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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.

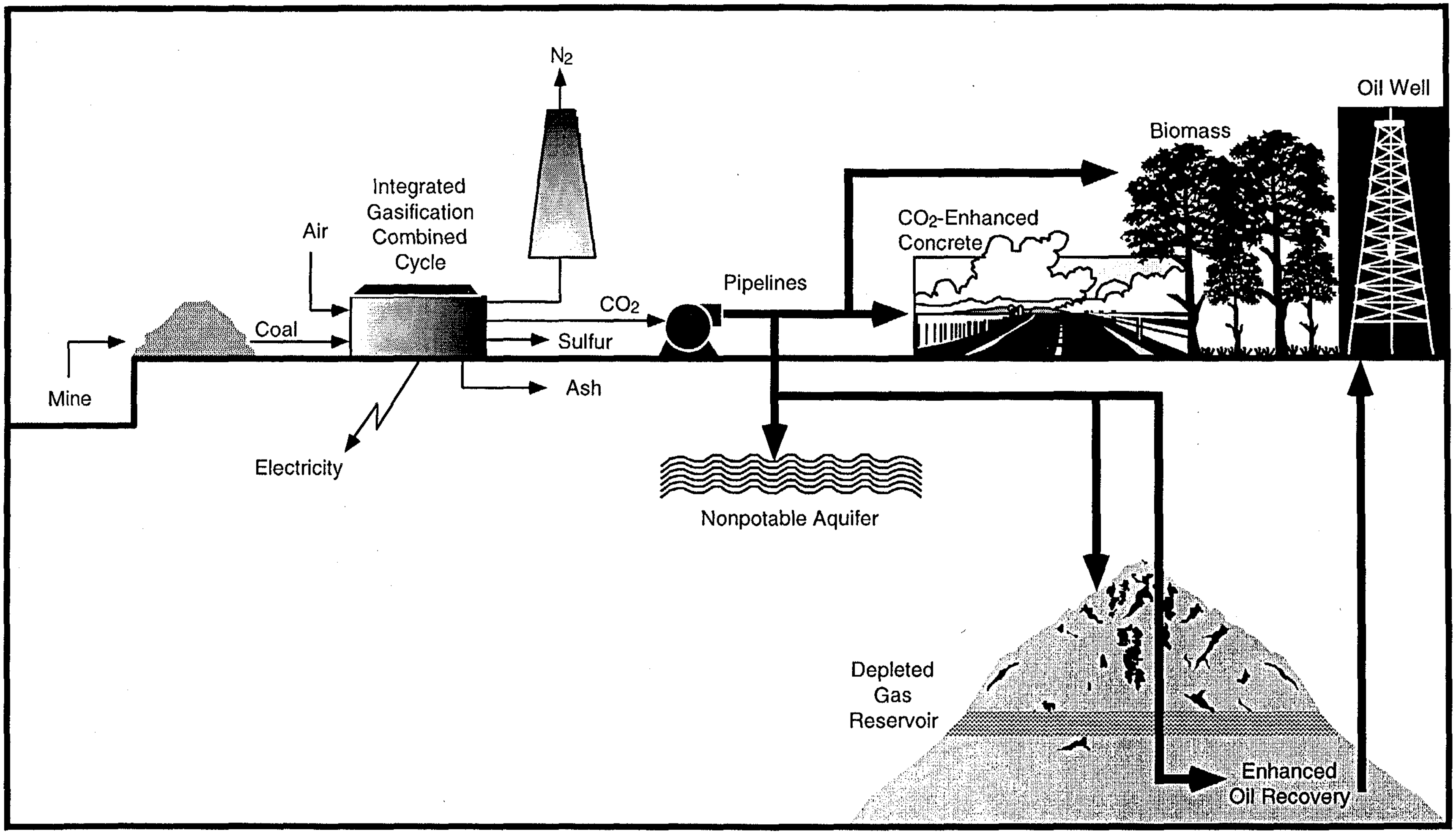

FIGURE S.1 Simplified Overview of the Energy Cycle Components for CO_2 Recovery

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.

