CO_2 has been recovered from the fuel-gas stream, its transportation, utilization, and disposal remain significant issues. In a previous study for METC (Doctor et al. 1994), the issues associated with the transport and sequestering of CO_2 were considered in greater detail; that information serves as the basis for this work. The CO_2 represents a large-volume, relatively low-value by-product that cannot be sequestered in the same way as most coal-utilization wastes (i.e., by landfilling). Large volumes of recovered CO_2 are likely to be moved by pipeline, and if sequestering were required, new pipelines would likely need to be constructed. In some cases, existing pipelines could be used, perhaps in a shared mode with other products. Costs for pipeline construction and use vary greatly on a regional basis within the United States. The recovered CO_2 represents more than 3 million normal cubic meters per day of gas volume. It is assumed that the transport and sequestering process releases approximately 2% of the recovered CO_2.

8.2 CO_2 Sequestering

Proposals have been made to dispose of CO_2 in the ocean depths. However, many questions of engineering and ecological concern associated with such options remain unanswered, and the earliest likely reservoir is a land-based geological repository (Hangebrauck 1992). A portion of the CO_2 can be used for enhanced oil recovery, which sequesters a portion of the CO_2, or the CO_2 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; they take into account that the power required for compression will rise throughout the life cycle of these sequestering reservoirs. The first reservoirs to be used will, in fact, be capable of accepting all IGCC CO_2 gas for a 30-year period without requiring any additional compression costs for operation. The pipeline transport and sequestering process represents approximately 26 mills/kWh for the CO_2-recovery cases.

9 Conclusions — Energy Cycle/Economic Comparisons

9.1 Energy Consumption and CO_2 Emissions

An adjustment of 9.7% between the oxygen-blown and air-blown KRW IGCC cases was needed to make the coal feed rates match. A second minor adjustment was required because the design basis coal was different for these two sets of studies. Efficiencies calculated previously were matched, while the CO_2 emission rates for the air-blown cases decreased slightly by 4.2%.

Data on energy consumption and CO_2 emissions for all seven cases appear in Tables 9.1-9.7. The IGCC power plant performance and emission factors within traditional battery limits have been bounded to clarify what in the net energy cycle falls outside the plant battery. The most significant contributor to the net CO_2 emissions for the CO_2-recovery cases is the makeup power to match the base case performance.

9.2 Capital Costs for KRW Integrated Gasification Combined-Cycle Power Generation

Capital costs for each of the IGCC power plants appear in Tables 9.8-9.13. For convenience in comparison, the O_2-blown and air-blown cases are next to each other. The large cost difference between these two systems for the coal preparation system is a consequence of the fact that the air-blown system employs the sulfator section off-gases for coal drying. The O_2-blown case requires an air-separation system and compression. Here the air-blown case is lower in cost as a consequence of needing only compression. From this section of the plant forward, the O_2-blown case shows lower costs for comparable plant subsystems as a consequence of the reduced gas volumes being handled.

Whenever a standard turn-key package system was part of the design, a zero percent contingency was taken. In addition, throughout the study, the Handy-Whitman Index was employed to bring all capital estimates to a fourth quarter of 1994 dollar basis. The plant cost for the O_2-blown base case comes to $1,332/kW; for the air-blown case, it is slightly lower, at $1,253/kW. For the optimal O_2-blown CO_2-recovery case, this cost increases to $1,687/kW, while for the optimal air-blown CO_2-recovery case, this cost rises to $1,773/kW.

9.3 Costs of Electricity

The costs of electricity appear in Tables 9.14-9.20. Following this, Table 9.21 summarizes the major costs for each of the combined-cycle cases. For the air-blown cases, the cost of limestone and the cost of ash disposal have been adjusted to typical values as given by the TAG study (EPRI 1993). A comparison of the cost of electricity for the CO_2-release base cases found the cost of the air-blown IGCC case to be 58.29 mills/kWh and the cost of the O_2-blown IGCC case to be 56.86 mills/kWh. There was no clear advantage for the optimal cases employing glycol CO_2 recovery; the cost of the air-blown IGCC was 95.48 mills/kWh, and the cost of the O_2-blown case was slightly lower, at 94.55 mills/kWh.

TABLE 9.1 Energy Consumption and CO_2 Emissions for Oxygen-Blown Base Case: KRW IGCC with No CO_2 Recovery

	Electricity MW	CO2 release kg/h
Mining and Transport		
Raw Coal in Mine	-2.36	2,356
Coal Rail Transport	-0.05	523
Subtotal	-2.41	2,879
IGCC Power Plant		
Coal Preparation	-0.85	0
Gasifier Island	-36.82	6,153
Power Island	-7.02	320,387
Subtotal	-44.70	326,540
Power - Gas Turbine	298.80	
Power - Steam Turbine	159.40	
GROSS Power	**458.20**	
NET Power	**413.50**	
Pipeline/Sequester	0.00	0
Energy Cycle Power Use	**-47.11**	
NET Energy Cycle	**411.09**	**329,419**
CO2 emission rate/net cycle	**0.801**	**kg CO2/kWh**
Power use/CO2 in reservoir	**N/A**	**kWh/kg CO2**

