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

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Notation

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

Acronyms, Initialisms, and Abbreviations

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

Chemicals

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

O ₂	oxygen
SO ₂	sulfur dioxide
SO _x	sulfur oxides

Units of Measure

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

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

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Abstract

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

Summary

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

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

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

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

FIGURE S.1 Annual Atmospheric Release of CO_2 and Carbon

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

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

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

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

S.1 Overview of Energy Cycle

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

FIGURE S.2 Energy System Components for CO₂ Recovery

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

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

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

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

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

Source: Gallaspy et al. (1990).

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

S.3 Handling of Coal and Limestone

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

S.4 Base Case for Integrated Gasification Combined Cycle

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

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

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

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

S.5 Integrated Gasification Combined Cycle with CO₂ Recovery

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

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

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

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

FIGURE S.4 Block Diagram of the IGCC System with CO₂ Recovery

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

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

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

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

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

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

S.6 Pipeline Transport of CO₂

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

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

S.7 Sequestering of CO₂

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

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

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

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

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

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

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

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

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

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

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

^a Not applicable.

FIGURE S.5 Net Power Generation and Associated CO₂ Inventory

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

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

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

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

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

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

S.9.2 Direct Costs for Pipeline

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

S.9.3 Costs of Replacement Electricity

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

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

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

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

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

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

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

S.10 References for Summary

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

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

1 Introduction

1.1 Background

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

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

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

1.2 Goals, Objectives, and Approach

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

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

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

2 Mining

2.1 Extraction

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

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

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

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

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

TABLE 2.2 Coal Losses and Feed Rates for IGCC Base Case

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

^a Not applicable.

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

2.1.1 Deep Mining

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

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

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

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

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

Source: Gallaspy et al. (1990b).

2.1.2 Coal-Cleaning Circuit

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

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

TABLE 2.4 Power Requirements for Underground Mining of Coal

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

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

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

^c NA, Not applicable.

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

2.2 Rail Transport

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

2.3 Impact of Plant Location on CO₂ Budget

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

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

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

2.4 Economics

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

2.5 Limestone Mining

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

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

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

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

TABLE 2.6 Power Requirements for Surface Mining of Limestone

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

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

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

^c NA, Not applicable.

2.6 Limestone Rail Transport

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

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

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

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

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

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

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

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

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

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

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

4 Kellogg-Rust-Westinghouse IGCC: Gasifier

In the case of direct coal combustion, the recovery of CO₂ is complicated by dilution of the flue gas with nitrogen (N₂) from the combustion air, as well as with excess air, which invariably accompanies coal combustion. For a representative boiler using bituminous coal, the mole fraction of CO₂ in the flue gas is in the range of 10-15% (Babcock & Wilcox Co. [B&W] 1975). For coal gasification, no excess air is provided. In fact, the air supplied is inadequate for complete combustion. The mole fraction of CO plus CO₂ in the fuel gas from the KRW gasifier is about 22% for an air-blown gasifier and 59% for an oxygen-blown gasifier (Gallaspy et al. 1990a). The CO can be converted to CO₂ via the water-gas shift reaction. Because of the higher CO₂ partial pressure, recovery of the CO₂ from this shifted synthesis gas requires a less costly recovery system than that which would be required for recovery following direct coal combustion in air. Sections 4 through 10 are intended to provide a first-order estimate of the effects of CO₂ recovery on the cost and performance of a gasifier.

4.1 Overview

4.1.1 Description of the Gasifier

The baseline system for gasification is the air-blown KRW gasifier, which is an agglomerating fluidized-bed gasifier. For the base-case system (case 1 [Gallaspy et al. 1990a]), hot-gas cleanup is employed.¹ The baseline capacity of the plant is 458 MW net. Figure 4.1 is a simplified flow diagram of the configuration in the base case with the system for CO₂ recovery added. Introduction of CO₂ recovery will reduce the net output of the plant because of additional energy losses incurred in various stages of CO₂ recovery. Specific energy penalties include the work of pumping and compression, the demand for heat or steam for solvent recovery, and thermal energy losses that are impractical to recover. In addition, some modifications to the combustion turbine will be required because of the accompanying change in the composition of the synthesis gas. No accounting of this latter effect is included in the present analysis.

Because no additional capacity for gasification has been added to compensate for energy losses due to the CO₂ recovery system, the output of the gasifier from the base case is taken to be the raw gas input to the CO₂ recovery system. The configuration of the gasifier in the base case uses four trains, each of which includes all major process equipment. Therefore, the preliminary design of the CO₂ recovery system described here also employs four trains, each at 25% of the necessary total capacity. The descriptions of the CO₂ recovery process, including equipment specifications, flow rates, and power requirements, refer to a single train. Descriptions of overall plant output and energy use refer to the aggregate of all four trains.

¹ The use of hot-gas cleanup on a system using solvent-based CO₂ recovery is of no value to the efficiency of the system because the temperature of the product gas must be reduced for effective absorption.

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

4.1.2 Shift Reaction

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

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

TABLE 4.1 Temperature Sensitivity
of Equilibrium Constant
for Shift Reaction

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

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

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

4.1.3 Recovery Process

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

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

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

FIGURE 4.2 Gas Stream Composition at Various Stages in the Process

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

4.2 Shift Reaction

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

TABLE 4.2 Composition of Combustion Chamber Inlet Gas

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

TABLE 4.3 Composition of Turbine Inlet Gas and Turbine Exhaust Gas

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

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

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

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

TABLE 4.4 Composition and Enthalpy of Streams

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

^a Streams do not include these gases.

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

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

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

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

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

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

where

k = reaction rate constant,

A_0 = initial concentration of CO,

B_0 = initial concentration of water,

C_0 = initial concentration of CO_2 ,

D_0 = initial concentration of H_2 , and

x = fraction conversion of CO to CO_2 ;

and

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

where

$$w = K - 1,$$

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

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

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

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

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

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

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

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

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

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

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

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

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

which can also be expressed as Equation 4.10:

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

where

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

x_e = fraction conversion at equilibrium.

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

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

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

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

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

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

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

4.3 Heat Recovery and Process Integration

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

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

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

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

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

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

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

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

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

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

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

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

^a Parentheses indicate power consumption.

^b HHV, Higher heating value.

The results in Table 4.7 depend on the following assumptions:

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

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

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

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

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

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

^a ΔH , Change in enthalpy.

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

5 Integrated Gasification Combined Cycle with Amine CO₂ Recovery

5.1 Feed Gas Preparation and Process Integration

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

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

TABLE 5.1 Heat Recovery and Allocation for MEA Process

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

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

5.2 Carbon Dioxide Recovery by Amines

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

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

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

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

^a Parentheses indicate power consumption.

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

TABLE 5.3 Stream Flows of MEA Process^a

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

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

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

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

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

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

TABLE 5.4 Descriptions of Streams of MEA Process^a

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

TABLE 5.4 (Cont.)

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

TABLE 5.4 (Cont.)

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

^a Flow rates are for one of four trains.

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

5.3 Power System Performance Effects of the MEA Process

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

5.4 Major Equipment Specification and Preliminary Cost Estimation

5.4.1 Heat Exchanger before First-Stage Shift Reactor

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

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

where

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

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

A is the heat transfer area (ft²), and

ΔT is the logarithmic mean temperature difference, defined as

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

where

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

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

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

5.4.2 Heat Exchanger between First and Second Shift Reactors

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

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

1. Heat Exchanger before First Stage for Raising Steam

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

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

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

TABLE 5.5 (Cont.)

3. Heat Exchanger after First Stage for Saturated Water

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

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

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

TABLE 5.5 (Cont.)

5. Gas-Liquid Heat Exchanger after Second Stage for Reboiler		
Q = Load (Btu/h)	66,700,501	
Tha = Inlet temperature of hot fluid (°F)	480	
Thb = Outlet temperature of hot fluid (°F)	417	
Pressure of hot gases (psia)	285	
Tca = Inlet temperature of cold fluid (MEA) (°F)	220	
Tcb = Outlet temperature of cold fluid (MEA) (°F)	240	
ΔT_1	240	
ΔT_2	197	
Log mean temperature difference (°F)	218	
Overall heat transfer coefficient (Btu/h/ft ² /°F)	40	
(hot fluid: gas, cold fluid: liquid) ^c		
Heat transfer area (ft ²)	7,649	
Operating pressure (psia)	285	
Pressure factor	1.165	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$90,000	
(mild steel construction; shell and tube floating head)		
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$377,879
6. Gas-Gas Heat Exchanger after Second Stage for Fuel Gas Heating		
Q = Load (Btu/h)	145,047,122	
Tha = Inlet temperature of hot fluid (°F)	417	
Thb = Outlet temperature of hot fluid (°F)	120	
Pressure of hot gases (psia)	285	
Tca = Inlet temperature of cold fluid (fuel gas) (°F)	100	
Tcb = Outlet temperature of cold fluid (fuel gas) (°F)	401	
ΔT_1	16	
ΔT_2	20	
Log mean temperature difference (°F)	18	
Overall heat transfer coefficient (Btu/h/ft ² /°F) ^d	5	
Heat transfer area (ft ²)	1,599,152	
Operating pressure (psia)	285	
Pressure factor	1.165	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$419,706	
(mild steel construction; shell and tube floating head)		
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$1,762,202

TABLE 5.5 (Cont.)

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

TABLE 5.5 (Cont.)

10. Cost of Shift Catalyst

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

11. CO₂ Absorption Column

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

12. CO₂ Stripping Column

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

TABLE 5.5 (Cont.)

13. Heat Exchanger for Lean Solvent Cooling

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

14. Interstage Cooler for Absorber

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

TABLE 5.5 (Cont.)

15. Lean-Rich MEA Heat Exchanger

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

16. Stripper Overhead Condenser

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

TABLE 5.5 (Cont.)

17. Stripper Reboiler

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

18. Solvent Circulation Pump

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

19. CO₂ Product Gas Compressors

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

TABLE 5.5 (Cont.)

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

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

^b CE, Chemical engineering.

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

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

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

^f Source: Garrett (1989).

5.4.3 Heat Exchanger after Second-Stage Shift Reactor

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

5.4.4 First-Stage Shift Reactor

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

5.4.5 Second-Stage Shift Reactor

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

5.4.6 Cost of Catalyst

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

5.4.7 Absorption Tower

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

5.4.8 Stripping Column

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

5.4.9 Lean Solvent Cooler

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

5.4.10 Interstage Cooler for Absorber

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

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

5.4.11 Lean-Rich MEA Heat Exchanger

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

5.4.12 Stripper Overhead Condenser

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

5.4.13 Stripper Reboiler

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

5.4.14 Solvent Circulation Pump

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

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

where

H_s = total dynamic head (psia),

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

η = efficiency of the pump, and

1,714 = unit conversion factor.

5.4.15 Carbon Dioxide Product Gas Compressors

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

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

where

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

P_1 = inlet pressure (psia),

P_2 = outlet pressure (psia),

k = Cp/Cv = 1.395 (assumed),

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

0.00436 = unit conversion factor.

6 Integrated Gasification Combined Cycle with Glycol Solvent CO₂ Recovery

Two alternative designs using a glycol solvent (dimethyl ether of polyethylene glycol) have been investigated on the basis of commercial application of the Selexol®³ process. These designs are a high-pressure (1,000-psia absorber) and a low-pressure (250-psia) case. These alternative processes are presented in parallel discussion in the following subsections.

6.1 Feed Gas Preparation and Process Integration

As discussed in Section 4.3, the most significant issue in process integration is the effective use of process heat from cooling of the synthesis gas as required for the shift reaction and gas preparation for the CO₂ absorber. The most effective use of this thermal energy is preheating of the fuel gas feed to the gas turbine. With the addition of moisture to that feed, it can absorb most of the available heat. The balance is used for feedwater heating, raising steam for the steam cycle, and solvent regeneration. Because of differences in the absorber inlet gas specifications among the alternative CO₂ recovery systems, some variation exists in the amount of heat available. For the glycol solvent process, the CO₂-rich gas to the absorber must be cooled to about 30°F. Cooling to 100°F is accomplished in the feed gas preparation described here. The final cooling is accomplished in heat exchange with lean fuel gas and by direct contact with solvent. Sufficient solvent is provided to maintain the necessary temperature. The amount and allocation of thermal energy available from various processes are listed in Table 6.1 for the high-pressure case and for the low-pressure case. The glycol solvent cases differ from the other recovery processes in that thermal regeneration of the solvent is not required. Solvent recovery is accomplished strictly through flashing.

6.2 Carbon Dioxide Recovery by a Glycol Solvent

The glycol solvent process for a commercial glycol solvent (Selexol; dimethyl ether of polyethylene glycol) is summarized in Figures 6.1 and 6.2. Carbon dioxide from the cooled, shifted synthesis gas is absorbed by lean solvent at high pressure. The use of high pressure and low temperature reduces the required solvent flow rate and vessel volume. Figure 6.3 displays the solubility as a function of partial pressure for CO₂ and other gases. High pressure also requires greater investment in compression and higher-strength piping and vessels. An alternative design, with lower pressure, has been evaluated for comparison. The low-pressure design requires no compression because absorption takes place at 250 psia, approximately the exit pressure for the shift reactor; however, more solvent is required to maintain CO₂ removal. The high-pressure

³ A commercial version of the glycol solvent process is Selexol®, which uses a proprietary mixture of the dimethyl ether of polyethylene glycol and related chemicals. The analysis presented here relies on published information on Selexol® and has benefited from review by the Selexol® licensee but does not represent an optimized Selexol® system.

TABLE 6.1 Heat Recovery and Allocation for Glycol Solvent Process

Process	Enthalpy Change Available from Process (10 ⁶ Btu)	Allocation to Fuel Gas Preheating (10 ⁶ Btu)	Allocation to Steam Cycle (10 ⁶ Btu)	Allocation to Solvent Regeneration (10 ⁶ Btu)
High-pressure case				
Cooling before first-stage shift	32.9	32.9	0	0
Cooling after first-stage shift	68.8	68.8	0	0
Cooling after second-stage shift	847.0	651.3	195.7	0
Low-pressure case				
Initial synthesis gas cooling to 1,000°F	1,299.3	302.3	997.0	0
Cooling before first-stage shift	32.9	32.9	0	0
Cooling after first-stage shift	68.8	68.8	0	0
Cooling after second-stage shift	847.0	571.8	275.2	0

absorber conditions of 1,000 psia and 30°F are described as typical of operation of a gas purification plant (Sweeny and Valentine 1970), although a minimum pressure of 150 psi has been suggested as adequate for economic operation (Shah and Huurdeman 1990). Carbon dioxide is recovered from the solvent by a combination of pressure reduction through pressure recovery turbines and flashing.