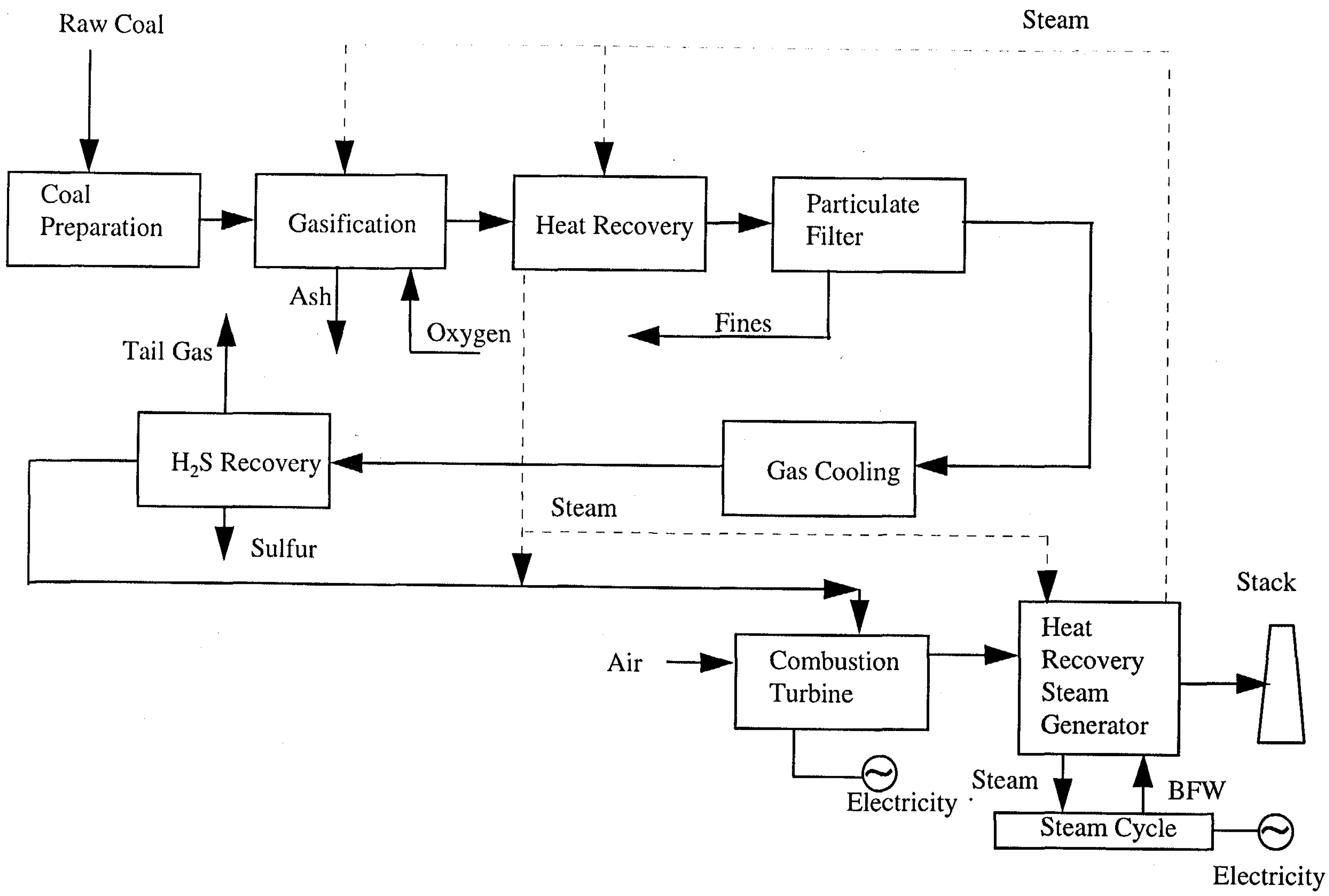

FIGURE S.2 Block Diagram of the Base-Case Oxygen-Blown KRW IGCC System

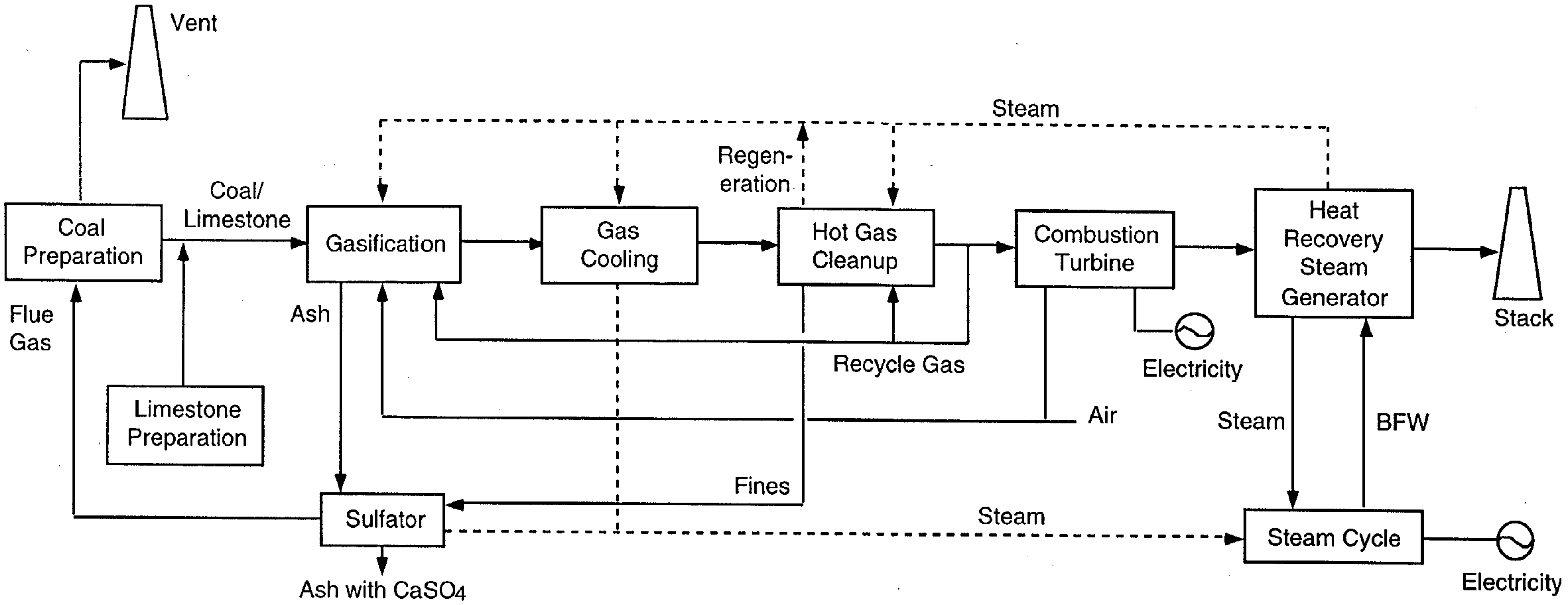

FIGURE S.3 Block Diagram of the Base-Case Air-Blown KRW IGCC System