TABLE 9.2 Energy Consumption and CO_2 Emissions for Air-Blown Base Case: KRW IGCC with No CO_2 Recovery

	Electricity	CO2 release
Mining and Transport	MW	kg/h
Raw Coal in Mine	-2.36	2,356
Coal Rail Transport	-0.05	523
Limestone Mining	-0.25	250
Limestone Rail Tansport	-0.02	156
Subtotal	-2.67	3,286
IGCC Power Plant		
Coal/Limestone Preparation	-3.49	11,374
Gasifier Island	-20.12	137
Power Island	-10.58	315,029
Subtotal	-34.19	326,540
Power - Gas Turbine	302.66	
Power - Steam Turbine	176.97	
GROSS Power	**479.63**	
NET Power	**445.44**	
Pipeline/Sequester	0.00	0
Energy Cycle Power Use	-36.87	
NET Energy Cycle	442.76	329,825
CO2 emission rate/net cycle	**0.745**	**kg CO2/kWh**
Power use/CO2 in reservoir	N/A	**kWh/kg CO2**

TABLE 9.3 Energy Consumption and CO_2 Emissions for Case 1: Oxygen-Blown KRW IGCC with Glycol CO_2 and H_2S Recovery and Gas Turbine Topping Cycle

	Electricity MW	CO2 release kg/h
Mining and Transport		
Raw Coal in Mine	-2.36	2,356
Coal Rail Transport	-0.05	523
Subtotal	-2.41	2,879
IGCC Power Plant		
Coal Preparation	-0.85	0
Gasifier Island	-36.82	6,153
Power Island	-7.02	320,387
Glycol Circulation	-5.80	-260,055
Glycol Refrigeration	-4.50	
Power Recovery Turbines	3.40	
CO2 Compression (to 2100psi)	-17.30	
Subtotal	-68.90	66,485
Power - Gas Turbine	284.80	
Power - Steam Turbine	161.60	
GROSS Power	**446.40**	
NET Power	**377.50**	
Pipeline/Sequester		
Pipeline CO2		260,055
Pipeline booster stations	-1.64	1,637
Geological reservoir (2% loss)	0.00	-254,854
Subtotal	-1.64	6,839
Energy Cycle Power Use	-72.95	
NET Energy Cycle	**373.45**	**76,202**
Derating from O2-Base Case	**37.64**	
Make-up Power	37.64	37,637
TOTAL	**411.09**	**113,840**
CO2 emission rate/net cycle	**0.277**	**kg CO2/kWh**
Power use/CO2 in reservoir	**0.148**	**kWh/kg CO2**

TABLE 9.4 Energy Consumption and CO_2 Emissions for Case 2: Oxygen-Blown KRW IGCC with Membrane CO_2 Recovery, Glycol H_2S Recovery, and Gas Turbine Topping Cycle

	Electricity MW	CO2 release kg/h
Mining and Transport		
Raw Coal in Mine	-2.36	2,356
Coal Rail Transport	-0.05	523
Subtotal	-2.41	2,879
IGCC Power Plant		
Coal Preparation	-0.85	
Gasifier Island	-36.82	6,153
Power Island	-7.02	320,387
Glycol Circulation	-0.90	-232,505
Glycol Refrigeration	-3.00	
Membrane Compression	-19.00	
CO2 Compression (to 2100psi)	-20.00	
Subtotal	-87.60	94,034
Power - Gas Turbine	262.80	
Power - Steam Turbine	154.80	
GROSS Power	**417.60**	
NET Power	**330.00**	
Pipeline/Sequester		
Pipeline CO2		232,505
Pipeline booster stations	-1.46	1,464
Geological reservoir (2% loss)	0.00	-227,855
Subtotal	-1.46	6,114
Energy Cycle Power Use	-91.47	
NET Energy Cycle	**326.13**	**103,028**
Derating from O2-Base Case	**84.96**	
Make-up Power	84.96	84,964
TOTAL	**411.09**	**187,992**
CO2 emission rate/net cycle	**0.457 kg CO2/kWh**	
Power use/CO2 in reservoir	**0.373 kWh/kg CO2**	

TABLE 9.5 Energy Consumption and CO_2 Emissions for Case 3: Oxygen-Blown KRW IGCC with Glycol CO_2 Recovery, Methanol H_2S Recovery, and Fuel Cell Topping Cycle

	Electricity MW	CO2 release kg/h
Mining and Transport		
Raw Coal in Mine	-2.36	2,356
Coal Rail Transport	-0.05	523
Subtotal	-2.41	2,879
IGCC Power Plant		
Coal Preparation	-0.85	
Gasifier Island	-36.82	6,153
Power Island	-11.24	320,387
CO2 Recovery	-4.54	-260,055
CO2 Compression (to 2100psi)	-24.93	
Subtotal	-78.39	66,485
Power - Gas Turbine	246.70	
Power - Steam Turbine	171.80	
GROSS Power	**418.50**	
NET Power	**340.11**	
Pipeline/Sequester		
Pipeline CO2		260,055
Pipeline booster stations	-1.64	1,637
Geological reservoir (2% loss)	0.00	-254,854
Subtotal	-1.64	6,839
Energy Cycle Power Use	-82.44	
NET Energy Cycle	**336.06**	**76,202**
Derating from O2-blown Base Case	75.03	
Make-up Power	75.03	75,030
TOTAL	411.09	151,232
CO2 emission rate/net cycle	**0.368**	**kg CO2/kWh**
CO2 Sequestering power use	**75.03**	**MW**
Power use/CO2 in reservoir	**0.294**	**kWh/kg CO2**