In addition to CH₄, CO₂, and H₂S, for which solubilities are given in Figure 6.3, the solubility in a glycol solvent (Selexol) for other gases, particularly steam, CO, and N₂, is needed for the process stream calculations. Solubility values for these gases were based on the solubility of CO₂, as given in Figure 6.3, and the relative solubilities listed in Table 6.2.

FIGURE 6.1 Summary of a Glycol Solvent Process: High-Pressure Case (numbers refer to streams)

FIGURE 6.2 Summary of a Glycol Solvent Process: Low-Pressure Case (numbers refer to streams)

FIGURE 6.3 Solubility of Acid Gases in a Glycol Solvent (Selexol) at 70°F (SCF, standard cubic feet) (Source: Sweeny and Valentine 1970)

In spite of the cooling and, in the high-pressure case, the compression of this stream to absorber design conditions, the relative amounts of CO, CO₂, and H₂ are assumed to be the same as those at the shift reactor exit. In the absence of any catalyst, no significant change in composition is expected over the short time required for compression and cooling. The major difference in stream composition is due to the removal of water by condensation. As shown in Table 4.4, the shifted syngas is 31 mol % water. Removal of water by condensation is done prior to compression to reduce the work of compression. Further cooling is then required to achieve an absorber inlet temperature of 100°F.

TABLE 6.2 Relative Solubilities of Gases in a Glycol Solvent^a

Gas	Solubility
CO ₂	1.0
N ₂	0
NH ₃	4.87
H ₂	0.013
H ₂ O	733
CO	0.028

^a Selexol.

Source: Bucklin and Schendel (1985).

An overview of the glycol solvent process is provided by Figures 6.1 and 6.2 (system flow diagrams for high-pressure and low-pressure cases, respectively) and by Tables 6.3 and 6.4 (the corresponding line lists). More detailed information and assumptions for each stream are provided in Table 6.5.

6.3 Power System Performance Effects of the Glycol Solvent Process

The fact that the synthesis gas must be cooled before processing by the CO₂ recovery system significantly affects the net power system output. This cooling substantially reduces the enthalpy of the synthesis gas, including the enthalpy contributed by the exothermic shift reaction. As discussed in Section 4.3, much of the thermal energy removed from the gas stream can be recovered, and the most effective use of that energy is preheating of humidified fuel gas. This effectiveness is shown in Table 4.7, where several alternative process integration strategies are compared. The heat demand for solvent regeneration in the glycol process is zero because solvent recovery is possible through flashing. As a result, the net power loss associated with the glycol-based CO₂ recovery is less than with the other CO₂ recovery systems. The overall power balance for the plant is summarized in Table 6.6. The energy penalty associated with CO₂ recovery includes both reduced output from the turbogenerator units and electricity for pumping, compressing, and refrigeration in the CO₂ recovery system. For the glycol solvent process, the total penalty is 6.1 MW in the low-pressure case and 79.8 MW in the high-pressure case.

6.4 Major Equipment Specification and Preliminary Cost Estimation

6.4.1 High-Pressure Glycol Solvent Case

6.4.1.1 Heat Exchanger before First-Stage Shift Reactor

The temperature of the gases from the KRW process is 1,000°F, and these gases have to be cooled to a temperature of 482°F. This temperature is found to be optimum for the first-stage shift reactor. The difference between the enthalpies of gases at these two temperatures gives the heat exchanger load. The majority of the heat available is used for raising saturated steam at 300 psia from saturated water. This saturated steam has to be injected into the first-stage shift reactor. During this process, the gases are cooled to a temperature of 531°F. Further cooling to the temperature of 482°F is achieved by heating and humidifying the fuel gas. The heat transfer area required for these two operations is estimated by using Equation 5.1, and the detailed calculations are given in Table 6.7.

TABLE 6.3 Stream Flows of a Glycol Solvent Process: High-Pressure Case^a

Stream Data	Stream 1A	Stream 1B	Stream 2A	Stream 2B	Stream 3	Stream 4A
Description of stream	Feed gas from shift system	Absorber feed	Fuel gas from absorber	Fuel gas after heat exchange	Lean glycol solvent	Rich glycol solvent from absorber
Gases (lb-mol/h)						
CO	77.37	77.37	76.60	76.60	0.00	77.37
CO ₂	4,335.99	4,335.99	216.93	216.93	115.55	4,234.74
H ₂	5,611.86	5,611.86	5,601.15	5,601.15	0.00	53.55
H ₂ O	61.91	61.91	0.30	0.30	0.07	61.91
N ₂	7,306.65	7,306.65	7,294.29	7,294.29	0.00	82.38
Ar	88.60	88.60	88.39	88.39	0.00	4.12
CH ₄	128.77	128.77	124.90	124.90	0.00	128.77
NH ₃	2.99	2.99	0.18	0.18	0.00	2.99
H ₂ S	0.40	0.40	0.04	0.04	0.00	0.40
HCl	0.04	0.04	0.00	0.00	0.00	0.04
O ₂	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00
SO ₂	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	17,614.58	17,614.58	13,402.79	13,402.79	115.62	4,646.27
Liquids (lb-mol/h)						
Glycol solvent	0.00	0.00	0.00	0.00	6,900.00	6,900.00
Temperature (°F)	100.00	68.17	30.00	75.00	30.00	83.82
Pressure (psia)	1,000.00	998.00	995.00	993.00	1,000.00	995.00
Enthalpy of stream (Btu/h) (reference, 32°F)	9.96×10^6	5.78×10^6	-1.80×10^5	4.00×10^6	-1.90×10^6	5.12×10^7

TABLE 6.3 Stream Flows of a Glycol Solvent Process: High-Pressure Case^a (Cont.)

Stream Data	Stream 4B	Stream 5	Stream 6	Stream 7	Stream 8	Stream 9	Stream 10
Description of stream	Rich glycol solvent after turbine 1	Recycle to absorber	Rich glycol solvent to turbine 2	Rich glycol solvent after turbine 2	CO ₂ -rich gas from 1st flash	Rich glycol solvent to 2nd flash	CO ₂ -rich gas from 2nd flash
Gases (lb-mol/h)							
CO	77.37	76.60	0.77	0.77	0.77	0.00	0.00
CO ₂	4,234.74	0.13	4,234.61	4,234.61	2,501.39	1,733.22	1,271.03
H ₂	53.55	42.84	10.71	10.71	10.71	0.00	0.00
H ₂ O	61.91	0.23	61.68	61.68	6.03	55.65	6.13
N ₂	82.38	70.03	12.36	12.36	12.36	0.00	0.00
Ar	4.12	3.91	0.21	0.21	0.21	0.00	0.00
CH ₄	128.77	124.90	3.86	3.86	3.86	0.00	0.00
NH ₃	2.99	0.18	2.81	2.81	2.81	0.00	0.00
H ₂ S	0.40	0.04	0.36	0.36	0.36	0.00	0.00
HCl	0.04	0.00	0.04	0.04	0.04	0.00	0.00
O ₂	0.00	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO ₂	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	4,646.27	318.86	4,327.41	4,327.41	2,538.53	1,788.87	1,277.16
Liquids (lb-mol/h)							
Glycol solvent	6,900.00	0.00	6,900.00	6,900.00	0.00	6,900.00	0.00
Temperature (°F)	82.69	82.69	82.69	79.57	50.48	50.48	35.05
Pressure (psia)	750.00	750.00	750.00	75.00	75.00	75.00	20.00
Enthalpy of stream (Btu/h) (reference, 32°F)	5.01×10^7	1.23×10^5	5.00×10^7	4.69×10^7	5.19×10^5	1.78×10^7	3.13×10^5

TABLE 6.3 Stream Flows of a Glycol Solvent Process: High-Pressure Case^a (Cont.)

Stream Data	Stream 11	Stream 12	Stream 13A	Stream 13B	Stream 13C	Stream 14
Description of stream	Rich glycol solvent to 3rd flash	CO ₂ -rich gas from 3rd flash	Lean glycol solvent	Lean glycol solvent after circulation pump	Lean glycol solvent after dehydration pump	CO ₂ -rich product
Gases (lb-mol/h)						
CO	0.00	0.00	0.00	0.00	0.00	0.77
CO ₂	462.19	346.64	115.55	115.55	115.55	4,119.06
H ₂	0.00	0.00	0.00	0.00	0.00	10.71
H ₂ O	49.52	2.60	46.92	46.92	0.07	14.76
N ₂	0.00	0.00	0.00	0.00	0.00	12.36
Ar	0.00	0.00	0.00	0.00	0.00	0.21
CH ₄	0.00	0.00	0.00	0.00	0.00	3.86
NH ₃	0.00	0.00	0.00	0.00	0.00	2.81
H ₂ S	0.00	0.00	0.00	0.00	0.00	0.36
HCl	0.00	0.00	0.00	0.00	0.00	0.04
O ₂	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00
SO ₂	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	511.71	349.24	162.47	162.47	115.62	4,164.94
Liquids (lb-mol/h)						
Glycol solvent	6,900.00	0.00	6,900.00	6,900.00	6,900.00	0.00
Temperature (°F)	35.05	30.74	30.74	38.47	38.47	142.58
Pressure (psia)	20.00	11.00	11.00	1,000.00	1,000.00	75.00
Enthalpy of stream (Btu/h) (reference, 32°F)	2.90×10^6	1.01×10^5	-1.19×10^6	6.14×10^6	6.14×10^6	4.44×10^6

^a Data are for a commercially available glycol solvent (Selexol). Flow rates are for one of four trains.

TABLE 6.4 Stream Flows of a Glycol Solvent Process: Low-Pressure Case^a

Stream Data	Stream 1A	Stream 1B	Stream 2A	Stream 2B	Stream 3	Stream 4A
Description of stream	Feed gas from shift system	Absorber feed	Fuel gas from absorber	Fuel gas after heat exchange	Lean glycol solvent	Rich glycol solvent from absorber
Gases (lb-mol/h)						
CO	77.37	77.37	76.60	76.60	0.00	77.37
CO ₂	4,355.99	4,355.99	218.15	218.15	395.00	4,514.19
H ₂	5,611.86	5,611.86	5,601.15	5,601.15	0.00	53.55
H ₂ O	61.91	61.91	0.49	0.49	2.25	64.16
N ₂	7,306.65	7,306.65	7,284.41	7,284.41	0.00	82.38
Ar	88.60	88.60	88.56	88.56	0.00	4.12
CH ₄	128.77	128.77	123.62	123.62	0.00	128.77
NH ₃	2.99	2.99	0.25	0.25	0.00	2.99
H ₂ S	0.40	0.40	0.05	0.05	0.00	0.40
HCl	0.04	0.04	0.01	0.01	0.00	0.04
O ₂	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00
SO ₂	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	17,614.58	17,614.58	13,393.30	13,393.30	397.25	4,927.97
Liquids (lb-mol/h)						
Glycol solvent	0.00	0.00	0.00	0.00	23,000.00	23,000.00
Temperature (°F)	100.00	68.13	30.00	75.00	30.00	46.68
Pressure (psia)	250.00	248.00	245.00	243.00	250.00	245.00
Enthalpy of stream (Btu/h) (reference, 32°F)	9.96×10^6	5.78×10^6	-1.85×10^5	4.00×10^6	-6.32×10^6	4.70×10^7

TABLE 6.4 Stream Flows of a Glycol Solvent Process: Low-Pressure Case^a (Cont.)

Stream Data	Stream 4B	Stream 5	Stream 6	Stream 7	Stream 8	Stream 9	Stream 10
Description of stream	Rich glycol solvent after turbine 1	Recycle to absorber	Rich glycol solvent to turbine 2	Rich glycol solvent after turbine 2	CO ₂ -rich gas from 1st flash	Rich glycol solvent to 2nd flash	CO ₂ -rich gas from 2nd flash
Gases (lb-mol/h)							
CO	77.37	76.60	0.77	0.77	0.77	0.00	0.00
CO ₂	4,514.19	1.35	4,512.83	4,512.83	2,227.15	2,285.68	941.70
H ₂	53.55	42.84	10.71	10.71	10.71	0.00	0.00
H ₂ O	64.16	0.49	63.67	63.67	10.27	53.40	6.45
N ₂	82.38	60.14	22.24	22.24	22.24	0.00	0.00
Ar	4.12	4.08	0.04	0.04	0.04	0.00	0.00
CH ₄	128.77	123.62	5.15	5.15	5.15	0.00	0.00
NH ₃	2.99	0.25	2.73	2.73	2.73	0.00	0.00
H ₂ S	0.40	0.05	0.35	0.35	0.35	0.00	0.00
HCl	0.04	0.01	0.03	0.03	0.03	0.00	0.00
O ₂	0.00	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO ₂	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	4,927.97	309.43	4,618.53	4,618.53	2,279.44	2,339.08	948.15
Liquids (lb-mol/h)							
Glycol solvent	23,000.00	0.00	23,000.00	23,000.00	0.00	23,000.00	0.00
Temperature (°F)	46.07	46.07	46.07	45.76	37.72	37.72	34.27
Pressure (psia)	100.00	100.00	100.00	25.00	25.00	25.00	14.70
Enthalpy of stream (Btu/h) (reference, 32°F)	4.50×10^7	3.18×10^4	4.50×10^7	4.40×10^7	2.90×10^5	1.82×10^7	1.29×10^5

TABLE 6.4 Stream Flows of a Glycol Solvent Process: Low-Pressure Case^a (Cont.)