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

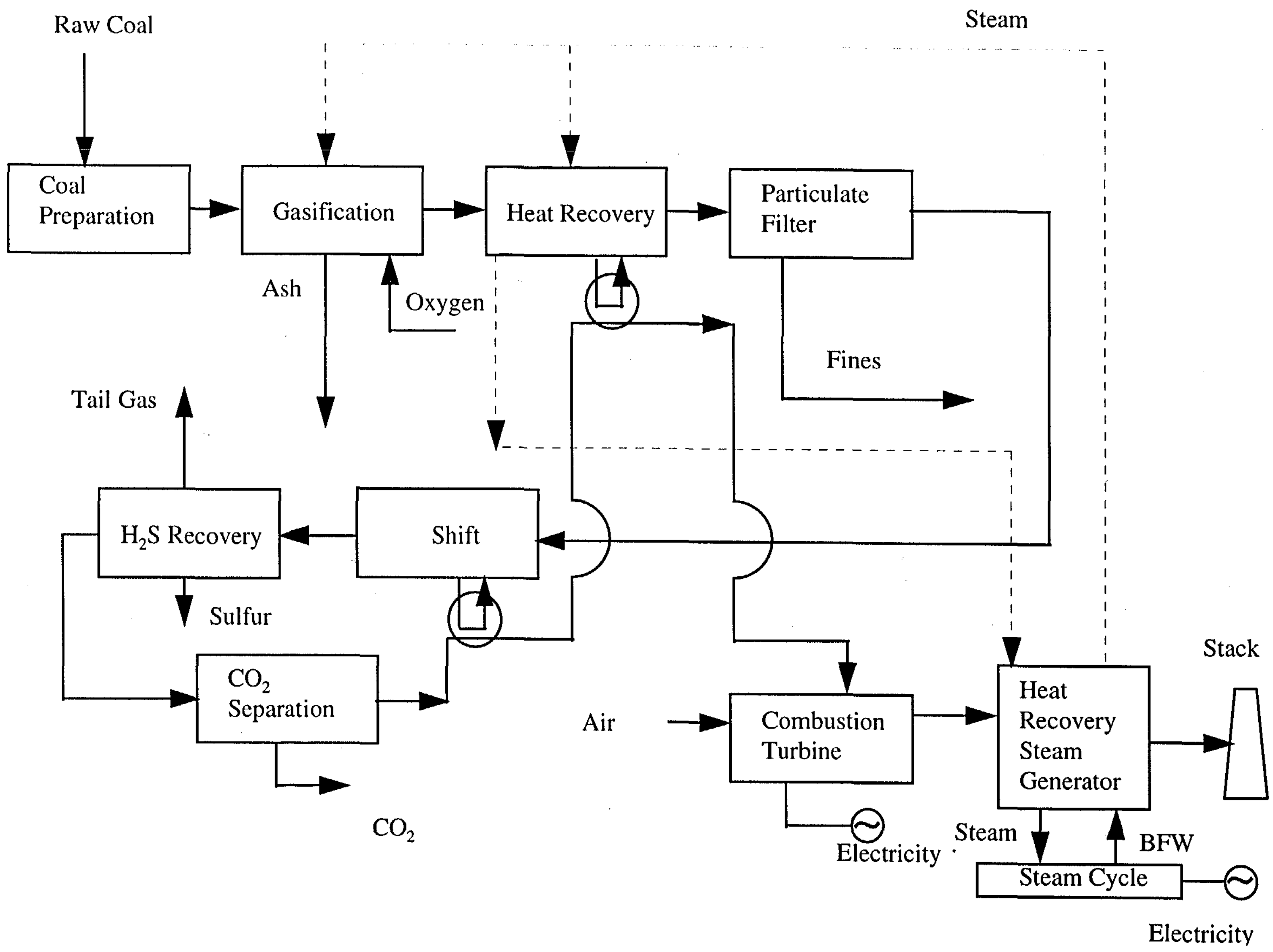

FIGURE S.4 Block Diagram of the Base-Case Oxygen-Blown KRW IGCC System Modified for CO_2 Recovery