TABLE 9.6 Energy Consumption and CO_2 Emissions for Case 4: Oxygen-Blown KRW IGCC with Membrane CO_2 Recovery, Methanol H_2S Recovery, and Fuel Cell Topping Cycle

	Electricity MW	CO2 release kg/h
Mining and Transport		
Raw Coal in Mine	-2.36	2,356
Coal Rail Transport	-0.05	523
Subtotal	-2.41	2,879
IGCC Power Plant		
Coal Preparation	-0.85	
Gasifier Island	-36.82	6,153
Power Island	-11.22	320,387
CO2 Recovery	-21.80	-272,137
CO2 Compression (to 2100psi)	-28.70	
Subtotal	-99.40	54,403
Power - Fuel Cells	247.40	
Power - Steam Turbine	165.80	
GROSS Power	**413.20**	
NET Power	**313.80**	
Pipeline/Sequester		
Pipeline CO2		272,137
Pipeline booster stations	-1.71	1,713
Geological reservoir (2% loss)	0.00	-266,694
Subtotal	-1.71	7,156
Energy Cycle Power Use	-103.52	
NET Energy Cycle	**309.68**	**64,438**
Derating from O2-blown Base Case	101.41	
Make-up Power	101.41	101,413
TOTAL	**411.09**	**165,852**
CO2 emission rate/net cycle	**0.403**	**kg CO2/kWh**
CO2 Sequestering power use	**101.41**	**MW**
Power use/CO2 in reservoir	**0.380**	**kWh/kg CO2**

TABLE 9.7 Energy Consumption and CO_2 Emissions for Optimal Air-Blown Case: KRW IGCC with Glycol CO_2 Recovery, In-Bed H_2S Recovery, and Gas Turbine Topping Cycle

	Electricity MW	CO2 release kg/h
Mining and Transport		
Raw Coal in Mine	-2.36	2,356
Coal Rail Transport	-0.05	523
Limestone Mining	-0.25	250
Limestone Rail Tansport	-0.02	156
Subtotal	-2.67	3,286
IGCC Power Plant		
Coal/Limestone Preparation	-3.49	11,374
Gasifier Island	-21.11	137
Power Island	-11.10	315,029
CO2 Recovery	-17.21	-285,499
CO2 Compression (to 2100psi)	-32.21	
Subtotal	-85.11	41,041
Power - Gas Turbine	274.39	
Power - Steam Turbine	186.50	
GROSS Power	**460.88**	
NET Power	**375.77**	
Pipeline/Sequester		
Pipeline CO2		285,499
Pipeline booster stations	-1.80	1,798
Geological reservoir (2% loss)	0.00	-279,789
Subtotal	-1.80	7,508
Energy Cycle Power Use	-89.58	
NET Energy Cycle	**371.30**	**51,834**
Derating from O2-blown Base Case	39.79	
Make-up Power	39.79	39,794
TOTAL	**411.09**	**91,628**

CO2 emission rate/net cycle	**0.223**	**kg CO2/kWh**
CO2 Sequestering power use	**39.79**	**MW**
Power use/CO2 in reservoir	**0.142**	**kWh/kg CO2**

TABLE 9.8 Capital Costs for Air-Blown and Oxygen-Blown Base Cases with No CO_2 Recovery

	KRW O2-Blown Base Case 413.50 MW		KRW Air-Blown Base Case 445.44 MW	
System	cont.*	**Capital Cost, $K**	cont.*	**Capital Cost, $K**
Direct Costs				
Coal Handling & Preparation	0.0%	$8,339	0.0%	$18,208
Limestone Handling & Prep.			0.0%	$10,388
Air-Separation Plant/Comprs.	0.0%	$66,249	0.0%	$10,099
Gasification	20.0%	$99,714	20.0%	$118,866
Fines and Ash Handling	15.0%	$2,650	15.0%	$6,628
Acid Gas Treatment (H2S)	10.0%	$12,286	10.0%	$37,902
Sulfur Recovery (Claus)	0.0%	$6,777		
Tail-Gas Treatment (SCOT)	0.0%	$6,116		
Sour-water Stripping	10.0%	$4,408		
Wastewater Treatment	30.0%	$5,116		
Gas Turbine System	5.0%	$77,837	5.0%	$80,654
HRSG System	5.0%	$25,808	5.0%	$28,407
Steam Turbine System	0.0%	$47,900	0.0%	$52,722
Sub-total		**$363,199**		**$363,873**
Indirect Costs				
General Facilities	10.5%	$38,136	10.5%	$38,207
Engineering Fees	8.0%	$29,056	8.0%	$29,110
Process Contingency	7.9%	$28,727	9.3%	$34,011
Project Contingency	20.0%	$91,823	20.0%	$93,040
Sub-total		**$187,742**		**$194,367**
Total Plant Cost-TPC		**$550,941**		**$558,241**
Cost($/kW-net output)		**$1,332**		**$1,253**
Interest & Inflation (AFUDC)**	20.5%	$112,943	20.5%	$114,439
Total Plant Investment-TPI		**$663,884**		**$672,680**
Royalties	0.6%	$2,179	0.6%	$2,183
Initial Inventory	3.3%	$11,986	3.3%	$12,008
Start-up Costs	4.6%	$16,707	4.6%	$16,738
Spare Parts	2.2%	$7,990	2.2%	$8,005
Working Capital	3.3%	$11,986	3.3%	$12,008
TOTAL		**$714,731**		**$723,622**