Stream Data	Stream 11	Stream 12	Stream 13A	Stream 13B	Stream 13C	Stream 14
Description of stream	Rich glycol solvent to 3rd flash	CO ₂ -rich gas from 3rd flash	Lean glycol solvent	Lean glycol solvent after circulation pump	Lean glycol solvent after dehydration pump	CO ₂ -rich product
Gases (lb-mol/h)						
CO	0.00	0.00	0.00	0.00	0.00	0.77
CO ₂	1,343.98	948.98	395.00	395.00	395.00	4,117.83
H ₂	0.00	0.00	0.00	0.00	0.00	10.71
H ₂ O	46.95	20.64	26.31	26.31	2.30	37.36
N ₂	0.00	0.00	0.00	0.00	0.00	22.24
Ar	0.00	0.00	0.00	0.00	0.00	0.04
CH ₄	0.00	0.00	0.00	0.00	0.00	5.15
NH ₃	0.00	0.00	0.00	0.00	0.00	2.73
H ₂ S	0.00	0.00	0.00	0.00	0.00	0.35
HCl	0.00	0.00	0.00	0.00	0.00	0.03
O ₂	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00
SO ₂	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	1,390.93	969.62	421.31	421.31	397.30	4,197.22
Liquids (lb-mol/h)						
Glycol solvent	23,000.00	0.00	23,000.00	23,000.00	23,000.00	0.00
Temperature (°F)	34.27	30.64	30.64	32.56	32.56	124.54
Pressure (psia)	14.70	4.00	4.00	250.00	250.00	25.00
Enthalpy of stream (Btu/h) (reference, 32°F)	7.20×10^6	3.39×10^5	-4.31×10^6	1.77×10^6	1.77×10^6	4.17×10^6

^a Data are for a commercially available glycol solvent (Selexol). Flow rates are for one of four trains.

TABLE 6.5 Descriptions of Streams of a Glycol Solvent Process^a

Stream and Characteristic	High-Pressure Case	Low-Pressure Case	Comments on Stream Calculations
Stream 1A: CO ₂ -rich gas feed to absorber			
Temperature (°F)	100	100	—
Pressure (psia)	1,000	250	
Flow rate (lb-mol/h)	17,615	17,615	
CO ₂ (mole fraction)	0.246	0.246	
Stream 2A: CO ₂ -lean gas			
Temperature (°F)	30	30	Lean gas stream reflects design specification of 95% CO ₂ removal in absorber. In addition, virtually all remaining water in input stream is absorbed, along with small amount of H ₂ and CH ₄ . Absorption is based on equilibrium values from Figure 6.3 and Table 6.1. Because equilibrium will not be attained, excess solvent is used.
Pressure (psia)	995	245	
Flow rate (lb-mol/h)	13,403	13,393	
CO ₂ (mole fraction)	0.016	0.016	
Stream 3: Lean glycol solvent			
Temperature (°F)	30	30	Lean glycol solvent contains residual CO ₂ as indicated and 0.01% H ₂ O. In high-pressure case, flow rate is 135% of amount of glycol solvent required to contain 95% of input CO ₂ at equilibrium under absorber conditions. In low-pressure case, only 150% of equilibrium requirement is used to limit vessel size.
Pressure (psia)	1,000	250	
Flow rate (lb-mol/h)	7,016	23,397	
CO ₂ (mole fraction)	0.017	0.017	
Stream 4A: Rich glycol solvent; high pressure			
Temperature (°F)	84	47	Flow rate reflects lean glycol solvent, plus absorbed CO ₂ and H ₂ O and small amounts of other absorbed species. Increase in temperature is because of heat of absorption. In absence of actual values, heat of absorption for CO ₂ has been taken to equal latent heat of condensation. Heat of absorption of other species has been ignored.
Pressure (psia)	995	245	
Flow rate (lb-mol/h)	11,546	27,923	
CO ₂ (mole fraction)	0.367	0.162	

TABLE 6.5 (Cont.)

Stream and Characteristic	High-Pressure Case	Low-Pressure Case	Comments on Stream Calculations
Stream 4B: Rich glycol solvent; medium pressure			
Temperature (°F)	83	46	This stream is exit stream from high-pressure power recovery turbine. Exit pressure has been selected to avoid release of CO ₂ while allowing some recovery of work of pressurization. Carbon dioxide is not released at this pressure because of large amount of excess glycol solvent introduced in absorber. Change in temperature over pressure recovery turbine is estimated from change in enthalpy, which is taken to be equal to flow work, $\int v dp$. The overall efficiency of the turbine is assumed to be 72%.
Pressure (psia)	750	100	
Flow rate (lb-mol/h)	11,546	27,923	
CO ₂ (mole fraction)	0.367	0.162	
Stream 5: H ₂ recycle			
Temperature (°F)	83	46	Carbon monoxide, H ₂ , and CH ₄ are released in slump tank and returned to absorber. In this ideal system, released gas could be sent directly to fuel gas; but in actual system, some contamination with CO ₂ is likely, so that recycle to absorber is recommended.
Pressure (psia)	750	100	
Flow rate (lb-mol/h)	319	309	
CO ₂ (mole fraction)	0.001	0.004	
CH ₄ (mole fraction)	0.392	0.400	
H ₂ (mole fraction)	0.134	0.139	
CO (mole fraction)	0.240	0.248	
Stream 6: Inlet to low-pressure power recovery turbine			
Temperature (°F)	83	46	Change in composition simply reflects flashing of fuel gases to stream 5.
Pressure (psia)	750	100	
Flow rate (lb-mol/h)	11,227	27,619	
CO ₂ (mole fraction)	0.377	0.163	
Stream 7: Rich glycol solvent; low pressure			
Temperature (°F)	80	46	This stream is exit stream from low-pressure power recovery turbine. Residence time in turbine is insufficient to permit significant desorption of CO ₂ . Change in temperature has been calculated as for stream 4B.
Pressure (psia)	75	25	
Flow rate (lb-mol/h)	11,227	27,619	
CO ₂ (mole fraction)	0.377	0.163	

TABLE 6.5 (Cont.)

Stream and Characteristic	High- Pressure Case	Low- Pressure Case	Comments on Stream Calculations
Stream 8: Flash gas from high-pressure flash tank			
Temperature (°F)	50	38	For simplicity, minor constituents have been ignored, leaving this pure CO ₂ stream. Amount of CO ₂ remaining in solvent is estimated by assuming equilibrium at flash pressure. Amount released is calculated from mass balance.
Pressure (psia)	75	25	
Flow rate (lb-mol/h)	2,539	2,279	
CO ₂ (mole fraction)	0.985	0.977	
Stream 9: Glycol solvent from high-pressure flash tank			
Temperature (°F)	50	38	—
Pressure (psia)	75	25	
Flow rate (lb-mol/h)	8,689	25,339	
CO ₂ (mole fraction)	0.199	0.090	
Stream 10: Flash gas from medium-pressure flash tank			
Temperature (°F)	35	34	Equilibrium is assumed to be attained in flash tank.
Pressure (psia)	20	14.7	
Flow rate (lb-mol/h)	1,277	948	
CO ₂ (mole fraction)	0.995	0.993	
Stream 11: Glycol solvent from medium-pressure flash tank			
Temperature (°F)	35	34	—
Pressure (psia)	20	14.7	
Flow rate (lb-mol/h)	7,412	24,391	
CO ₂ (mole fraction)	0.062	0.055	
Stream 12: Flash Gas from low-pressure flash tank			
Temperature (°F)	31	31	Small amount of H ₂ O is released at this pressure, along with CO ₂ .
Pressure (psia)	11	4	
Flow rate (lb-mol/h)	349	970	
CO ₂ (mole fraction)	0.993	0.979	

TABLE 6.5 (Cont.)

Stream and Characteristic	High- Pressure Case	Low- Pressure Case	Comments on Stream Calculations
Stream 13A: Low-pressure lean glycol solvent			
Temperature (°F)	31	31	—
Pressure (psia)	11	4	
Flow rate (lb-mol/h)	7,062	23,421	
CO ₂ (mole fraction)	0.016	0.017	
Stream 13B: High-pressure lean glycol solvent			
Temperature (°F)	38	33	—
Pressure (psia)	1,000	250	
Flow rate (lb-mol/h)	7,062	23,421	
CO ₂ (mole fraction)	0.016	0.017	
Stream 13C: Dehydrated high-pressure lean glycol solvent			
Temperature (°F)	38	33	—
Pressure (psia)	1,000	250	
Flow rate (lb-mol/h)	7,016	23,397	
CO ₂ (mole fraction)	0.017	0.017	
Stream 14: CO ₂ product; summary of stream characteristics			
Temperature (°F)	143	125	—
Pressure (psia)	75	25	
Flow rate (lb-mol/h)	4,165	4,197	
CO ₂ (mole fraction)	0.989	0.981	

^a Data are for a commercially available glycol solvent (Selexol). Flow rates are for one of four trains.

TABLE 6.6 Turbine Power Output, Plant Power Use, and Net Power Output for Glycol Solvent Process

Power Variable	Power (MW)		
	Base Case	Low Pressure	High Pressure
Power output			
Gas turbine	311.6	281.6	281.8
Steam turbine	182.2	191.4	187.7
Expansion turbine	— ^a	—	23.3
Internal power consumption ^b			
CO ₂ recovery			
CO ₂ compression	0	(33.8)	(72.4)
Solvent circulation	0	(7.1)	(8.6)
Solvent refrigeration	0	(2.7)	(2.7)
Others	0	3.4	4.9
Gasifier system	(35.4)	(35.4)	(35.4)
Net power output	458.4	397.4	378.6
Energy penalty	0	61	79.8

^a Not applicable.

^b Parentheses indicate power consumption.

6.4.1.2 Heat Exchanger between First and Second Shift Reactors

The shift reaction is an exothermic reaction, and the heat released in the first stage has to be removed to obtain an optimum temperature for the second-stage shift reactor. The gases have to be cooled from 680°F to 460°F. The heat load of the heat exchanger is again the difference between the enthalpies of the gases. Most of the available heat is used in heating the water from 100°F to the saturation temperature of 417.4°F at 300 psia. During this process, the gases are cooled to a temperature of 537°F. Further cooling to the required temperature of 460°F is achieved by heating and humidifying the fuel gas. The heat transfer area is estimated by using Equation 5.1, and the installed cost of the heat exchanger is given in Table 6.7.

6.4.1.3 Heat Exchanger after Second-Stage Shift Reactor

The shifted gases from the second stage are at a temperature of 480°F. These gases have to be cooled to a temperature of 100°F. The cooling is achieved in stages. First, the high-pressure fuel gas is heated from 75°F to a temperature of 401°F before the expansion turbine. Expansion of

TABLE 6.7 Sizing and Cost Estimation of Major Equipment for High-Pressure Glycol Solvent Process (Selexol)

1. Heat Exchanger before First Stage for Raising Steam

Q = Load (Btu/h)	78,838,137	
Tha = Inlet temperature of hot fluid (°F)	1,000	
Thb = Outlet temperature of hot fluid (°F)	531	
Pressure of hot gases (psia)	295	
Tca = Inlet temperature of saturated water (°F)	417	
Tcb = Outlet temperature of saturated steam (°F)	417	
ΔT_1	583	
ΔT_2	114	
Log mean temperature difference (°F)	287	
Overall heat transfer coefficient (Btu/h/ft ² /°F) ^a	260	
Heat transfer area (ft ²)	1,057	
Operating pressure (psia)	300	
Pressure factor	1.165	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	\$23,000	
CE index for process equipment in 1987 ^b	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$96,569

2. Gas-Gas Heat Exchanger before First Stage for Heating Fuel Gas

Q = Load (Btu/h)	8,219,649	
Tha = Inlet temperature of hot fluid (°F)	531	
Thb = Outlet temperature of hot fluid (°F)	482	
Pressure of hot gases (psia)	295	
Tca = Inlet temperature of saturated water (°F)	100	
Tcb = Outlet temperature of saturated steam (°F)	510	
ΔT_1	21	
ΔT_2	382	
Log mean temperature difference (°F)	124	
Overall heat transfer coefficient (Btu/h/ft ² /°F) ^a	5	
Heat transfer area (ft ²)	13,210	
Operating pressure (psia)	300	
Pressure factor	1.165	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	\$122,616	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$514,823

TABLE 6.7 (Cont.)

3. Heat Exchanger after First Stage for Saturation Water

Q = Load (Btu/h)	31,747,489	
Tha = Inlet temperature of hot fluid (°F)	680	
Thb = Outlet temperature of hot fluid (°F)	537	
Pressure of hot gases (psia)	285	
Tca = Inlet temperature of water (°F)	100	
Tcb = Outlet temperature of saturated water (°F)	417	
ΔT_1	263	
ΔT_2	437	
Log mean temperature difference (°F)	342	
Overall heat transfer coefficient (Btu/h/ft ² /°F)	40	
(hot fluid: gas, cold fluid: gas) ^c		
Heat transfer area (ft ²)	2,319	
Operating pressure (psia)	300	
Pressure factor	1.165	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$40,000	
(mild steel construction; shell and tube floating head)		
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$167,946

4. Gas-Gas Heat Exchanger after First Stage for Heating Fuel Gas

Q = Load (Btu/h)	17,195,044	
Tha = Inlet temperature of hot fluid (°F)	537	
Thb = Outlet temperature of hot fluid (°F)	460	
Pressure of hot gases (psia)	295	
Tca = Inlet fuel gas/water temperature (°F)	100	
Tcb = Outlet fuel gas/water temperature (°F)	510	
ΔT_1	27	
ΔT_2	360	
Log mean temperature difference (°F)	128	
Overall heat transfer coefficient (Btu/h/ft ² /°F) ^a	5	
Heat transfer area (ft ²)	26,872	
Operating pressure (psia)	300	
Pressure factor	1.165	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$161,742	
(mild steel construction; shell and tube floating head)		
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$679,099

TABLE 6.7 (Cont.)