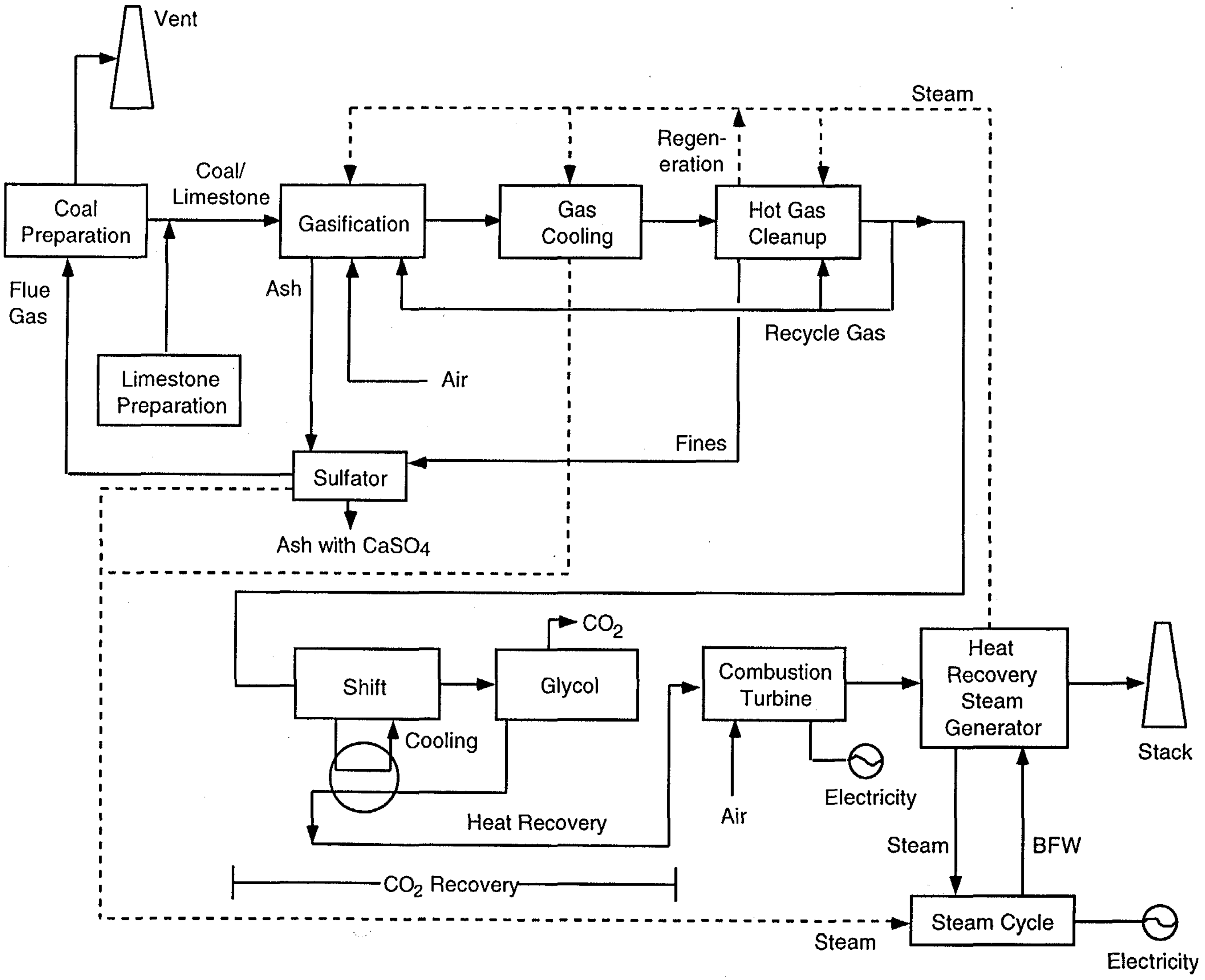

FIGURE S.5 Block Diagram of the Base-Case Air-Blown KRW IGCC System Modified for CO_2 Recovery heat to be absorbed before combustion and increases the mass flow rate through the gas turbine. The balance of the thermal energy is used in the heat recovery steam generator for feedwater heating and steam 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	CO2 release
Mining and Transport	MW	kg/h
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.