TABLE 9.9 Capital Costs for Case 1: Oxygen-Blown KRW IGCC with Glycol CO_2 and H_2S Recovery and Gas Turbine Topping Cycle

		Net Power Case #1 377.5 MW
System	cont.*	**Capital Cost, $K**
Direct Costs		
Coal Handling & Preparation	0.0%	$8,339
Air-Separation Plant/Comprs.	0.0%	$66,249
Gasification	20.0%	$99,714
Fines and Ash Handling	15.0%	$2,650
Glycol (H2S)	10.0%	$17,756
Sulfur Recovery (Claus)	0.0%	$6,777
Tail-Gas Treatment (SCOT)	0.0%	$6,116
Sour-water Stripping	10.0%	$4,408
Shift System	10.0%	$21,571
Glycol (CO2 Recovery)	10.0%	$28,597
Wastewater Treatment	30.0%	$5,116
Gas Turbine System	5.0%	$77,837
HRSG System	5.0%	$25,808
Steam Turbine System	0.0%	$47,900
Sub-total		**$418,838**
Indirect Costs		
General Facilities	10.5%	$43,978
Engineering Fees	8.0%	$33,507
Process Contingency	8.2%	$34,291
Project Contingency	20.0%	$106,123
Sub-total		**$217,898**
Total Plant Cost-TPC		**$636,737**
Cost($/kW-net output)		**$1,687**
Interest & Inflation (AFUDC)**	20.5%	$130,531
Total Plant Investment-TPI		**$767,268**
Royalties	0.6%	$2,513
Initial Inventory	3.3%	$13,822
Start-up Costs	4.6%	$19,267
Spare Parts	2.2%	$9,214
Working Capital	3.3%	$13,822
TOTAL		**$825,905**

TABLE 9.10 Capital Costs for Case 2: Oxygen-Blown KRW IGCC with Membrane CO_2 Recovery, Glycol H_2S Recovery, and Gas Turbine Topping Cycle

	Net Power Case #2 330.0 MW	
System	**cont.***	**Capital Cost, $K**
Direct Costs		
Coal Handling & Preparation	0.0%	$8,339
Air-Separation Plant/Comprs.	0.0%	$66,249
Gasification	20.0%	$99,714
Fines and Ash Handling	15.0%	$2,650
Glycol (H2S)	10.0%	$17,756
Sulfur Recovery (Claus)	0.0%	$6,777
Tail-Gas Treatment (SCOT)	0.0%	$6,116
Sour-water Stripping	10.0%	$4,408
Shift System	10.0%	$19,980
Membrane (CO2 Recovery)	10.0%	$110,448
Wastewater Treatment	30.0%	$5,116
Gas Turbine System	5.0%	$77,837
HRSG System	5.0%	$25,808
Steam Turbine System	0.0%	$47,900
Sub-total		**$499,097**
Indirect Costs		
General Facilities	10.5%	$52,405
Engineering Fees	8.0%	$39,928
Process Contingency	8.5%	$42,316
Project Contingency	20.0%	$126,749
Sub-total		**$261,399**
Total Plant Cost-TPC		**$760,496**
Cost($/kW-net output)		**$2,305**
Interest & Inflation (AFUDC)**	20.5%	$155,902
Total Plant Investment-TPI		**$916,397**
Royalties	0.6%	$2,995
Initial Inventory	3.3%	$16,470
Start-up Costs	4.6%	$22,958
Spare Parts	2.2%	$10,980
Working Capital	3.3%	$16,470
TOTAL		$986,271

TABLE 9.11 Capital Costs for Case 3: Oxygen-Blown KRW IGCC with Glycol CO_2 Recovery, Methanol H_2S Recovery, and Fuel Cell Topping Cycle

	Net Power Case #3 340.11 MW	
System	**cont.***	**Capital Cost, $K**
Direct Costs		
Coal Handling & Preparation	0.0%	$8,339
Air-Separation Plant/Comprs.	0.0%	$66,249
Gasification	20.0%	$99,714
Fines and Ash Handling	15.0%	$2,650
Chilled Methanol (H2S)	10.0%	$22,825
Sulfur Recovery (Claus)	0.0%	$6,777
Tail-Gas Treatment (SCOT)	0.0%	$6,116
Sour-water Stripping	10.0%	$4,408
Glycol (CO2 Recovery)	10.0%	$31,555
Wastewater Treatment	30.0%	$5,116
Molten Carbonate Fuel Cells	15.0%	$285,637
Steam Turbine System	0.0%	$47,900
Sub-total		**$587,286**
Indirect Costs		
General Facilities	10.5%	$61,665
Engineering Fees	8.0%	$46,983
Process Contingency	12.0%	$70,599
Project Contingency	20.0%	$153,307
Sub-total		**$332,554**
Total Plant Cost-TPC		**$919,840**
Cost($/kW-net output)		**$2,705**
Interest & Inflation (AFUDC)**	20.5%	$188,567
Total Plant Investment-TPI		**$1,108,407**
Royalties	0.6%	$3,524
Initial Inventory	3.3%	$19,380
Start-up Costs	4.6%	$27,015
Spare Parts	2.2%	$12,920
Working Capital	3.3%	$19,380
TOTAL		$1,190,627