5. Heat Exchanger after Second Stage for Fuel Gas Heating before Expansion

Q = Load (Btu/h)	30,748,825	
Tha = Inlet temperature of hot fluid (°F)	480	
Thb = Outlet temperature of hot fluid (°F)	350	
Pressure of hot gases (psia)	993	
Tca = Inlet temperature of fuel gas (°F)	75	
Tcb = Outlet temperature of fuel gas (°F)	401	
ΔT_1	79	
ΔT_2	275	
Log mean temperature difference (°F)	157	
Overall heat transfer coefficient (Btu/h/ft ² /°F) ^d	5	
Heat transfer area (ft ²)	39,136	
Operating pressure (psia)	1,000	
Pressure factor	1.23	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	\$110,000	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$487,621

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

Q = Load (Btu/h)	132,075,366	
Tha = Inlet temperature of hot fluid (°F)	480	
Thb = Outlet temperature of hot fluid (°F)	350	
Pressure of hot gases (psia)	275	
Tca = Inlet temperature of fuel gas (°F)	191.6	
Tcb = Outlet temperature of fuel gas (°F)	401	
ΔT_1	79	
ΔT_2	158	
Log mean temperature difference (°F)	114	
Overall heat transfer coefficient (Btu/h/ft ² /°F) ^d	5	
Heat transfer area (ft ²)	231,440	
Operating pressure (psia)	300	
Pressure factor	1.165	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	\$110,000	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$461,853

TABLE 6.7 (Cont.)

7. Raw Synthesis Gas Compressor

Inlet pressure (psia)	250	
Outlet pressure (psia)	1,000	
Compressor size (hp)	17,413	
Purchased cost of 1,000-hp centrifugal compressor in 1987 (includes electric motor drive and gear reducer)	\$1,500,000	
Size factor for compressor	0.8	
Module factor ^e	2.6	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of compressor in 1993		\$6,845,283

8. Heat Exchanger after Raw Gas Compressor

Q = Load (Btu/h)	41,381,754	
Tha = Inlet temperature of hot fluid (°F)	414	
Thb = Outlet temperature of hot fluid (°F)	100	
Pressure of hot gases (psia)	1,000	
Tca = Inlet temperature of water (°F)	70	
Tcb = Outlet temperature of saturated water (°F)	200	
ΔT_1	214	
ΔT_2	30	
Log mean temperature difference (°F)	94	
Overall heat transfer coefficient (Btu/h/ft ² /°F)	40	
(hot fluid: gas, cold fluid: water) ^c		
Heat transfer area (ft ²)	11,056	
Operating pressure (psia)	1,000	
Pressure factor	1.23	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	\$114,391	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$507,088

9. Gas-Gas Heat Exchanger for Raw Gas Cooling

Q = Load (Btu/h)	4,177,943	
Tha = Inlet temperature of hot fluid (°F)	100	
Thb = Outlet temperature of hot fluid (°F)	68	
Pressure of hot gases (psia)	1,000	
Tca = Inlet temperature of cold fluid (°F)	30	
Tcb = Outlet temperature of gas (°F)	75	
ΔT_1	25	
ΔT_2	38	
Log mean temperature difference (°F)	31	
Overall heat transfer coefficient (Btu/h/ft ² /°F) ^d	5	
Heat transfer area (ft ²)	26,853	
Operating pressure (psia)	1,000	
Pressure factor	1.23	

TABLE 6.7 (Cont.)

9. (Cont.)		
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$161,695	
(mild steel construction: shell and tube floating head)		
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$716,782
10. First-Stage Shift Reactor		
Catalyst volume (ft ³)	1,000	
Reactor volume (ft ³) (1.2 times catalyst volume)	1,200	
Reactor volume (gal)	8,963	
Pressure factor	1	
Module factor	3.05	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Purchased cost of reactor in 1987	\$9,000	
Installed cost of reactor in 1993		\$30,916
11. Second-Stage Shift Reactor		
Catalyst volume (ft ³)	340	
Reactor volume (ft ³) (1.2 times catalyst volume)	408	
Reactor volume (gal)	3,041	
Pressure factor	1	
Module factor	3.05	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Purchased cost of reactor in 1987	\$5,000	
Installed cost of reactor in 1993		\$17,175
12. Cost of Shift Catalyst		
Volume of catalyst in first stage (ft ³)	1,000	
Volume of catalyst in second stage (ft ³)	340	
Cost of high-temperature catalyst (\$/ft ³)	\$50	
Cost of low-temperature catalyst (\$/ft ³)	\$250	
Total cost of catalyst		\$135,000
13. CO₂ Absorption Column		
Diameter of tower (ft)	11	
HETP (ft)	3	
No. of theoretical stages	10	
Absorber tower height (ft)	34	
(4 ft for inlet and outlet and for gas and liquid distributors)		
Volume of packing (ft ³)	2,852	
Pressure factor	4.2	
Cost per foot of column height (\$/ft)	\$1,300	
(mild steel construction)		
Module factor	4.16	

TABLE 6.7 (Cont.)

13. (Cont.)

CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
1993 installed cost of absorber		\$869,761
Cost of packing (\$/ft ³)	\$63.5	
(2-in. pall rings-metal)		
Total cost of packing		\$181,111

14. Power Recovery Turbine 1

Turbine size (kW)	567	
Purchased cost in 1979 ^f	\$180,000	
Module factor	1	
CE index for process equipment in 1979	256	
CE index for process equipment in 1993	360.4	
Installed cost of solvent pump in 1993		\$253,406

15. Power Recovery Turbine 2

Turbine size (kW)	293	
Purchased cost in 1979 ^f	\$140,000	
Module factor	1	
CE index for process equipment in 1979	256	
CE index for process equipment in 1993	360.4	
Installed cost of solvent pump in 1993		\$197,094

16. Slump Tank

Glycol solvent flow rate (lb/h)	6,440,000	
Density of glycol solvent (lb/gal)	8.6	
Residence time (s)	180	
Slump tank volume (gal)	37,442	
Pressure factor	1	
Module factor	2.08	
Purchased cost of slump tank in 1987 (mild steel construction)	\$70,000	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of slump tank in 1993		\$163,982

17. Recycle Compressor

Inlet pressure (psia)	750	
Outlet pressure (psia)	1,000	
Compressor size (hp)	151	
Purchased cost of 197-hp reciprocating compressor in 1987 (includes electric motor drive and gear reducer)	\$58,000	
Size factor for compressor	1	
Module factor ^e	2.6	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of compressor in 1993		\$169,839

TABLE 6.7 (Cont.)

18. Flash Tank 1			
Glycol solvent flow rate (lb/h)	6,440,000		
Density of glycol solvent (lb/gal)	8.6		
Residence time (s)	180		
Slump tank volume (gal)	37,442		
Pressure factor	1		
Module factor	2.08		
Purchased cost of slump tank in 1987 (mild steel construction)	\$70,000		
CE index for process equipment in 1987	320		
CE index for process equipment in 1993	360.4		
Installed cost of slump tank in 1993			\$163,982
19. Flash Tank 2			
Glycol solvent flow rate (lb/h)	6,440,000		
Density of glycol solvent (lb/gal)	8.6		
Residence time (s)	180		
Slump tank volume (gal)	37,442		
Pressure factor	1		
Module factor	2.08		
Purchased cost of slump tank in 1987 (mild steel construction)	\$70,000		
CE index for process equipment in 1987	320		
CE index for process equipment in 1993	360.4		
Installed cost of slump tank in 1993			\$163,982
20. Flash Tank 3			
Glycol solvent flow rate (lb/h)	6,440,000		
Density of glycol solvent (lb/gal)	8.6		
Residence time (s)	180		
Slump tank volume (gal)	37,442		
Pressure factor	1		
Module factor	2.08		
Purchased cost of slump tank in 1987 (mild steel construction)	\$70,000		
CE index for process equipment in 1987	320		
CE index for process equipment in 1993	360.4		
Installed cost of slump tank in 1993			\$163,982
21. Flash Gas Compressor 1			
Inlet pressure (psia)	20		
Outlet pressure (psia)	75		
Compressor size (hp)	255		
Purchased cost of centrifugal compressor in 1987 (includes electric motor drive and gear reducer)	\$85,000		
Size factor for compressor	1		
Module factor ^e	2.6		
CE index for process equipment in 1987	320		
CE index for process equipment in 1993	360.4		
Installed cost of compressor in 1993			\$248,901

TABLE 6.7 (Cont.)

22. Flash Gas Compressor 2

Inlet pressure (psia)	11	
Outlet pressure (psia)	75	
Compressor size (hp)	1,086	
Purchased cost of centrifugal compressor in 1987 (includes electric motor drive and gear reducer)	\$280,000	
Size factor for compressor	1	
Module factor ^e	2.6	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of compressor in 1993		\$819,910

23. Solvent Circulation Pump

Horsepower	2,388	
Size exponent	0.79	
Purchased cost of 300-hp pump in 1987 (includes motor, coupling, base; cast iron, horizontal)	\$30,000	
Module factor	1.5	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of solvent pump in 1993		\$260,988

24. CO₂ Product Gas Compressors

Compressor 1 (hp)	3,369	
Compressor 2 (hp)	3,217	
Compressor 3 (hp)	3,250	
Purchased cost of centrifugal compressor 1 in 1987	\$700,000	
Purchased cost of centrifugal compressor 2 in 1987	\$650,000	
Purchased cost of centrifugal compressor 3 in 1987 (includes electric motor drive and gear reducer)	\$650,000	
Size factor for compressor	1	
Module factor ^e	2.6	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of compressor 1 in 1993		\$2,049,775
Installed cost of compressor 2 in 1993		\$1,903,363
Installed cost of compressor 3 in 1993		\$1,903,363

TABLE 6.7 (Cont.)

25. Solvent Refrigeration		
Refrigeration (tons)	644	
Purchased cost in 1987	\$300,000	
Temperature correction factor	1.25	
Module factor	1.46	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of solvent pump in 1993		\$616,622
Total Cost		\$20,790,216
Total Cost of Four Trains		\$83,160,864

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

^b CE, Chemical engineering.

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

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

^e Source: Garrett (1989).

^f Source: Humphreys and Katell (1981, p. 134).

the fuel gas from 993 psia to a pressure of 250 psia reduces the temperature of the gases to 192°F. The gases are again heated and humidified by the heat available after the second shift. The remaining heat available is used in preheating the water or steam going to the HRSG and then is used in the steam cycle. The heat exchanger areas required for the first two operations are estimated, and the details are given in Table 6.7. The heat transfer to the steam cycle through the HRSG is assumed to take place in the existing HRSG; and, therefore, its design and cost are not estimated.

6.4.1.4 Raw Gas Compressor

In the high-pressure glycol solvent case, the gases are absorbed into the solvent at a pressure of 1,000 psia. The gases from the shift system are at a pressure of 250 psia. Therefore the raw gases are compressed to 1,000 psia by using a three-stage compressor. The compressor size for this duty is 17,413 hp. The horsepower of the compressor is estimated by using Equation 5.4.

6.4.1.5 Heat Exchanger after Raw Gas Compressor

The compression of gases from 250 to 1,000 psia increases the temperature of the gases to 414°F. Therefore the gases are cooled to 100°F by using water at 70°F. The overall heat transfer coefficient for this gas-liquid heat exchanger is taken as 40 Btu/h/ft²/°F. The heat transfer is estimated, and the details are given in Table 6.7.

6.4.1.6 Heat Exchanger for Raw Gas Cooling

The fuel gas from the top of the absorber is at a temperature of 30°F, and the inlet gases to the absorber are at a temperature of 100°F. Therefore the inlet synthesis gas can be cooled to a temperature of 68°F against the fuel gas, which can be heated to a temperature of 75°F. The overall heat transfer coefficient for this gas-gas heat exchanger is taken as 5 Btu/h/ft²/°F. The area required for this duty is calculated by using Equation 5.1, and the details are given in Table 6.7.

6.4.1.7 First-Stage Shift Reactor

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

6.4.1.8 Second-Stage Shift Reactor

The volume of the reactor is assumed to be 20% greater than the volume of the catalyst. The volume of the catalyst is estimated to be 340 ft³ from the kinetics and the equilibrium considerations.

6.4.1.9 Cost of Catalyst

The cost of the iron-based catalyst is assumed to be \$50/ft³ and that of the copper-based catalyst \$250/ft³.

6.4.1.10 Absorption Tower

The diameter of the absorption tower is calculated for 60% flooding velocity. Diameters are calculated at the top and bottom of the tower, and the larger of the two is taken as the diameter of the tower. The height of the tower is calculated from the number of theoretical stages required

and the HETP. The assumption is made that 10 theoretical stages are required and that HETP is 3 ft. An additional 4 ft is provided for liquid and gas distributors. The diameter is 11 ft, and the total height is 34 ft.

6.4.1.11 Power Recovery Turbine 1

The rich glycol solvent is at a pressure of 1,000 psia. The gases absorbed in the glycol solvent can be released by lowering the pressure. In the first stage, a pressure of 750 psia is selected. At this pressure, not much CO₂ is desorbed from the glycol solvent; and the other less soluble gases, such as H₂, CH₄, and CO, are recycled to the absorption tower after compressing to the tower operating pressure. The turbine efficiency is assumed to be 72%. The power recovered from the liquid is calculated from the expression

$$\text{hydraulic hp} = H_s \cdot \frac{G}{1,714} \cdot \eta \quad , \quad (6.1)$$

where

H_s = total dynamic head (lb/in.²),

G = flow rate of liquid (gal/min), and

η = efficiency of the turbine.

6.4.1.12 Power Recovery Turbine 2

The pressure in the second stage is reduced from 750 psia to 75 psia. At this pressure, large quantities of CO₂ are desorbed from the glycol solvent. Again, the power recovered is estimated as flow work, as given previously.

6.4.1.13 Slump Tank

Sufficient residence time is required for the gases to escape from the liquid; otherwise, these gases will be carried over along with the glycol solvent and will be released along with CO₂ in subsequent stages. A residence time of 3 minutes is selected on the basis of the literature (Woelfer et al. 1977):

$$\text{volume of slump tank (gal)} = \text{flow rate (gal/min)} \cdot \text{residence time (min)} \quad . \quad (6.2)$$

6.4.1.14 Recycle Gas Compressor

The less soluble gases released from the slump tank are recycled to the tower by compressing them to the tower operating pressure. The gases are compressed from 750 psia to a pressure of 1,000 psia. The horsepower of the compressor is estimated by using Equation 5.4.

6.4.1.15 Flash Tank 1

The CO₂ is released from the glycol solvent by reducing the pressure. This reduction is carried out in stages. The first-stage flash tank is at a pressure of 75 psia. The flash tanks are designed to provide sufficient residence time for the gases to escape from the glycol solvent. A residence time of 3 minutes is selected.