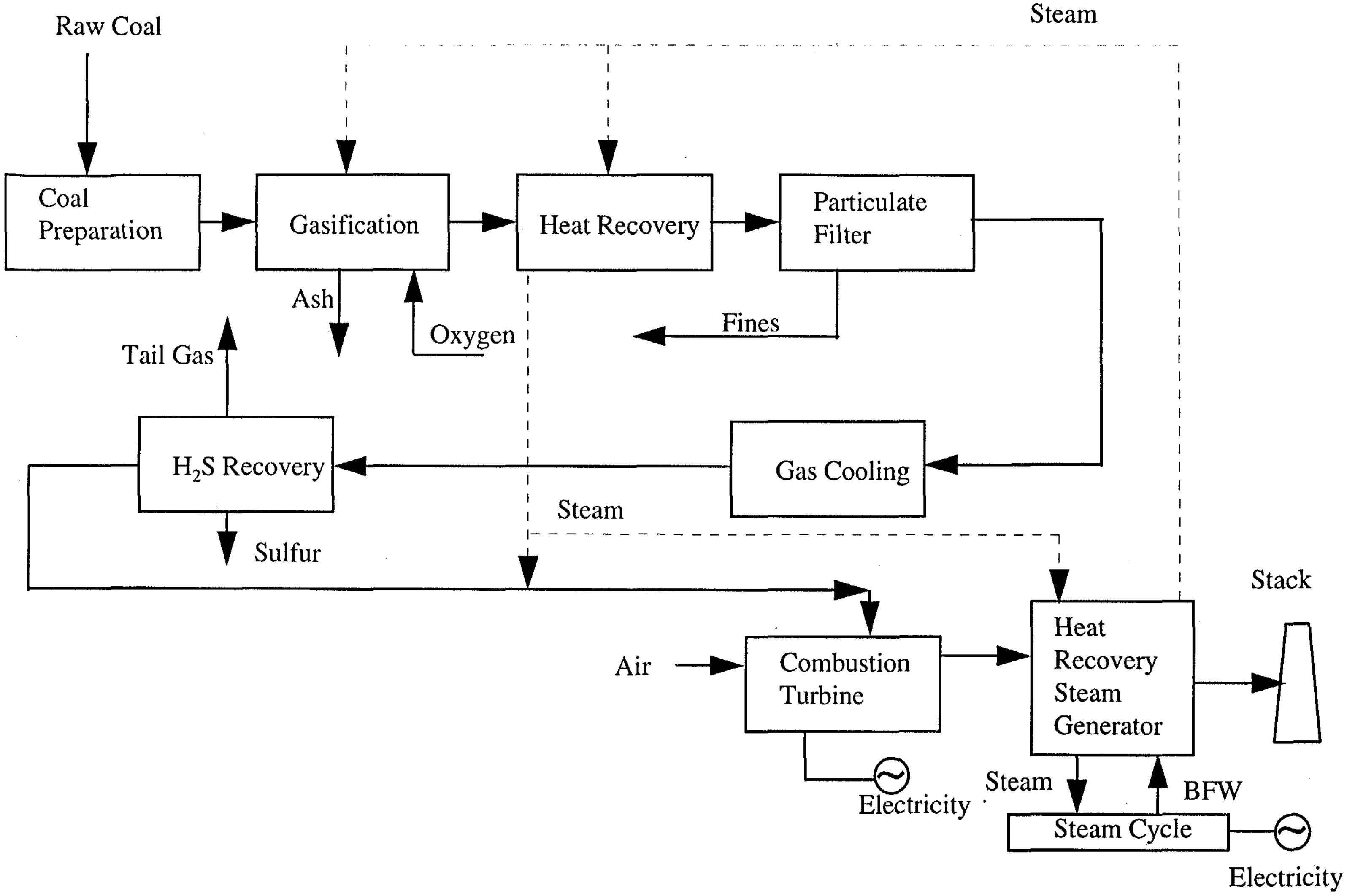

FIGURE 3.1 Block Diagram of the Base-Case Oxygen-Blown KRW IGCC System

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

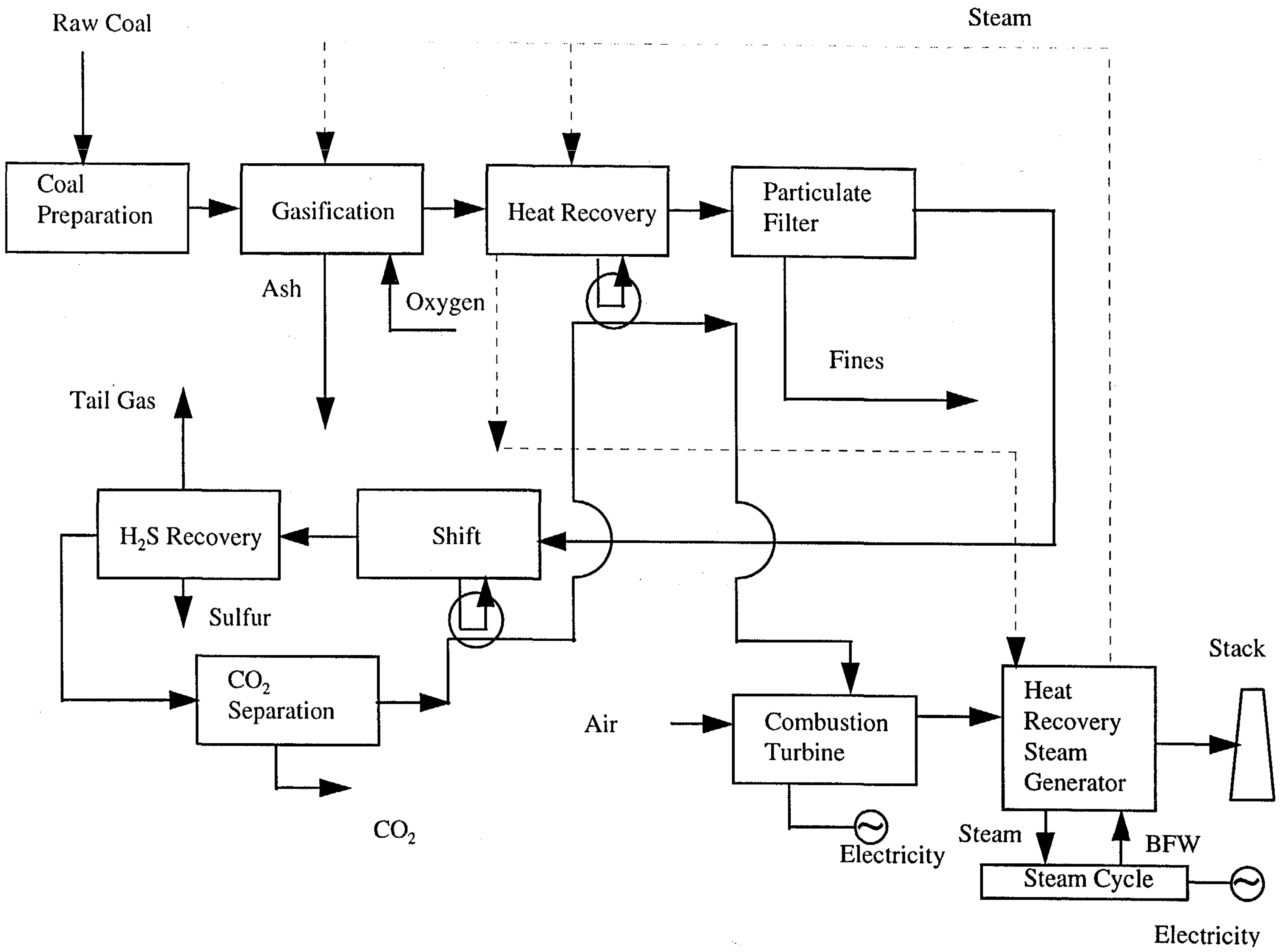

FIGURE 4.1 Block Diagram of the Base-Case Oxygen-Blown KRW IGCC System Modified for CO_2 Recovery