TABLE 9.12 Capital Costs for Case 4: Oxygen-Blown KRW IGCC with Membrane CO_2 Recovery, Methanol H_2S Recovery, and Fuel Cell Topping Cycle

	Net Power Case #4 313.77 MW	
System	cont.*	**Capital Cost, $K**
Direct Costs		
Coal Handling & Preparation	0.0%	$8,339
Air-Separation Plant/Comprs.	0.0%	$66,249
Gasification	20.0%	$99,714
Fines and Ash Handling	15.0%	$2,650
Chilled Methanol (H2S)	10.0%	$22,825
Sulfur Recovery (Claus)	0.0%	$6,777
Tail-Gas Treatment (SCOT)	0.0%	$6,116
Sour-water Stripping	10.0%	$4,408
CO2 Recovery - Membrane	10.0%	$181,868
Wastewater Treatment	30.0%	$5,116
Molten Carbonate Fuel Cells	15.0%	$281,920
Steam Turbine System	0.0%	$47,900
Sub-total		**$733,882**
Indirect Costs		
General Facilities	10.5%	$77,058
Engineering Fees	8.0%	$58,711
Process Contingency	11.6%	$85,073
Project Contingency	20.0%	$190,945
Sub-total		**$411,786**
Total Plant Cost-TPC		**$1,145,668**
Cost($/kW-net output)		**$3,651**
Interest & Inflation (AFUDC)**	20.5%	$234,862
Total Plant Investment-TPI		**$1,380,529**
Royalties	0.6%	$4,403
Initial Inventory	3.3%	$24,218
Start-up Costs	4.6%	$33,759
Spare Parts	2.2%	$16,145
Working Capital	3.3%	$24,218
TOTAL		$1,483,273

TABLE 9.13 Capital Costs for Optimal Air-Blown Case: KRW IGCC with Glycol CO_2 Recovery, In-Bed H_2S Recovery, and Gas Turbine Topping Cycle

	Net Power Glycol CO2 375.77 MW	
System	cont.*	**Capital Cost, $K**
Direct Costs		
Coal Handling & Preparation	0.0%	$18,208
Limestone Handling & Prep.	0.0%	$10,388
Air-Separation Plant/Comprs.	0.0%	$10,099
Gasification	20.0%	$118,866
Fines and Ash Handling	15.0%	$6,628
Glycol H2S	10.0%	$37,902
Shift/Glycol CO2/Compression	10.0%	$60,321
Gas Turbine System	5.0%	$80,654
HRSG System	5.0%	$28,407
Steam Turbine System	0.0%	$52,722
Sub-total		**$424,194**
Indirect Costs		
General Facilities	10.5%	$44,540
Engineering Fees	8.0%	$33,936
Process Contingency	9.4%	$40,043
Project Contingency	20.0%	$108,542
Sub-total		**$227,061**
Total Plant Cost-TPC		**$651,255**
Cost($/kW-net output)		**$1,733**
Interest & Inflation (AFUDC)**	20.5%	$133,507
Total Plant Investment-TPI		**$784,762**
Royalties	0.6%	$2,545
Initial Inventory	3.3%	$13,998
Start-up Costs	4.6%	$19,513
Spare Parts	2.2%	$9,332
Working Capital	3.3%	$13,998
TOTAL		**$844,149**

TABLE 9.14 Operating Costs for Oxygen-Blown Base Case: KRW IGCC with No CO_2 Recovery

Net Power (MW) =	**413.50**
Capacity factor=	**65%**
Annual Net Power Production (MW)=	**2,354,469**
Net Energy-cycle Power (MW)=	411.09

OPERATING COSTS	**Basis**	**Units**	**Unit Cost**		**Annual Cost**
Fuel - Illinois #6 Coal (ROM)	4,110	T/D	$35.00	S/T	**$34,126,136**
Coal - prepared	3,845	T/D			
Consumable material					
Catalyst, etc.					$1,640,000
Miscellaneous					$603,730
Ash/Sorbent Disposal	491.4	T/D	$11.00	S/T	$1,282,432
Plant Labor					
Oper Labor (w benefits)	23.0	men/shift	$25.50	S/h	$5,137,198
Supervision/support	25%	of above			$1,284,300
Maintenance	2.7%	of Direct			$9,806,370
Insurance & Local Taxes	0.9%	of Direct			$3,268,790
Other - % of Oper Labor	12.5%	of above			$642,150
By-Product Credit	102.1	TPD	$30.00	S/T	($726,857)
Net Operating Cost					**$22,938,113**