6.4.1.16 Flash Tank 2

The CO₂ is released from the glycol solvent by reducing the pressure. This reduction is carried out in stages. The second-stage flash tank is at a pressure of 20 psia. The flash tanks are designed to provide sufficient residence time for the gases to escape from the glycol solvent. A residence time of 3 minutes is selected.

6.4.1.17 Flash Tank 3

The CO₂ is released from the glycol solvent by reducing the pressure. This reduction is carried out in stages. The third-stage flash tank is at a pressure of 11 psia. The flash tanks are designed to provide sufficient residence time for the gases to escape from the glycol solvent. A residence time of 3 minutes is selected.

6.4.1.18 Flash Gas Compressor 1

The CO₂ released at 20 psia is compressed to a pressure of 75 psia. The compressed CO₂ is combined with the CO₂ from the 75-psia flash tank. The horsepower of the compressor is estimated by using Equation 5.4.

6.4.1.19 Flash Gas Compressor 2

The CO₂ released at 11 psia is compressed to a pressure of 75 psia. The compressed CO₂ is combined with the CO₂ from the 75-psia flash tank. The horsepower of the compressor is estimated by using Equation 5.4.

6.4.1.20 Solvent Circulation Pump

The lean glycol solvent is pumped back to the tower operating pressure by using a circulation pump. The horsepower required is calculated from the work of pressurization (flow work).

6.4.1.21 Carbon Dioxide Product Gas Compressors

The CO₂ from the flash tanks is at a pressure of 75 psia. These gases have to be compressed to the pipeline pressure of 1,000 psia. This compression is done in three stages, with interstage cooling.

6.4.1.22 Solvent Refrigeration

The temperature of the glycol solvent increases because of the heat of absorption and the heat of pumping. The glycol solvent is again cooled to the tower operating temperature by using refrigeration. The amount of refrigeration is calculated by the difference in enthalpy of the solvent:

$$\text{refrigeration power (kW)} = \frac{\text{refrigeration load (Btu/h)}}{EER \cdot 1,000} \quad (6.3)$$

The energy efficiency ratio (*EER*) is defined as

$$EER = 9.0 + \frac{T_{\text{evap}}}{10} \quad (6.4)$$

6.4.2 Low-Pressure Glycol Solvent Case

6.4.2.1 Heat Exchanger before First-Stage Shift Reactor

The temperature of the gases from the KRW process is 1,000°F, and these gases have to be cooled to a temperature of 482°F. This temperature is found to be optimum for the first-stage shift reactor. The difference between the enthalpies of gases at these two temperatures gives the heat exchanger load. The majority of the heat available is used for raising saturated steam at 300 psia from saturated water. This saturated steam has to be injected into the first-stage shift reactor. During this process, the gases are cooled to a temperature of 531°F. Further cooling to the temperature of 482°F is achieved by heating and humidifying the fuel gas. The heat transfer area required for these two operations is estimated by using Equation 5.1, and the detailed calculations are given in Table 6.8.

6.4.2.2 Heat Exchanger between First and Second Shift Reactors

The shift reaction is an exothermic reaction, and the heat released in the first stage has to be removed to obtain an optimum temperature for the second-stage shift reactor. The gases have to be cooled from 680°F to 460°F. The heat load of the heat exchanger is again the difference between the enthalpies of the gases. Most of the available heat is used in heating the water from 100°F to the saturation temperature of 417.4°F at 300 psia. During this process, the gases are cooled to a temperature of 537°F. Further cooling to the required temperature of 460°F is achieved by heating and humidifying the fuel gas. The heat transfer area is estimated by using Equation 5.1, and the installed cost of the heat exchanger is given in Table 6.8.

6.4.2.3 Heat Exchanger after Second-Stage Shift Reactor

The shifted gases from the second stage are at a temperature of 480°F. These gases have to be cooled to a temperature of 100°F. The cooling is achieved in stages. First, the fuel gas is humidified and heated from 75°F to a temperature of 401°F. The remaining heat available is used in preheating the water or steam going to the HRSG and then is used in the steam cycle. The heat exchanger area required for the first operation is estimated, and the details are given in Table 6.8. The heat transfer to the steam cycle through the HRSG is assumed to take place in the existing HRSG; and, therefore, its design and cost are not estimated.

TABLE 6.8 Sizing and Cost Estimation of Major Equipment for Low-Pressure Glycol Solvent Process (Selexol)

1. Heat Exchanger before First Stage for Raising Steam

Q = Load (Btu/h)	78,838,137	
Tha = Inlet temperature of hot fluid (°F)	1,000	
Thb = Outlet temperature of hot fluid (°F)	531	
Pressure of hot gases (psia)	295	
Tca = Inlet temperature of saturated water (°F)	417	
Tcb = Outlet temperature of saturated steam (°F)	417	
ΔT_1	583	
ΔT_2	114	
Log mean temperature difference (°F)	287	
Overall heat transfer coefficient (Btu/h/ft ² /°F) ^a	260	
Heat transfer area (ft ²)	1,057	
Operating pressure (psia)	300	
Pressure factor	1.165	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$23,000	
(mild steel construction; shell and tube floating head)		
CE index for process equipment in 1987 ^b	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$96,569

2. Heat Exchanger before First Stage for Heating Fuel Gas

Q = Load (Btu/h)	8,219,649	
Tha = Inlet temperature of hot fluid (°F)	531	
Thb = Outlet temperature of hot fluid (°F)	482	
Pressure of hot gases (psia)	295	
Tca = Inlet fuel gas/water temperature (°F)	100	
Tcb = Outlet fuel gas/water temperature (°F)	510	
ΔT_1	21	
ΔT_2	382	
Log mean temperature difference (°F)	124	
Overall heat transfer coefficient (Btu/h/ft ² /°F) ^c	5	
Heat transfer area (ft ²)	13,210	
Operating pressure (psia)	300	
Pressure factor	1.165	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$122,616	
(mild steel construction; shell and tube floating head)		
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$514,823

TABLE 6.8 (Cont.)

3. Heat Exchanger after First Stage for Saturated Water

Q = Load (Btu/h)	31,747,489	
Tha = Inlet temperature of hot fluid (°F)	680	
Thb = Outlet temperature of hot fluid (°F)	537	
Pressure of hot gases (psia)	285	
Tca = Inlet temperature of cold fluid (fuel gas) (°F)	100	
Tcb = Outlet temperature of cold fluid (fuel gas) (°F)	417	
ΔT_1	263	
ΔT_2	437	
Log mean temperature difference (°F)	342	
Overall heat transfer coefficient (Btu/h/ft ² /°F)	40	
(hot fluid: gas, cold fluid: water) ^d		
Heat transfer area (ft ²)	2,319	
Operating pressure (psia)	285	
Pressure factor	1.165	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$40,000	
(mild steel construction; shell and tube floating head)		
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$167,946

4. Gas-Gas Heat Exchanger after First Stage for Heating Fuel Gas

Q = Load (Btu/h)	17,195,044	
Tha = Inlet temperature of hot fluid (°F)	537	
Thb = Outlet temperature of hot fluid (°F)	460	
Pressure of hot gases (psia)	285	
Tca = Inlet temperature of fuel gas/water (°F)	100	
Tcb = Outlet temperature of fuel gas/water (°F)	510	
ΔT_1	27	
ΔT_2	360	
Log mean temperature difference (°F)	129	
Overall heat transfer coefficient (Btu/h/ft ² /°F) ^c	5	
Heat transfer area (ft ²)	26,751	
Operating pressure (psia)	285	
Pressure factor	1.165	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$161,455	
(mild steel construction; shell and tube floating head)		
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$677,897

TABLE 6.8 (Cont.)

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

Q = Load (Btu/h)	142,954,526	
Tha = Inlet temperature of hot fluid (°F)	480	
Thb = Outlet temperature of hot fluid (°F)	120	
Pressure of hot gases (psia)	275	
Tca = Inlet fuel gas/water temperature (°F)	75	
Tcb = Outlet fuel gas/water temperature (°F)	401	
ΔT_1	79	
ΔT_2	45	
Log mean temperature difference (°F)	60	
Overall heat transfer coefficient (Btu/h/ft ² /°F)	5	
(hot fluid: gas, cold fluid: water) ^e		
Heat transfer area (ft ²)	473,251	
Operating pressure (psia)	275	
Pressure factor	1.165	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$495,084	
(mild steel construction; shell and tube floating head)		
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$2,078,687

6. Gas-Gas Heat Exchanger for Raw Gas Cooling

Q = Load (Btu/h)	4,183,448	
Tha = Inlet temperature of hot fluid (°F)	100	
Thb = Outlet temperature of hot fluid (°F)	68	
Pressure of hot gases (psia)	250	
Tca = Inlet temperature of cold fluid (°F)	30	
Tcb = Outlet temperature of steam (°F)	75	
ΔT_1	25	
ΔT_2	38	
Log mean temperature difference (°F)	31	
Overall heat transfer coefficient (Btu/h/ft ² /°F) ^c	5	
Heat transfer area (ft ²)	26,903	
Operating pressure (psia)	275	
Pressure factor	1.165	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$161,814	
(mild steel construction; shell and tube floating head)		
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$679,401

TABLE 6.8 (Cont.)

7. First-Stage Shift Reactor

Catalyst volume (ft ³)	1,000	
Reactor volume (ft ³) (1.2 times catalyst volume)	1,200	
Reactor volume (gal)	8,963	
Pressure factor	1	
Module factor	3.05	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Purchased cost of reactor in 1987	\$9,000	
Installed cost of reactor in 1993		\$30,916

8. Second-Stage Shift Reactor

Catalyst volume (ft ³)	340	
Reactor volume (ft ³) (1.2 times catalyst volume)	408	
Reactor volume (gal)	3,041	
Pressure factor	1	
Module factor	3.05	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Purchased cost of reactor in 1987	\$5,000	
Installed cost of reactor in 1993		\$17,175

9. Cost of Shift Catalyst

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

10. CO₂ Absorption Column

Diameter of tower (ft)	17	
HETP (ft)	3	
No. of theoretical stages	12	
Absorber tower height (ft)	40	
(4 ft for inlet and outlet and for gas and liquid distributors)		
Volume of packing (ft ³)	8,175	
Pressure factor	2	
Cost per foot of column height (\$/ft)	\$1,800	
(mild steel construction)		
Module factor	416	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
1993 installed cost of absorber		\$674,669
Cost of packing (\$/ft ³)	\$63.5	
(2-in. pall rings-metal)		
Total cost of packing		\$519,085

TABLE 6.8 (Cont.)

11. Power Recovery Turbine 1

Turbine size (kW)	567	
Purchased cost in 1979 ^f	\$180,000	
Module factor	1	
CE index for process equipment in 1979	256	
CE index for process equipment in 1993	360.4	
Installed cost of solvent pump in 1993		\$253,406

12. Power Recovery Turbine 2

Turbine size (kW)	293	
Purchased cost in 1979 ^f	\$140,000	
Module factor	1	
CE index for process equipment in 1979	256	
CE index for process equipment in 1993	360.4	
Installed cost of solvent pump in 1993		\$197,094

13. Slump Tank

Glycol solvent flow rate (lb/h)	6,440,000	
Density of glycol solvent (lb/gal)	8.6	
Residence time (s)	180	
Slump tank volume (gal)	37,442	
Pressure factor	1	
Module factor	2.08	
Purchased cost of slump tank in 1987 (mild steel construction)	\$70,000	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of slump tank in 1993		\$163,982

14. Recycle Compressor

Inlet pressure (psia)	100	
Outlet pressure (psia)	250	
Compressor size (hp)	151.30	
Purchased cost of 197-hp reciprocating compressor in 1987 (includes electric motor drive and gear reducer)	\$58,000	
Size factor for compressor	1	
Module factor ^g	2.6	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of compressor in 1993		\$169,839

15. Flash Tank 1

Glycol solvent flow rate (lb/h)	6,440,000	
Density of glycol solvent (lb/gal)	8.6	
Residence time (s)	180	
Slump tank volume (gal)	37,442	
Pressure factor	1	
Module factor	2.08	

TABLE 6.8 (Cont.)

15. (Cont.)

Purchased cost of slump tank in 1987 (mild steel construction)	\$70,000	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of slump tank in 1993		\$163,982

16. Flash Tank 2

Glycol solvent flow rate (lb/h)	6,440,000	
Density of glycol solvent (lb/gal)	8.6	
Residence time (s)	180	
Slump tank volume (gal)	37,442	
Pressure factor	1	
Module factor	2.08	
Purchased cost of slump tank in 1987 (mild steel construction)	\$70,000	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of slump tank in 1993		\$163,982

17. Flash Tank 3

Glycol solvent flow rate (lb/h)	6,440,000	
Density of glycol solvent (lb/gal)	8.6	
Residence time (s)	180	
Slump tank volume (gal)	37,442	
Pressure factor	1	
Module factor	2.08	
Purchased cost of slump tank in 1987 (mild steel construction)	\$70,000	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of slump tank in 1993		\$163,982

18. Flash Gas Compressor 1

Inlet pressure (psia)	14.7	
Outlet pressure (psia)	25	
Compressor size (hp)	255	
Purchased cost of centrifugal compressor in 1987 (includes electric motor drive and gear reducer)	\$85,000	
Size factor for compressor	1	
Module factor ⁹	2.6	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of compressor in 1993		\$248,901

TABLE 6.8 (Cont.)

19. Flash Gas Compressor 2

Inlet pressure (psia)	4	
Outlet pressure (psia)	25	
Compressor size (hp)	1,086	
Purchased cost of centrifugal compressor in 1987 (includes electric motor drive and gear reducer)	\$280,000	
Size factor for compressor	1	
Module factor ⁹	2.6	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of compressor in 1993		\$819,910

20. Solvent Circulation Pump

Horsepower	2,388	
Size exponent	0.79	
Purchased cost of 300-hp pump in 1987 (includes motor, coupling, base; cast iron, horizontal)	\$30,000	
Module factor	1.5	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of solvent pump in 1993		\$260,988

21. CO₂ Product Gas Compressors

Compressor 1 (hp)	3,368.96	
Compressor 2 (hp)	3,217.25	
Compressor 3 (hp)	3,250.33	
Purchased cost of centrifugal compressor 1 in 1987	\$700,000	
Purchased cost of centrifugal compressor 2 in 1987	\$650,000	
Purchased cost of centrifugal compressor 3 in 1987 (includes electric motor drive and gear reducer)	\$650,000	
Size factor for compressor	1	
Module factor ⁹	2.6	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of compressor 1 in 1993		\$2,049,775
Installed cost of compressor 2 in 1993		\$1,903,363
Installed cost of compressor 3 in 1993		\$1,903,363

TABLE 6.8 (Cont.)