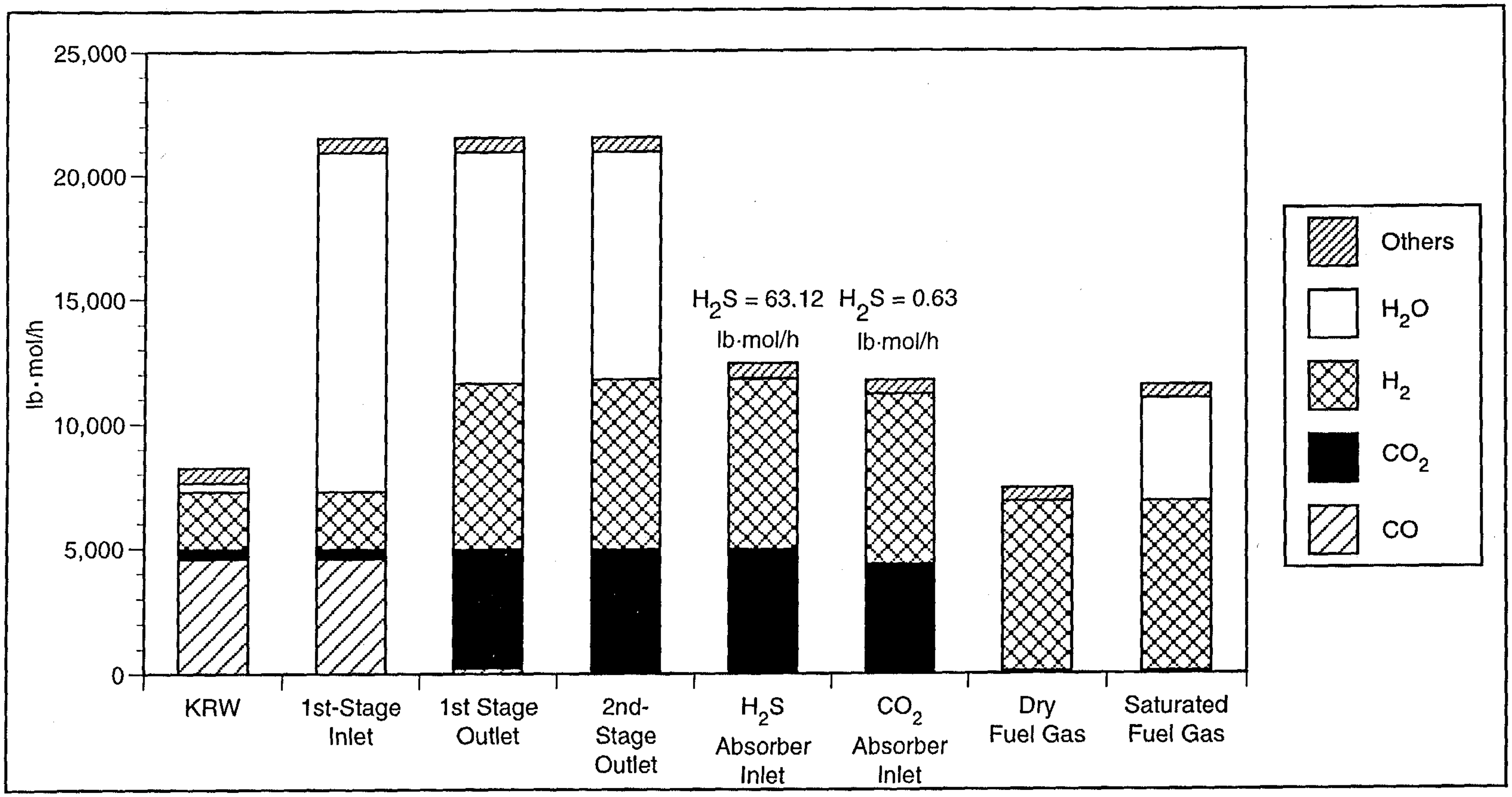

FIGURE 4.2 Gas Stream Composition at Various Stages in the Process in Case 1

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.

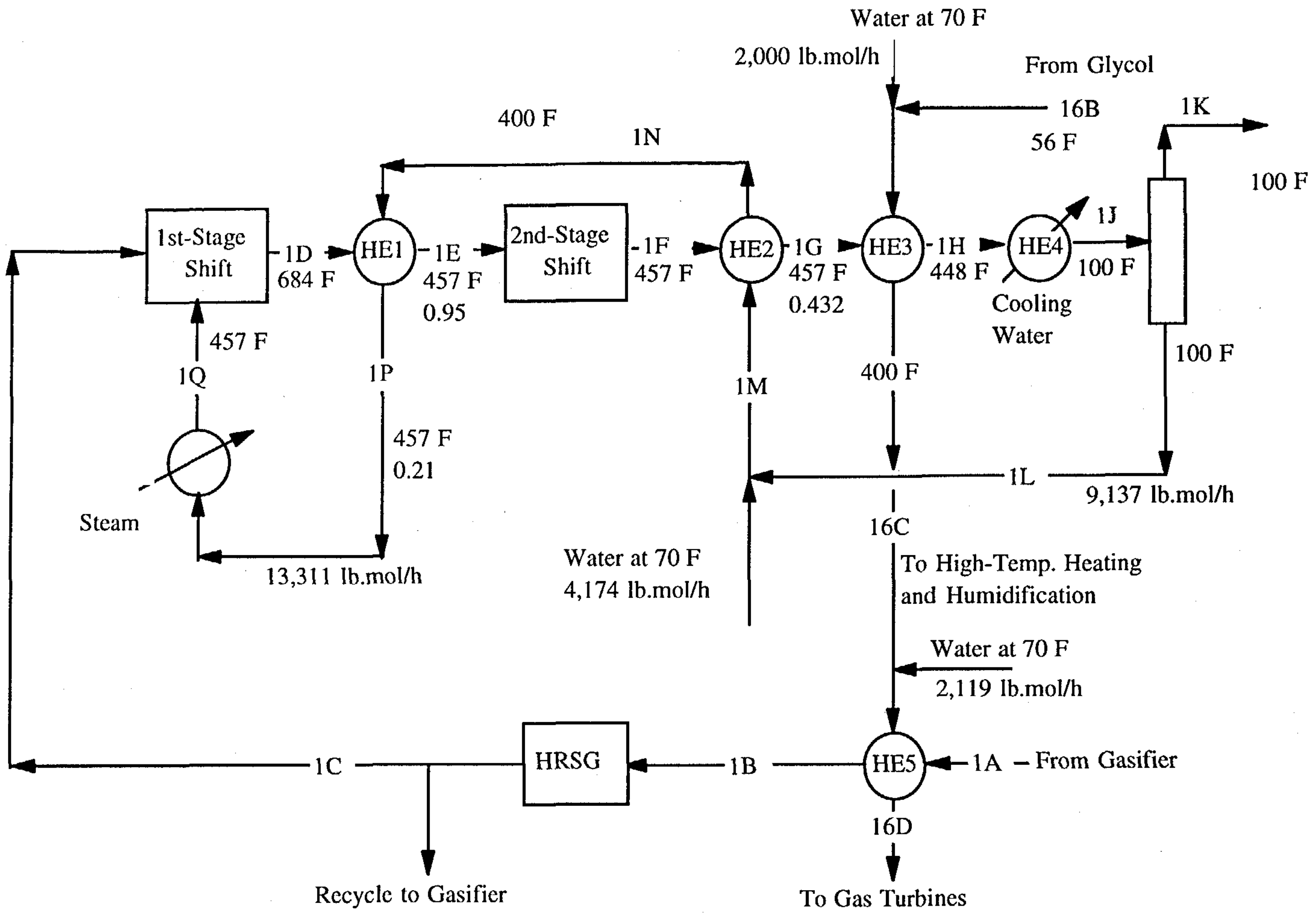

FIGURE 4.3 Flow Diagram of Shift System and Associated Heat Integration in Case 1