COSTS OF ELECTRICITY

Levelizing Factors	**Constant ($)**	**Basis (K$)**	**Annual (K$)**
Capital Charge	0.111	$714,731	$137,253
Fuel	1.025	$34,126	
Operating & Maintenance	1.000	$22,938	
Cost of Electricity - Levelized	**mills/kWh**		
Capital Charge	33.70		
Fuel	14.86		
Operating & Maintenance	9.74		
Total Cost of Electricity	**58.29**	Basis (MW) 413.5	
Energy-cycle Cost of Electricity	58.64	Basis (MW) 411.1	

TABLE 9.15 Operating Costs for Air-Blown Base Case: KRW IGCC with No CO_2 Recovery

		Net Power (MW) =	**445.44**
		Capacity factor=	**65%**
		Annual Net Power Production (MW)=	**2,536,335**
		Net Energy-cycle Power (MW)=	441.26

OPERATING COSTS	Basis	Units	Unit Cost		Annual Cost
Fuel - Illinois #6 Coal (ROM)	4109.7	T/D	$35.00	$/T	**$34,126,164**
Coal - prepared	3,845	T/D			
Consumable material					
Limestone	1100.8	T/D	$11.20	$/T	$2,925,032
Nahcolite	4.9	T/D	$261.25	$/T	$301,676
Zinc Ferrite	1.1	T/D	$6,270.00	$/T	$1,659,216
Miscellaneous					$603,730
Ash/Sorbent Disposal	1248.2	T/D	$11.00	$/T	$3,257,569
Plant Labor					
Oper Labor (w benefits)	23.0	men/shift	$25.50	$/h	$5,137,198
Supervision/support	25%	of above			$1,284,300
Maintenance	2.7%	of Direct			$9,824,580
Insurance & Local Taxes	0.9%	of Direct			$3,274,860
Other - % of Oper Labor	12.5%	of above			$642,150
By-Product Credit					$0
Net Operating Cost					**$28,910,311**

COSTS OF ELECTRICITY

Levelizing Factors	**Constant ($)**	**Basis (K$)**		**Annual (K$)**
Capital Charge	0.111	$723,622		$144,212
Fuel	1.025	$34,126		
Operating & Maintenance	1.000	$28,910		
Cost of Electricity - Levelized	**mills/kWh**			
Capital Charge	31.67			
Fuel	13.79			
Operating & Maintenance	11.40			
Total Cost of Electricity	**56.86**	Basis (MW)	445.4	
Energy-cycle Cost of Electricity	57.40	Basis (MW)	441.3	

TABLE 9.16 Operating Costs for Case 1: Oxygen-Blown KRW IGCC with Glycol CO_2 and H_2S Recovery and Gas Turbine Topping Cycle

Net Power (MW) =	**377.50**
Capacity factor=	**65%**
Annual Net Power Production (MW)=	**2,149,485**
Net Energy-cycle Power (MW)=	373.45

OPERATING COSTS	**Basis**	**Units**	**Unit Cost**		**Annual Cost**
Fuel - Illinois #6 Coal (ROM)	4,110	T/D	$35.00	$/T	**$34,126,136**
Coal - prepared	3,845	T/D			
Consumable material					
Catalyst, etc.					$1,895,096
Miscellaneous					$603,730
Ash/Sorbent Disposal	491.4	T/D	$11.00	$/T	$1,282,408
Plant Labor					
Oper Labor (w benefits)	23.0	men/shift	$25.50	$/h	$5,137,198
Supervision/support	25%	of above			$1,284,300
Maintenance	2.7%	of Direct			$11,308,637
Insurance & Local Taxes	0.9%	of Direct			$3,769,546
Other - % of Oper Labor	12.5%	of above			$642,150
By-Product Credit	102.1	TPD	$30.00	$/T	($726,857)
Net Operating Cost					**$25,196,207**

COSTS OF ELECTRICITY

Levelizing Factors	**Constant ($)**	**Basis (K$)**		**Annual (K$)**
Capital Charge	0.111	$825,905		$203,238
Fuel	1.025	$34,126		
Operating & Maintenance	1.000	$25,196		
Pipeline	1.000	$51,387		
Cost of Electricity - Levelized	**mills/kWh**			
Capital Charge	42.65			
Fuel	16.27			
Operating & Maintenance	11.72			
Pipeline	23.91			
Total Cost of Electricity	**94.55**	Basis (MW)	377.5	
Energy-cycle Cost of Electricity	95.58	Basis (MW)	373.5	

TABLE 9.17 Operating Costs for Case 2: Oxygen-Blown KRW IGCC with Membrane CO_2 Recovery, Glycol H_2S Recovery, and Gas Turbine Topping Cycle