22. Solvent Refrigeration		
Refrigeration (tons)	644.01	
Purchased cost in 1987	\$300,000	
Temperature correction factor	1.25	
Module factor	1.46	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of solvent pump in 1993		\$ 616,622
Total Cost		\$14,671,357
Total Cost of Four Trains		\$58,685,428

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

^b CE, Chemical engineering.

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

^d Source: Perry and Chilton (1973, p. 10-14).

^e Source: Peters and Timmerhaus (1980, p. 651).

^f Source: Humphreys and Katell (1981, p. 134).

^g Source: Garrett (1989).

6.4.2.4 Heat Exchanger for Raw Gas Cooling

The fuel gas from the top of the absorber is at a temperature of 30°F, and the inlet gases to the absorber are at a temperature of 100°F. Therefore the inlet synthesis gas can be cooled to a temperature of 68°F against the fuel gas, which can be heated to a temperature of 75°F. The overall heat transfer coefficient for this gas-gas heat exchanger is taken as 5 Btu/h/ft²/°F. The area required for this duty is calculated by using Equation 5.1, and the details are given in Table 6.8.

6.4.2.5 First-Stage Shift Reactor

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

6.4.2.6 Second-Stage Shift Reactor

The volume of the reactor is assumed to be 20% greater than the volume of the catalyst. The volume of the catalyst is estimated to be 340 ft³ from the kinetics and the equilibrium considerations.

6.4.2.7 Cost of Catalyst

The cost of the iron-based catalyst is assumed to be \$50/ft³ and that of the copper-based catalyst \$250/ft³.

6.4.2.8 Absorption Tower

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

6.4.2.9 Power Recovery Turbine 1

The rich glycol solvent is at a pressure of 250 psia. The gases absorbed in the glycol solvent can be released by lowering the pressure. In the first stage, a pressure of 100 psia is selected. At this pressure, not much CO₂ is desorbed from the glycol solvent; and the other less soluble gases, such as H₂, CH₄, and CO, are recycled to the absorption tower after compressing to the tower operating pressure. The turbine efficiency is assumed to be 72%. The power recovered from the liquid is calculated from Equation 6.1.

6.4.2.10 Power Recovery Turbine 2

The pressure in the second stage is reduced from 100 psia to 25 psia. At this pressure, large quantities of CO₂ are desorbed from the glycol solvent. Again, the power recovered is estimated as flow work, as given previously.

6.4.2.11 Slump Tank

Sufficient residence time is required for the gases to escape from the liquid; otherwise, these gases will be carried over along with the glycol solvent and will be released along with CO₂ in subsequent stages. A residence time of 3 minutes is selected on the basis of the literature (Woelfer et al. 1977).

6.4.2.12 Recycle Gas Compressor

The less soluble gases released from the slump tank are recycled to the tower by compressing them to the tower operating pressure. The gases are compressed from 100 psia to a pressure of 250 psia. The horsepower of the compressor is estimated from Equation 5.4.

6.4.2.13 Flash Tank 1

The CO₂ is released from the glycol solvent by reducing the pressure. This reduction is carried out in stages. The first-stage flash tank is at a pressure of 25 psia. The flash tanks are designed to provide sufficient residence time for the gases to escape from the glycol solvent. A residence time of 3 minutes is selected.

6.4.2.14 Flash Tank 2

The CO₂ is released from the glycol solvent by reducing the pressure. This reduction is carried out in stages. The second-stage flash tank is at a pressure of 14.7 psia. The flash tanks are designed to provide sufficient residence time for the gases to escape from the glycol solvent. A residence time of 3 minutes is selected.

6.4.2.15 Flash Tank 3

The CO₂ is released from the glycol solvent by reducing the pressure. This reduction is carried out in stages. The third-stage flash tank is at a pressure of 4 psia. The flash tanks are designed to provide sufficient residence time for the gases to escape from the glycol solvent. A residence time of 3 minutes is selected.

6.4.2.16 Flash Gas Compressor 1

The CO₂ released at 14.7 psia is compressed to a pressure of 25 psia. The compressed CO₂ is combined with the CO₂ from the 25-psia flash tank. The horsepower of the compressor is estimated by using Equation 5.4.

6.4.2.17 Flash Gas Compressor 2

The CO₂ released at 4 psia is compressed to a pressure of 25 psia. The compressed CO₂ is combined with the CO₂ from the 25-psia flash tank. The horsepower of the compressor is estimated by using Equation 5.4.

6.4.2.18 Solvent Circulation Pump

The lean glycol solvent is pumped back to the tower operating pressure by using a circulation pump. The horsepower required is calculated from the work of pressurization (flow work).

6.4.2.19 Carbon Dioxide Product Gas Compressors

The CO₂ from the flash tanks is at pressure of 25 psia. These gases have to be compressed to the pipeline pressure of 1,000 psia. This compression is done in three stages, with interstage cooling.

6.4.2.20 Solvent Refrigeration

The temperature of the glycol solvent increases because of the heat of absorption and the heat of pumping. The glycol solvent is again cooled to the tower operating temperature by using refrigeration. The amount of refrigeration is calculated by the difference in enthalpy of the solvent.

7 Integrated Gasification Combined Cycle with Chilled Methanol CO₂ Recovery

7.1 Feed Gas Preparation and Process Integration

As discussed in Section 4.3, the most significant issue in process integration is the effective use of process heat from cooling of the synthesis gas as required for the shift reaction and gas preparation for the CO₂ absorber. The most effective use of this thermal energy is preheating of the fuel gas feed to the gas turbine. With the addition of moisture to that feed, it can absorb most of the available heat. The balance is used for feedwater heating, raising steam for the steam cycle, and solvent regeneration.

Because of differences in the absorber inlet gas specifications among the alternative CO₂ recovery systems, some variation exists in the amount of heat available. For the chilled methanol process, the CO₂-rich gas to the absorber must be cooled to about -15°F. This temperature is substantially cooler than the absorber inlet temperature required for the other recovery systems. This temperature is attained by cooling the incoming shifted gas stream, stream 1A (Figure 7.1), with the fuel gas leaving the absorber. This lean fuel gas is at -70°F because of contact with refrigerated solvent. Thus, this final cooling is accomplished internal to the recovery process and does not necessarily affect the thermal energy balance for gas preparation as it has been outlined in Section 4.3. The processed gas, therefore, effectively enters the recovery process at 100°F, and the final cooling is accomplished by the heat exchange with lean fuel gas and, subsequently, by direct contact with solvent in the absorber.

The amount and allocation of thermal energy available from initial gas cooling, various processes in the shift reaction, and cooling to 100°F are listed in Table 7.1. A portion of the heat available after the second-stage shift is allocated to solvent regeneration. The heating requirement for solvent regeneration in this case is substantially lower than that required for MEA or hot potassium carbonate (K₂CO₃).

7.2 Carbon Dioxide Recovery by Chilled Methanol

An overview of the methanol-based recovery system is provided by Figure 7.1 (the system flow diagram) and by Table 7.2 (the corresponding line lists). A brief overview of the system is provided subsequently. More detailed information and assumptions for each stream are provided in Table 7.3.

The chilled methanol process uses nearly pure methanol as the solvent. During the process, moisture is absorbed by the methanol and must be removed by fractional distillation to recover the solvent. The distillation results in discarding some methanol with the separated water. Compensation for this loss requires a methanol makeup stream.

TABLE 7.1 Heat Recovery and Allocation for Chilled Methanol Process

Process	Enthalpy Change Available from Process (10 ⁶ Btu)	Allocation to Fuel Gas Preheating (10 ⁶ Btu)	Allocation to Steam Cycle (10 ⁶ Btu)	Allocation to Solvent Regeneration (10 ⁶ Btu)
Initial synthesis gas cooling to 1,000°F	1,299.3	328.5	970.8	0
Cooling before first-stage shift	32.9	32.9	0	0
Cooling after first-stage shift	68.8	68.8	0	0
Cooling after second-stage shift	847.0	350.2	0	496.8

The low absorber operating temperature is dictated by the solubility data curves for CO₂ in methanol. At the low absorber temperature, moisture in the feed gas to the absorber would freeze and possibly interfere with process operation. The methanol solvent is introduced to the gas stream ahead of the absorber in an amount sufficient to act as an antifreeze agent for the incoming moisture, which is absorbed by the methanol. A phase separator removes the resulting methanol-water liquid mixture, and the methanol is recovered by distillation. This methanol stream is only a small fraction of the total solvent.

Solvent regeneration is accomplished through pressure reduction and heating. Pressure reduction is done in a series of flash tanks, which provide sufficient residence time at each pressure to release CO₂. The first flash drum in the series is used to recover combustible gases, which are recycled to the absorber. The remaining flash drums recover CO₂. Partial recovery of CO₂ in this manner reduces the compressor work required for the CO₂ product.

Final desorption of the CO₂ is accomplished in the stripping column through heating with steam. A portion of the heat required is obtained through exchange of heat between the lean solvent (stream 6B) and the rich solvent (stream 5E). Most of the heat of regeneration is obtained from steam supplied to the stripper. The sources of steam or heat for steam supply are listed in Table 7.1. The lean solvent requires further cooling, which is accomplished by refrigeration. The cooled solvent enters the absorber at -70°F. The regeneration process releases the CO₂, which is cooled to condense accompanying solvent vapors and to prepare the CO₂ for compression to pipeline disposal conditions.

TABLE 7.2 Stream Flows of Chilled Methanol Process^a

Stream Data	Stream 1A	Stream 1B	Stream 1C	Stream 1D	Stream 2A	Stream 2B
Description of stream	Feed gas from shift system	Absorber feed	Feed gas phase separator	Feed gas to absorber	Fuel gas from absorber	Fuel gas
Gases (lb-mol/h)						
CO	77.37	77.37	77.37	77.37	75.70	75.70
CO ₂	4,335.99	4,336.95	4,336.95	4,336.95	159.90	159.90
H ₂	5,611.86	5,611.86	5,611.86	5,611.86	5,607.20	5,607.20
H ₂ O	61.91	61.91	61.91	0.00	0.00	0.00
N ₂	7,306.65	7,306.65	7,306.65	7,306.65	7,271.50	7,271.50
Ar	88.60	88.60	88.60	88.60	87.40	87.40
CH ₄	128.77	128.77	128.77	128.77	123.10	123.10
NH ₃	2.99	2.99	2.99	2.99	0.00	0.00
H ₂ S	0.40	0.40	0.40	0.40	0.00	0.00
HCl	0.04	0.04	0.04	0.04	0.04	0.04
O ₂	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00
SO ₂	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	17,614.58	17,615.54	17,615.54	17,553.63	13,324.84	13,324.84
Liquids (lb-mol/h)						
Methanol	0.00	48.06	48.06	0.00	0.11	0.11
Temperature (°F)	100.00	99.37	-6.27	-6.27	-70.00	80.00
Pressure (psia)	250.00	250.00	245.00	240.00	235.00	235.00
Enthalpy of stream (Btu/h) (reference, 32°F)	9.96×10^6	9.92×10^6	-5.01×10^6	-4.93×10^6	-9.35×10^6	4.43×10^6

TABLE 7.2 Stream Flows of Chilled Methanol Process^a (Cont.)

Stream Data	Stream 3A	Stream 3B	Stream 3C	Stream 3D	Stream 4A	Stream 4B
Description of stream	Methanol-water mixture from phase separator	Methanol-water to distillation column	Methanol to stripper	Wastewater from distillation column	CO ₂ -rich gas from flash drum 2	CO ₂ -rich gas from heat exchanger
Gases (lb-mol/h)						
CO	0.00	0.00	0.00	0.00	0.74	0.74
CO ₂	0.00	0.00	0.00	0.00	1,044.02	1,044.02
H ₂	0.00	0.00	0.00	0.00	5.59	5.59
H ₂ O	61.91	61.91	2.53	59.38	0.00	0.00
N ₂	0.00	0.00	0.00	0.00	34.90	34.90
Ar	0.00	0.00	0.00	0.00	1.20	1.20
CH ₄	0.00	0.00	0.00	0.00	5.67	5.67
NH ₃	0.00	0.00	0.00	0.00	0.00	0.00
H ₂ S	0.00	0.00	0.00	0.00	0.00	0.00
HCl	0.00	0.00	0.00	0.00	0.00	0.00
O ₂	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00
SO ₂	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	61.91	61.91	2.53	59.38	1,092.12	1,092.12
Liquids (lb-mol/h)						
Methanol	48.06	48.06	48.06	0.01	0.00	0.00
Temperature (°F)	-6.27	-6.27	150.00	280.00	-36.44	84.50
Pressure (psia)	240.00	50.00	50.00	50.00	100.00	95.00
Enthalpy of stream (Btu/h) (reference, 32°F)	-7.60 × 10 ⁴	-7.60 × 10 ⁴	8.61 × 10 ⁵	2.65 × 10 ⁵	-6.44 × 10 ⁵	5.10 × 10 ⁵

TABLE 7.2 Stream Flows of Chilled Methanol Process^a (Cont.)