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)	13,311.44	13,311.44	13,311.44	0.00	0.00	0.00
H_2O						
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

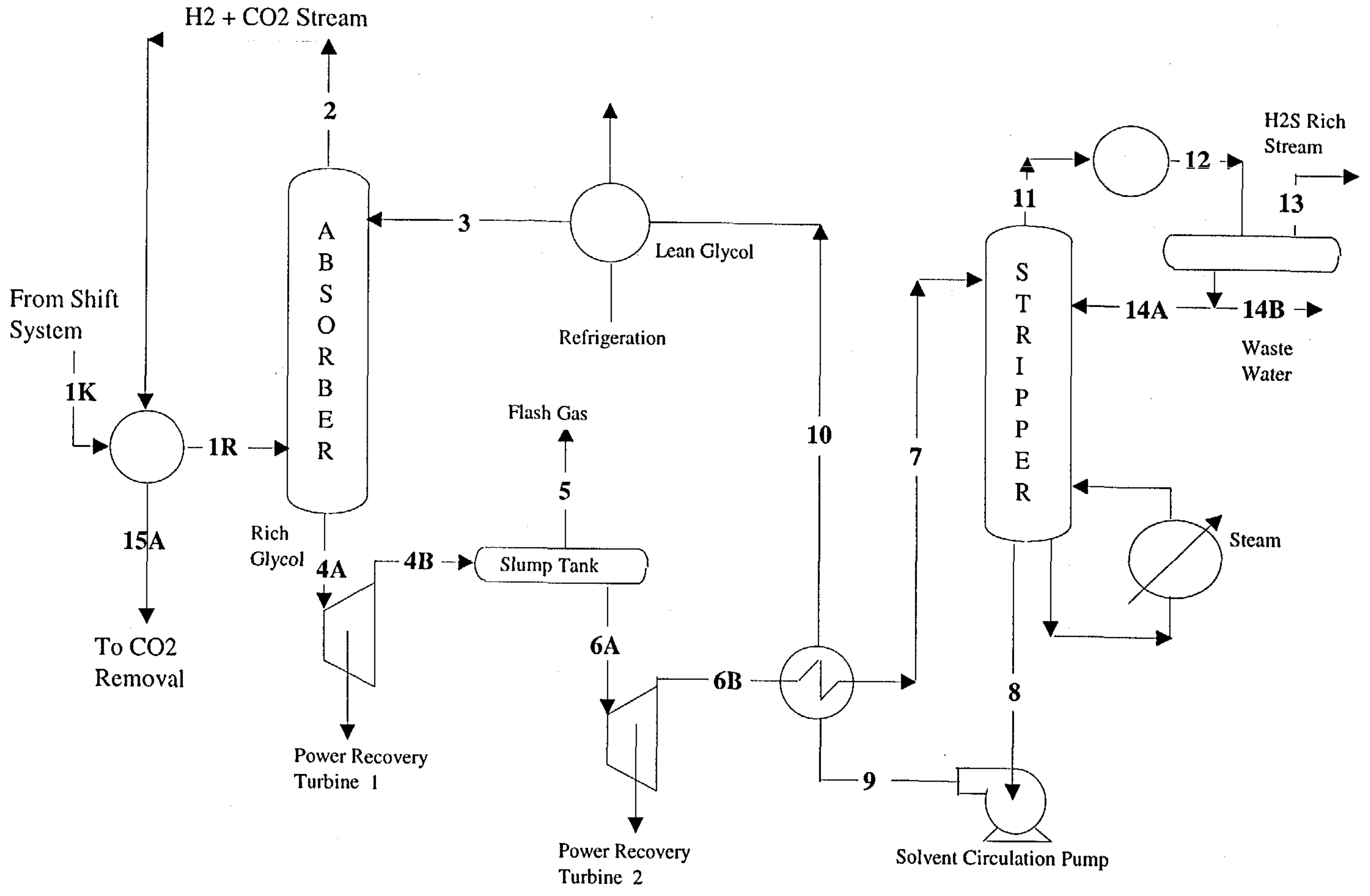

FIGURE 4.4 Flow Diagram of Glycol Process for H_2S Recovery in Case 1

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	

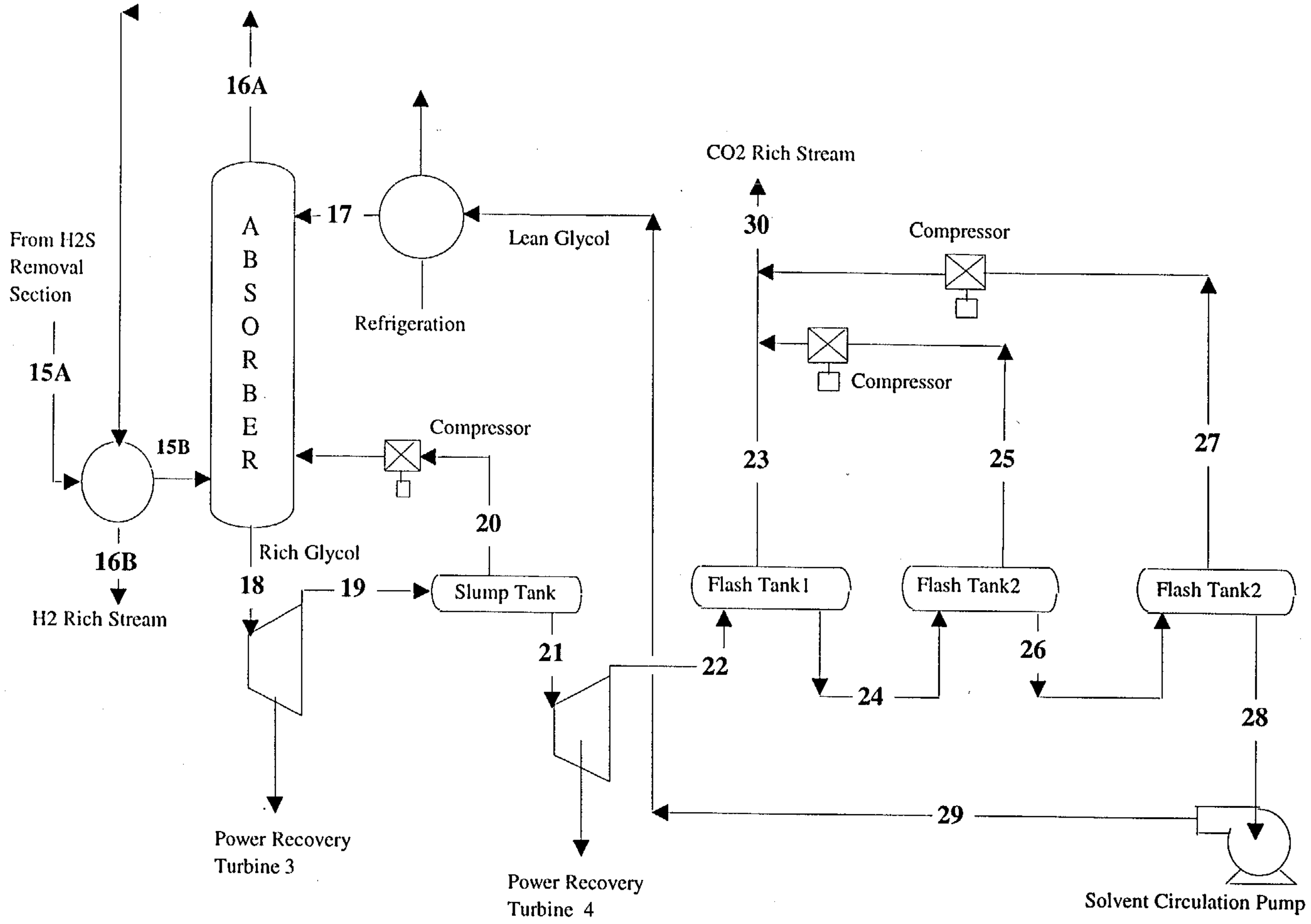

FIGURE 4.5 Flow Diagram of Glycol Process for CO_2 Recovery in Case 1

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	3.2	
	(includes all of the supporting equipment and connections and installation)		
	Purchased cost of heat exchanger in 1987	$160,000	
	(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		**$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	$44,000	
	(mild steel construction)		
	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	3.2	
	(includes all of the supporting equipment and connections and installation)		
	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		**$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	3.2	
	(includes all of the supporting equipment and connections and installation)		