			Net Power (MW) =		**330.00**
			Capacity factor=		**65%**
			Annual Net Power Production (MW)=		**1,879,020**
			Net Energy-cycle Power (MW)=		295.02
OPERATING COSTS	**Basis**	**Units**	**Unit Cost**		**Annual Cost**
Fuel - Illinois #6 Coal (ROM)	4,110	T/D	$35.00	$/T	**$34,126,136**
Coal - prepared	3,845	T/D			
Consumable material					
Catalyst, etc.					$1,895,096
Miscellaneous					$603,730
Ash/Sorbent Disposal	491.4	T/D	$11.00	$/T	$1,282,408
Plant Labor					
Oper Labor (w benefits)	23.0	men/shift	$25.50	$/h	$5,137,198
Supervision/support	25%	of above			$1,284,300
Maintenance	2.7%	of Direct			$13,475,622
Membrane Replacement (6 yr)	16.7%	of capital			$18,407,981
Insurance & Local Taxes	0.9%	of Direct			$4,491,874
Other - % of Oper Labor	12.5%	of above			$642,150
By-Product Credit	102.1	TPD	$30.00	$/T	($726,857)
Net Operating Cost					**$46,493,502**

COSTS OF ELECTRICITY				
Levelizing Factors	**Constant ($)**	**Basis (K$)**		**Annual (K$)**
Capital Charge	0.111	$986,271		$242,336
Fuel	1.025	$34,126		
Operating & Maintenance	1.000	$46,494		
Pipeline	1.000	$51,387		
Cost of Electricity - Levelized	**mills/kWh**			
Capital Charge	58.26			
Fuel	18.62			
Operating & Maintenance	24.74			
Pipeline	27.35			
Total Cost of Electricity	**128.97**	Basis (MW)	330.0	
Energy-cycle Cost of Electricity	144.26	Basis (MW)	295.0	

TABLE 9.18 Operating Costs for Case 3: Oxygen-Blown KRW IGCC with Glycol CO_2 Recovery, Methanol H_2S Recovery, and Fuel Cell Topping Cycle

Net Power (MW) =	**340.11**
Capacity factor=	**65%**
Annual Net Power Production (MW)=	**1,936,586**
Net Energy-cycle Power (MW)=	336.06

OPERATING COSTS	Basis	Units	Unit Cost		Annual Cost
Fuel - Illinois #6 Coal (ROM)	4,110	T/D	$35.00	$/T	**$34,126,136**
Coal - prepared	3,845	T/D			
Consumable material					
Catalyst, etc.					$1,895,096
Miscellaneous					$603,730
Ash/Sorbent Disposal	491.4	T/D	$11.00	$/T	$1,282,408
Plant Labor					
Oper Labor (w benefits)	23.0	men/shift	$25.50	$/h	$5,137,198
Supervision/support	25%	of above			$1,284,300
Maintenance	2.7%	of Direct			$15,856,718
Insurance & Local Taxes	0.9%	of Direct			$5,285,573
Other - % of Oper Labor	12.5%	of above			$642,150
By-Product Credit	102.1	TPD	$30.00	$/T	($726,857)
Net Operating Cost					**$31,260,315**

COSTS OF ELECTRICITY

Levelizing Factors	**Constant ($)**	**Basis (K$)**		**Annual (K$)**
Capital Charge	0.111	$1,190,627		$249,786
Fuel	1.025	$34,126		
Operating & Maintenance	1.000	$31,260		
Pipeline	1.000	$51,387		
Cost of Electricity - Levelized	**mills/kWh**			
Capital Charge	68.24			
Fuel	18.06			
Operating & Maintenance	16.14			
Pipeline	26.53			
Total Cost of Electricity	**128.98**	Basis (MW)	340.1	
Energy-cycle Cost of Electricity	130.54	Basis (MW)	336.1	

TABLE 9.19 Operating Costs for Case 4: Oxygen-Blown KRW IGCC with Membrane CO_2 Recovery, Methanol H_2S Recovery, and Fuel Cell Topping Cycle

Net Power (MW) =	**313.77**
Capacity factor=	**65%**
Annual Net Power Production (MW)=	**1,786,606**
Net Energy-cycle Power (MW)=	309.41

OPERATING COSTS	Basis	Units	Unit Cost		Annual Cost
Fuel - Illinois #6 Coal (ROM)	4,110	T/D	$35.00	$/T	**$34,126,136**
Coal - prepared	3,845	T/D			
Consumable material					
Catalyst, etc.					$1,895,096
Miscellaneous					$603,730
Ash/Sorbent Disposal	491.4	T/D	$11.00	$/T	$1,282,408
Plant Labor					
Oper Labor (w benefits)	23.0	men/shift	$25.50	$/h	$5,137,198
Supervision/support	25%	of above			$1,284,300
Maintenance	2.7%	of Direct			$19,814,807
Insurance & Local Taxes	0.9%	of Direct			$6,604,936
Other - % of Oper Labor	12.5%	of above			$642,150
By-Product Credit	102.1	TPD	$30.00	$/T	($726,857)
Net Operating Cost					**$36,537,767**

COSTS OF ELECTRICITY

Levelizing Factors	**Constant ($)**	**Basis (K$)**		**Annual (K$)**
Capital Charge	0.111	$1,483,273		$287,547
Fuel	1.025	$34,126		
Operating & Maintenance	1.000	$36,538		
Pipeline	1.000	$51,387		
Cost of Electricity - Levelized	**mills/kWh**			
Capital Charge	92.15			
Fuel	19.58			
Operating & Maintenance	20.45			
Pipeline	28.76			
Total Cost of Electricity	**160.95**	Basis (MW)	313.8	
Energy-cycle Cost of Electricity	163.21	Basis (MW)	309.4	