Stream Data	Stream 4C	Stream 4D	Stream 4E	Stream 4F	Stream 4G	Stream 4H
Description of stream	CO ₂ -rich gas from stripper	CO ₂ -rich gas from phase separator	Compressed CO ₂ -rich gas	CO ₂ -rich gas from flash drum 3	Compressed CO ₂ -rich gas	Final CO ₂ product
Gases (lb-mol/h)						
CO	0.00	0.00	0.00	0.93	0.93	1.67
CO ₂	1,437.37	1,437.37	1,437.37	1,694.69	1,694.69	4,176.09
H ₂	0.00	0.00	0.00	-0.93	-0.93	4.66
H ₂ O	2.53	2.53	2.53	0.00	0.00	2.53
N ₂	0.00	0.00	0.00	0.25	0.25	35.15
Ar	0.00	0.00	0.00	0.00	0.00	1.20
CH ₄	0.00	0.00	0.00	0.00	0.00	5.67
NH ₃	0.00	0.00	0.00	2.99	2.99	2.99
H ₂ S	0.40	0.40	0.40	0.00	0.00	0.40
HCl	0.00	0.00	0.00	0.00	0.00	0.00
O ₂	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00
SO ₂	0.00	0.00	0.00	0.00	-0.93	0.00
Total gas flow	1,440.30	1,440.30	1,440.30	1,697.94	1,697.01	4,230.36
Liquids (lb-mol/h)						
Methanol	5,394.33	514.21	514.21	0.00	0.00	514.21
Temperature (°F)	135.00	100.00	479.86	-49.62	167.22	273.28
Pressure (psia)	14.70	14.70	95.00	20.00	95.00	95.00
Enthalpy of stream (Btu/h) (reference, 32°F)	9.26 × 10 ⁷	9.52 × 10 ⁶	1.60 × 10 ⁷	-1.20 × 10 ⁶	2.10 × 10 ⁶	1.86 × 10 ⁷

TABLE 7.2 Stream Flows of Chilled Methanol Process^a (Cont.)

Stream Data	Stream 4I	Stream 5A	Stream 5B	Stream 5C	Stream 5D	Stream 5E
Description of stream	Recycle to stripper	CO ₂ -rich solvent from absorber	Recycle to absorber	CO ₂ -rich solvent from flash drum 1	CO ₂ -rich solvent from flash drum 2	CO ₂ -rich solvent from flash drum 3
Gases (lb-mol/h)						
CO	0.00	16.69	15.02	1.67	0.93	0.00
CO ₂	0.00	5,088.49	0.00	5,088.49	4,044.47	2,349.77
H ₂	0.00	46.63	41.97	4.66	-0.93	0.00
H ₂ O	0.00	0.00	0.00	0.00	0.00	0.00
N ₂	0.00	351.52	316.37	35.15	0.25	0.00
Ar	0.00	12.00	10.80	1.20	0.00	0.00
CH ₄	0.00	56.66	50.99	5.67	0.00	0.00
NH ₃	0.00	2.99	0.00	2.99	2.99	0.00
H ₂ S	0.00	0.40	0.00	0.40	0.40	0.40
HCl	0.00	0.00	0.00	0.00	0.00	0.00
O ₂	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00
SO ₂	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	0.00	5,575.37	435.15	5,140.23	4,048.11	2,350.17
Liquids (lb-mol/h)						
Methanol	4,880.13	45,571.78	0.00	45,571.78	45,571.78	45,571.78
Temperature (°F)	100.00	-28.41	-28.41	-28.41	-36.44	-49.62
Pressure (psia)	95.00	235.00	200.00	200.00	100.00	20.00
Enthalpy of stream (Btu/h) (reference, 32°F)	6.01×10^6	-5.27×10^7	-1.82×10^5	-5.26×10^7	-5.89×10^7	-6.90×10^7

TABLE 7.2 Stream Flows of Chilled Methanol Process^a (Cont.)

Stream Data	Stream 5F	Stream 6A	Stream 6B	Stream 6C	Stream 6D	Stream 6E	Stream 7
Description of stream	CO ₂ -rich solvent to stripper	CO ₂ -lean solvent from stripper	Lean solvent after circulation pump	Lean solvent to absorber	Lean solvent to feed gas injection	Lean solvent to absorber	Methanol makeup
Gases (lb-mol/h)							
CO	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO ₂	2,349.77	912.40	912.40	911.44	0.96	911.44	0.00
H ₂	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H ₂ O	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N ₂	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ar	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH ₄	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH ₃	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H ₂ S	0.40	0.00	0.00	0.00	0.00	0.00	0.00
HCl	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O ₂	0.00	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO ₂	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	2,350.17	912.40	912.40	911.44	0.96	911.44	0.00
Liquids (lb-mol/h)							
Methanol	45,571.78	45,619.95	45,619.95	45,571.89	48.06	45,571.89	514.32
Temperature (°F)	114.20	149.00	151.12	-15.00	-15.00	-70.00	70.00
Pressure (psia)	14.70	14.70	250.00	250.00	250.00	250.00	14.70
Enthalpy of stream (Btu/h) (reference, 32°F)	6.96 × 10 ⁷	9.76 × 10 ⁷	9.94 × 10 ⁷	-3.92 × 10 ⁷	-4.13 × 10 ⁴	-8.50 × 10 ⁷	3.54 × 10 ⁵

^a Flow rates are for one of four trains.

TABLE 7.3 Descriptions of Streams of Chilled Methanol Process^a

Stream and Characteristic	Data	Comments on Stream Calculations
Stream 1A: Gas feed from shift system	100	This stream is coming from shift system. This stream will be cooled against cold fuel gas from absorber and cold CO ₂ -rich gas from flash drum 2.
Temperature (°F)	250	
Pressure (psia)	17,615	
Flow rate (lb-mol/h)	0.2461	
CO ₂ (mole fraction)		
Stream 2A: Fuel gas from top of absorber		Chilled methanol enters top of column at temperature of -70°F. Therefore, gases leaving column are in equilibrium with methanol; and, hence, they are at temperature of -70°F. Gas composition corresponds to 95% removal of CO ₂ .
Temperature (°F)	-70	
Pressure (psia)	235	
Flow rate (lb-mol/h)	13,325	
CO ₂ (mole fraction)	0.0120	
Stream 3A: Methanol-water mixture from phase separator		Methanol is added to feed gas prior to absorption column to prevent icing of water in feed gas. Condensed water and methanol are separated from gas in phase separator.
Temperature (°F)	-6	
Pressure (psia)	240	
Flow rate (lb-mol/h)	110	
CO ₂ (mole fraction)	0	
Stream 3B: Methanol-water mixture to distillation column		Methanol is separated from methanol-water mixture in distillation column.
Temperature (°F)	-6	
Pressure (psia)	50	
Flow rate (lb-mol/h)	110	
CO ₂ (mole fraction)	0	
Stream 3C: Methanol from distillation column to stripper		Methanol from distillation column is sent to stripper.
Temperature (°F)	150	
Pressure (psia)	50	
Flow rate (lb-mol/h)	51	
CO ₂ (mole fraction)	0	
Stream 3D: Wastewater from distillation column		Water from distillation column is removed from bottom of column for disposal.
Temperature (°F)	280	
Pressure (psia)	50	
Flow rate (lb-mol/h)	59	
CO ₂ (mole fraction)	0	

TABLE 7.3 (Cont.)

Stream and Characteristic	Data	Comments on Stream Calculations
Stream 4A: CO ₂ -rich gas from flash drum 2		
Temperature (°F)	-36	Rich methanol from flash drum 1 is flashed to pressure of 100 psia to desorb major portion of CO ₂ from solvent.
Pressure (psia)	100	
Flow rate (lb-mol/h)	1,092	
CO ₂ (mole fraction)	0.9560	
Stream 4C: CO ₂ -rich gas from stripper		
Temperature (°F)	135	Final removal of CO ₂ is achieved in stripper by heat. Because of low vapor pressure of methanol, substantial amounts of methanol will be vaporized along with CO ₂ .
Pressure (psia)	14.7	
Flow rate (lb-mol/h)	6,835	
CO ₂ (mole fraction)	0.2103	
Stream 4D: CO ₂ -rich gas from phase separator		
Temperature (°F)	100	Methanol is condensed from CO ₂ -methanol mixture, and CO ₂ is separated in phase separator.
Pressure (psia)	14.7	
Flow rate (lb-mol/h)	1,955	
CO ₂ (mole fraction)	0.7354	
Stream 4F: CO ₂ -rich gas from flash drum 3		
Temperature (°F)	-50	Rich methanol solution from flash drum 2 is further flashed to pressure of 20 psia in flash drum 3 to desorb CO ₂ from solvent.
Pressure (psia)	20	
Flow rate (lb-mol/h)	1,698	
CO ₂ (mole fraction)	0.9981	
Stream 4H: Final CO ₂ -rich product		
Temperature (°F)	273	The CO ₂ -rich streams from stripper and flash drum 3 are compressed to pressure of 95 psia and then are combined with CO ₂ -rich stream from flash drum 2. This stream is further processed for pipeline specifications.
Pressure (psia)	95	
Flow rate (lb-mol/h)	4,745	
CO ₂ (mole fraction)	0.8802	
Stream 5A: Rich methanol from the absorber		
Temperature (°F)	-28	Rich methanol, which contains CO ₂ and other soluble gases, is withdrawn from bottom of tower. Temperature rise of solvent is due to heat of absorption of CO ₂ into methanol.
Pressure (psia)	235	
Flow rate (lb-mol/h)	51,147	
CO ₂ (mole fraction)	0.0995	

TABLE 7.3 (Cont.)

Stream and Characteristic	Data	Comments on Stream Calculations
Stream 5B: Recycle to absorption tower		
Temperature (°F)	-28	Rich methanol is flashed to pressure of 200 psia to desorb gases like H ₂ and CH ₄ , and the desorbed gases are recycled to absorption tower.
Pressure (psia)	200	
Flow rate (lb-mol/h)	435	
CO ₂ (mole fraction)	0	
Stream 6A: Lean methanol from stripper		
Temperature (°F)	149	Lean methanol from stripper bottom is to be circulated to absorption tower. The CO ₂ content in lean methanol is 0.02 mol/mol of methanol.
Pressure (psia)	14.7	
Flow rate (lb-mol/h)	46,532	
CO ₂ (mole fraction)	0.0196	
Stream 6B: Lean methanol from circulation pump		
Temperature (°F)	151	Lean methanol from stripper is at pressure of 14.7 psia and is pressurized to absorption tower operating pressure of 250 psia by using circulation pump.
Pressure (psia)	250	
Flow rate (lb-mol/h)	46,532	
CO ₂ (mole fraction)	0.0196	
Stream 6C: Lean methanol from heat exchanger		
Temperature (°F)	-15	Lean methanol from circulation pump is cooled against cold rich methanol from flash drum 3 to temperature of -15°F. Small portion of methanol is injected into feed gas prior to absorption to prevent icing of water.
Pressure (psia)	250	
Flow rate (lb-mol/h)	46,483	
CO ₂ (mole fraction)	0.0196	
Stream 6E: Lean methanol to absorber		
Temperature (°F)	-70	Lean methanol from heat exchanger is further cooled to temperature of -70°F by refrigeration.
Pressure (psia)	250	
Flow rate (lb-mol/h)	46,483	
CO ₂ (mole fraction)	0.0196	
Stream 7: Methanol makeup		
Temperature (°F)	70	Methanol has low vapor pressure and, hence, is lost in stripper along with CO ₂ . Also, some methanol is lost in distillation column along with wastewater.
Pressure (psia)	14.7	
Flow rate (lb-mol/h)	514	
CO ₂ (mole fraction)	0	

^a Flow rates are for one of four trains.

7.3 Power System Performance Effects of the Chilled Methanol Process

The fact that the synthesis gas must be cooled before processing by the CO₂ recovery system significantly affects the net power system output. This cooling substantially reduces the enthalpy of the synthesis gas, including the enthalpy contributed by the exothermic shift reaction. As discussed in Section 4.3, much of the thermal energy removed from the gas stream can be recovered, and the most effective use of that energy is preheating of humidified fuel gas. This effectiveness is shown in Table 4.7, where several alternative process integration strategies are compared. The heat demand for solvent regeneration (stripping and distillation) in the methanol process can be met entirely by heat recovered during cooling of the synthesis gas. The overall power balance for the plant is summarized in Table 7.4. The high power use by the CO₂ recovery system reflects the refrigeration demand. The energy penalty associated with CO₂ recovery includes both reduced output from the turbogenerator units and electricity consumption for pumping, compression, and refrigeration in the CO₂ recovery system. For the chilled methanol process, the total penalty is 179.3 MW.

TABLE 7.4 Turbine Power Output, Plant Power Use, and Net Power Output for Chilled Methanol Process

Power Variable	Power (MW)	
	Base Case	Chilled Methanol CO ₂ Recovery
Power output		
Gas turbine	311.6	269.0
Steam turbine	182.2	170.1
Internal power consumption ^a		
CO ₂ recovery		
CO ₂ compression	0	(30.9)
Solvent circulation	0	(2.1)
Solvent refrigeration	0	(91.6)
Others	0	0
Gasifier system	(35.4)	(35.4)
Net power output	458.4	279.1
Energy penalty	0	179.3

^a Parentheses indicate power consumption.

7.4 Major Equipment Specification and Preliminary Cost Estimation

7.4.1 Heat Exchanger before First-Stage Shift Reactor

The temperature of the gases from the KRW process is 1,000°F, and these gases have to be cooled to a temperature of 482°F. This temperature is found to be optimum for the first-stage shift reactor. The difference between the enthalpies of gases at these two temperatures gives the heat exchanger load. Most of this heat is used for raising saturated steam from saturated water at 417.4°F and 300 psia. This steam is used in the first-stage shift reactor. The remaining heat is used in humidifying and heating the fuel gas. The heat transfer area can be calculated from Equation 5.1.

The area required for the two operations (raising steam and heating fuel gas) is estimated separately because the approach temperature and the overall heat transfer coefficients are different. For raising steam from saturated water, a boiling film coefficient of 260 Btu/h/ft²/°F is used (Fraas 1982, Figure 4.5). The overall heat transfer coefficient for heating fuel gas is a gas-gas heat transfer coefficient and is assumed to be 5 Btu/h/ft²/°F (B&W 1975, p. 19-10). The detailed calculations are shown in Table 7.5.

7.4.2 Heat Exchanger between First and Second Shift Reactors

The shift reaction is an exothermic reaction, and the heat released in the first stage has to be removed to obtain an optimum temperature for the second-stage shift reactor. The gases have to be cooled from 680°F to 460°F. The heat load of the heat exchanger is again the difference between the enthalpies of the gases. A major portion of available heat is used in heating water from 100°F to saturated water conditions at 300 psia and 417.4°F, and the remaining heat is used for humidifying and heating the fuel gas. The surface area is calculated by using Equation 5.1, and the details are given in Table 7.5.