	Purchased cost of heat exchanger in 1987	$118,000	
	(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		**$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

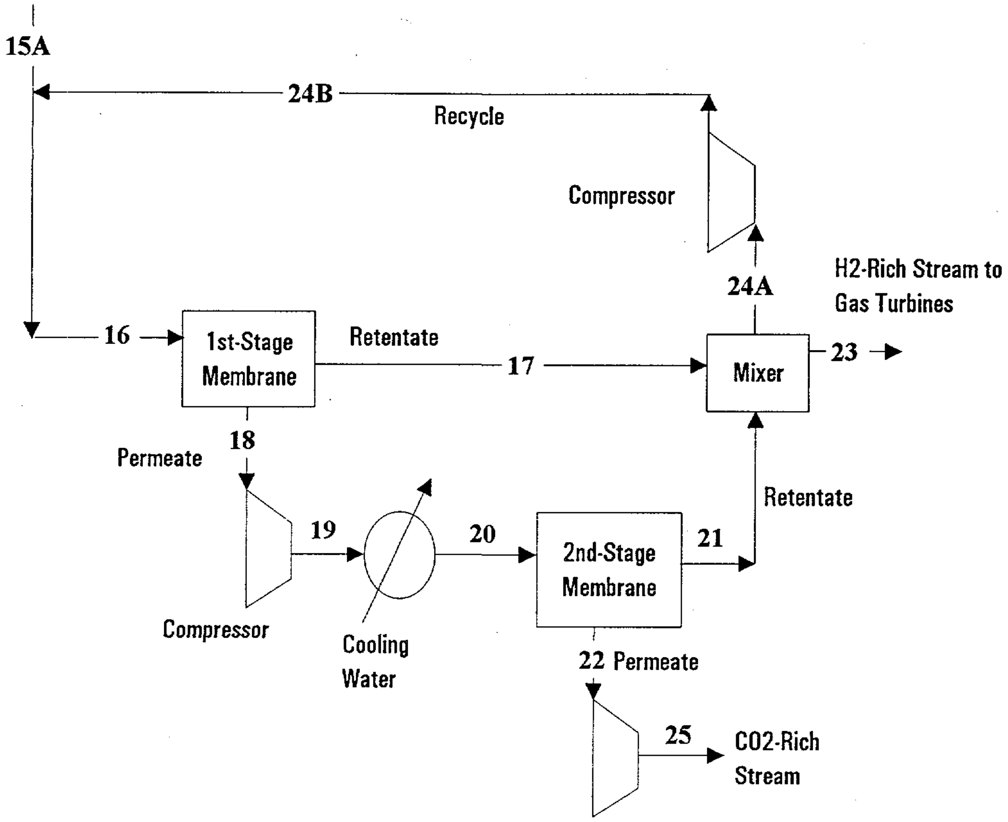

FIGURE 5.1 Flow Diagram of Membrane Process for CO_2 Recovery in Case 2

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 (MW)	
Power Variable	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 (MW)		
Power Variable	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	3.2	
	(includes all of the supporting equipment and connections and installation)		
	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 ft 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 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 (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)	$36,000	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of heat exchanger in 1995		**$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 (includes electric motor drive and gear reducer)	$540,000	
	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		**$1,640,486**
	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