TABLE 9.20 Operating Costs for Optimal Air-Blown Case: KRW IGCC with Glycol CO_2 Recovery, In-Bed H_2S Recovery, and Gas Turbine Topping Cycle

Net Power (MW) =	**375.77**
Capacity factor=	**65%**
Annual Net Power Production (MW)=	**2,139,634**
Net Energy-cycle Power (MW)=	371.30

OPERATING COSTS	Basis	Units	Unit Cost		Annual Cost
Fuel - Illinois #6 Coal (ROM)	4,110	T/D	$35.00	$/T	**$34,126,136**
Coal - prepared	3,845	T/D			
Consumable material					
Catalyst, etc.					$0
Miscellaneous					$603,730
Ash/Sorbent Disposal	491.4	T/D	$11.00	$/T	$1,282,408
Plant Labor					
Oper Labor (w benefits)	23.0	men/shift	$25.50	$/h	$5,137,198
Supervision/support	25%	of above			$1,284,300
Maintenance	2.7%	of Direct			$11,453,237
Insurance & Local Taxes	0.9%	of Direct			$3,817,746
Other - % of Oper Labor	12.5%	of above			$642,150
By-Product Credit	0.0	TPD	$30.00	$/T	$0
Net Operating Cost					**$24,220,768**

COSTS OF ELECTRICITY

Levelizing Factors	**Constant ($)**	**Basis (K$)**		**Annual (K$)**
Capital Charge	0.111	$844,149		$204,288
Fuel	1.025	$34,126		
Operating & Maintenance	1.000	$24,221		
Pipeline	1.000	$51,387		
Cost of Electricity - Levelized	**mills/kWh**			
Capital Charge	43.79			
Fuel	16.35			
Operating & Maintenance	11.32			
Pipeline	24.02			
Total Cost of Electricity	**95.48**	Basis (MW)	375.8	
Energy-cycle Cost of Electricity	96.63	Basis (MW)	371.3	

TABLE 9.21 Summary of Comparative Costs of IGCC Systems

Case		BASE	BASE	Case #1	Case #2	Case#3	Case #4	ESD-24/Glycol
Gasifier Oxidant		Oxygen	Air	Oxygen	Oxygen	Oxygen	Oxygen	Air
H2S Recovery		Glycol	In-Bed/ZnTi	Glycol	Glycol	Methanol	Methanol	In-Bed/ZnTi
CO2 Recovery		none	none	Glycol	Membrane	Glycol	Membrane	Glycol
Topping Cycle		Turbine	Turbine	Turbine	Turbine	Fuel Cell	Fuel Cell	Turbine
Bottoming Cycle		Steam	Steam	Steam	Steam	Steam	Steam	Steam
Component	Unit							
Base Plant Capital	$/kW	$1,332	$1,253	$1,485	$1,703	$2,560	$2,746	$1,487
CO2 Control Capital	$/kW	$0	$0	$202	$602	$145	$905	$246
Total Plant Capital	**$/kW**	**$1,332**	**$1,253**	**$1,687**	**$2,305**	**$2,705**	**$3,651**	**$1,733**
Power Plant Annual Cost	$K	$137,253	$144,212	$203,238	$242,336	$249,786	$287,547	$204,288
Power Cost								
Base Plant Power Cost	mills/kWh	58.29	56.86	70.64	101.62	102.45	132.19	71.46
Pipeline Cost	mills/kWh	0	0	23.91	27.35	26.53	28.76	24.02
Net Power Cost	**mills/kWh**	**58.29**	**56.86**	**94.55**	**128.97**	**128.98**	**160.95**	**95.48**
Coal Energy Input	10^6Btu/h	3839	3839	3839	3839	3839	3839	3839
Gross Power Output	MW	458.20	479.63	446.40	417.60	418.50	413.20	460.88
In Plant Power Use	MW	44.70	34.19	68.90	87.60	78.39	99.40	85.11
Net Plant Output	**MW**	**413.50**	**445.44**	**377.50**	**330.00**	**340.11**	**313.80**	**375.77**
Net Heat Rate	Btu/kWh	9284	8618	10170	11633	11288	12234	10216
Thermal Efficiency - HHV	%	36.78%	39.62%	33.58%	29.35%	30.25%	27.91%	33.42%
Out of Plant Power Use	MW	2.41	4.18	4.05	3.87	4.05	4.12	4.47
Net Energy Cycle Power	**MW**	**411.09**	**441.26**	**373.45**	**326.13**	**336.06**	**309.68**	**371.30**
Net Energy Cycle Heat Rate	Btu/kWh	9339	8700	10280	11771	11424	12397	10339
Thermal Efficiency - HHV	%	36.56%	39.25%	33.21%	29.01%	29.89%	27.54%	33.02%
Net Energy Cycle Power	MW	411.09	441.26	373.45	326.13	336.06	309.68	371.30
Net Replacement [Added] Power	MW	0.00	(30.17)	37.64	84.96	75.03	101.41	39.79
Net Grid Power	**MW**	**411.09**	**411.09**	**411.09**	**411.09**	**411.09**	**411.09**	**411.09**

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