7.4.3 Heat Exchanger after Second-Stage Shift Reactor

The shifted gases from the second stage are at a temperature of 480°F. These gases have to be cooled to a temperature of 100°F. This cooling is achieved in three stages. In the first stage, the gases are cooled to about 417.4°F in the reboiler; in the second stage, the gases are cooled to about 120°F by heating CO₂-lean fuel gas from the absorber at 80°F to a temperature of 401°F. The final cooling to 100°F is achieved by circulating cooling water. The cooling is achieved by raising saturated steam (200 psia and 382°F) from cooling water, which is available at 70°F. The steam then can be used in the steam turbine cycle through an HRSG. The overall heat transfer coefficient

TABLE 7.5 Sizing and Cost Estimation of Major Equipment for Chilled Methanol Process

1. Heat Exchanger before First Stage for Raising**Steam**

Q = Load (Btu/h)	78,838,137
Tha = Inlet temperature of hot fluid (°F)	1,000
Thb = Outlet temperature of hot fluid (°F)	531
Pressure of hot gases (psia)	290
Tca = Inlet temperature of saturated water (°F)	417
Tcb = Outlet temperature of saturated steam (°F)	417
ΔT_1	583
ΔT_2	114
Log mean temperature difference (°F)	287
Overall heat transfer coefficient (Btu/h/ft ² /°F) ^a	260
Heat transfer area (ft ²)	1,057
Operating pressure (psia)	300
Pressure factor	1.165
Module factor	3.2

(includes all supporting equipment and connections and installation)

Purchased cost of heat exchanger in 1987 \$23,000

(mild steel construction; shell and tube floating head)

CE index for process equipment in 1987^b 320

CE index for process equipment in 1993 360.4

Installed cost of heat exchanger in 1993

\$96,569

2. Gas-Gas Heat Exchanger before First Stage for Heating Fuel Gas

Q = Load (Btu/h)	8,219,649
Tha = Inlet temperature of hot fluid (°F)	531
Thb = Outlet temperature of hot fluid (°F)	482
Pressure of hot gases (psia)	285
Tca = Inlet temperature of saturated water (°F)	401
Tcb = Outlet temperature of saturated steam (°F)	401
ΔT_1	130
ΔT_2	81
Log mean temperature difference (°F)	104
Overall heat transfer coefficient (Btu/h/ft ² /°F) ^c	5
Heat transfer area (ft ²)	15,872
Operating pressure (psia)	300
Pressure factor	1.165
Module factor	3.2

(includes all supporting equipment and connections and installation)

Purchased cost of heat exchanger in 1987 \$131,716

(mild steel construction; shell and tube floating head)

CE index for process equipment in 1987 320

CE index for process equipment in 1993 360.4

Installed cost of heat exchanger in 1993

\$553,029

TABLE 7.5 (Cont.)

3. Heat Exchanger after First Stage for Saturated Water

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

4. Heat Exchanger after First Stage for Fuel Gas Heating

Q = Load (Btu/h)	17,195,044	
Tha = Inlet temperature of hot fluid (°F)	536.6	
Thb = Outlet temperature of hot fluid (°F)	460	
Pressure of hot gases (psia)	275	
Tca = Inlet temperature of water (°F)	401	
Tcb = Outlet temperature of saturated water (°F)	401	
Saturated water pressure (psia)	200	
ΔT_1	136	
ΔT_2	59	
Log mean temperature difference (°F)	92	
Overall heat transfer coefficient (Btu/h/ft ² /°F)	5	
(hot fluid: gas, cold fluid: fuel gas) ^c		
Heat transfer area (ft ²)	37,361	
Operating pressure (psia)	275	
Pressure factor	1.165	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$183,923	
(mild steel construction; shell and tube floating head)		
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$772,229

TABLE 7.5 (Cont.)

5. Heat Exchanger after Second Stage for Fuel Gas**Heating**

Q = Load (Btu/h)	87,549,418	
Tha = Inlet temperature of hot fluid (°F)	480	
Thb = Outlet temperature of hot fluid (°F)	120	
Pressure of hot gases (psia)	275	
Tca = Inlet temperature of cold fluid (fuel gas) (°F)	80	
Tcb = Outlet temperature of cold fluid (fuel gas) (°F)	401	
ΔT_1	79	
ΔT_2	40	
Log mean temperature difference (°F)	57	
Overall heat transfer coefficient (Btu/h/ft ² /°F)	5	
(hot fluid: gas, cold fluid: fuel gas) ^c		
Heat transfer area (ft ²)	305,556	
Operating pressure (psia)	275	
Pressure factor	1.165	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$417,425	
(mild steel construction; shell and tube floating head)		
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$1,752,624

6. Gas-Gas Heat Exchanger for Raw Gas Cooling**(a) With CO₂-Rich Gas**

Q = Load (Btu/h)	1,153,557	
Tha = Inlet temperature of hot fluid (raw synthesis gas) (°F)	100	
Thb = Outlet temperature of hot fluid (raw synthesis gas) (°F)	-14	
Pressure of hot gases (psia)	250	
Tca = Inlet temperature of cold fluid (CO ₂ -rich gas) (°F)	-36.4	
Tcb = Outlet temperature of cold fluid (CO ₂ -rich gas) (°F)	84.5	
ΔT_1	15.5	
ΔT_2	22	
Log mean temperature difference (°F)	19	
Overall heat transfer coefficient (Btu/h/ft ² /°F) ^c	5	
Heat transfer area (ft ²)	12,418	
Operating pressure (psia)	275	
Pressure factor	1.165	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$119,696	
(mild steel construction; shell and tube floating head)		
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$502,562

TABLE 7.5 (Cont.)

6. (Cont.)**(b) With CO₂-Lean Fuel Gas**

Q = Load (Btu/h)	13,773,930	
Tha = Inlet temperature of hot fluid (°F)	100	
Thb = Outlet temperature of hot fluid (°F)	-14	
Pressure of hot gases (psia)	250	
Tca = Inlet temperature of cold fluid (°F)	-70	
Tcb = Outlet steam temperature (°F)	80	
ΔT_1	20	
ΔT_2	56	
Log mean temperature difference (°F)	35	
Overall heat transfer coefficient (Btu/h/ft ² /°F)	5	
(hot fluid: gas, cold fluid: gas) ^e		
Heat transfer area (ft ²)	79,086	
Operating pressure (psia)	275	
Pressure factor	1.165	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$246,407	
(mild steel construction; shell and tube floating head)		
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$1,034,578

7. First-Stage Shift Reactor

Catalyst volume (ft ³)	1,000	
Reactor volume (ft ³) (1.2 times catalyst volume)	1,200	
Reactor volume (gal)	8,963	
Pressure factor	1	
Module factor	3.05	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Purchased cost of reactor in 1987	\$9,000	
Installed cost of reactor in 1993		\$30,916

8. Second-Stage Shift Reactor

Catalyst volume (ft ³)	340	
Reactor volume (ft ³) (1.2 times catalyst volume)	408	
Reactor volume (gal)	3,041	
Pressure factor	1	
Module factor	3.05	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Purchased cost of reactor in 1987	\$5,000	
Installed cost of reactor in 1993		\$17,175

TABLE 7.5 (Cont.)

9. Cost of Shift Catalyst			
Volume of catalyst in first stage (ft ³)	1,000		
Volume of catalyst in second stage (ft ³)	340		
Cost of high-temperature catalyst (\$/ft ³)	\$50		
Cost of low-temperature catalyst (\$/ft ³)	\$250		
Total cost of catalyst			\$135,000
10. CO₂ Absorption Column			
Diameter of tower (ft)	17		
HETP (ft)	3		
No. of theoretical stages	15		
Absorber tower height (ft)	49		
(4 ft for inlet and outlet and for gas and liquid distributors)			
Volume of packing (ft ³)	10,218		
Pressure factor	2		
Cost per foot of column height (\$/ft)	\$1,800		
(mild steel construction)			
Module factor	4.16		
CE index for process equipment in 1987	320		
CE index for process equipment in 1993	360.4		
1993 installed cost of absorber			\$826,469
Cost of packing (\$/ft ³)	\$63.5		
(2-in. pall rings-metal)			
Total cost of packing			\$648,857
11. CO₂ Stripping Column			
Diameter of tower (ft)	17		
HETP (ft)	3		
No. of theoretical stages	17		
Absorber tower height (ft)	55		
(4 ft for inlet and outlet and for gas and liquid distributors)			
Volume of packing (ft ³)	11,581		
Pressure factor	2		
Cost per foot of column height (\$/ft)	\$1,800		
(mild steel construction)			
Module factor	4.16		
CE index for process equipment in 1987	320		
CE index for process equipment in 1993	360.4		
1993 installed cost of absorber			\$927,670
Cost of packing (\$/ft ³)	\$63.5		
(2-in. pall rings-metal)			
Total cost of packing			\$735,371
12. Flash Tank 1			
Methanol flow rate (lb/h)	1,458,093		
Density of methanol (lb/gal)	6.55		
Residence time (s)	180		
Slump tank volume (gal)	11,130		
Pressure factor	1		
Module factor	2.08		

TABLE 7.5 (Cont.)

12. (Cont.)

Purchased cost of slump tank in 1987 (mild steel construction)	\$28,000	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of slump tank in 1993		\$65,593

13. Recycle Compressor

Inlet pressure (psia)	200	
Outlet pressure (psia)	250	
scf/min	2,604	
Purchased cost of 197-hp reciprocating compressor in 1987 (includes electric motor drive and gear reducer)	\$80,000	
Size factor for compressor	1	
Module factor ^f	2.6	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of compressor in 1993		\$234,260

14. Flash Tank 2

Methanol flow rate (lb/h)	1,458,093	
Density of methanol (lb/gal)	6.55	
Residence time (s)	180	
Slump tank volume (gal)	11,130	
Pressure factor	1	
Module factor	2.08	
Purchased cost of slump tank in 1987 (mild steel construction)	\$28,000	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of slump tank in 1993		\$65,593

15. Flash Tank 3

Methanol flow rate (lb/h)	1,458,093	
Density of methanol (lb/gal)	6.55	
Residence time (s)	180	
Slump tank volume (gal)	11,130	
Pressure factor	1	
Module factor	2.08	
Purchased cost of slump tank in 1987 (mild steel construction)	\$28,000	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of slump tank in 1993		\$65,593

TABLE 7.5 (Cont.)

16. Flash Gas Compressor 1

Inlet pressure (psia)	20	
Outlet pressure (psia)	95	
Compressor size (hp)	1,318	
Purchased cost of centrifugal compressor in 1987 (includes electric motor drive and gear reducer)	\$320,000	
Size factor for compressor	1	
Module factor ^f	2.6	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of compressor in 1993		\$937,040

17. Flash Gas Compressor 2

Inlet pressure (psia)	14.7	
Outlet pressure (psia)	95	
Compressor size (hp)	2,561	
Purchased cost of centrifugal compressor in 1987 (includes electric motor drive and gear reducer)	\$540,000	
Size factor for compressor	1	
Module factor ^f	2.6	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of compressor in 1993		\$1,581,255

18. Solvent Circulation Pump

Horsepower	694	
Size exponent	0.79	
Purchased cost of 300-hp pump in 1987 (includes motor, coupling, base; cast iron, horizontal)	\$30,000	
Module factor	1.5	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of solvent pump in 1993		\$98,263

19. Lean-Rich Solvent Heat Exchanger

Q = Load (Btu/h)	138,607,492	
Tha = Inlet temperature of hot fluid (lean methanol) (°F)	151	
Thb = Outlet temperature of hot fluid (lean methanol) (°F)	-15	
Pressure of hot gases (psia)	250	
Tca = Inlet temperature of cold fluid (rich methanol) (°F)	-50	
Tcb = Outlet temperature of cold fluid (rich methanol) (°F)	114	
ΔT_1	37	
ΔT_2	35	
Log mean temperature difference (°F)	36	
Overall heat transfer coefficient (Btu/h/ft ² /°F) (hot fluid: gas, cold fluid: gas)	150	
Heat transfer area (ft ²)	28,851	
Operating pressure (psia)	275	
Pressure factor	1.165	

TABLE 7.5 (Cont.)

19. (Cont.)

Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$209,837	
(mild steel construction; shell and tube floating head)		
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$881,033

20. CO₂ Product Gas Compressors

Compressor 1 (hp)	2,462	
Compressor 2 (hp)	1,954	
Compressor 3 (hp)	1,955	
Purchased cost of centrifugal compressor 1 in 1987	\$540,000	
Purchased cost of centrifugal compressor 2 in 1987	\$450,000	
Purchased cost of centrifugal compressor 3 in 1987	\$450,000	
(includes electric motor drive and gear reducer)		
Size factor for compressor	1	
Module factor ^f	2.6	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of compressor 1 in 1993		\$1,581,255
Installed cost of compressor 2 in 1993		\$1,317,713
Installed cost of compressor 3 in 1993		\$1,317,713

21. Solvent Refrigeration

Refrigeration (tons)	3,818	
Purchased cost in 1987	\$1,100,000	
Temperature correction factor	3.5	
Module factor	1.46	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of solvent pump in 1993		\$6,330,651

Total Cost **\$22,676,957**

Total Cost of Four Trains **\$90,707,828**

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

^b CE, Chemical engineering.

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

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

^e Source: Perry and Chilton (1973, p. 10-40).

^f Source: Garrett (1989).

for this gas-liquid heat exchanger is taken as 5 Btu/h/ft²/°F. The area required for this duty is calculated as in Equation 5.1. The heat exchanger areas required for these operations are estimated, and the details are given in Table 7.5.

7.4.4 Heat Exchanger for Raw Gas Cooling

The fuel gas from the top of the absorber is at a temperature of -70°F, and the CO₂-rich gas from the flash drum is at a temperature of -36°F. The inlet gases to the absorber are at a temperature of 100°F. Therefore the inlet synthesis gases can be cooled to a temperature of -6°F against the fuel gas, which can be heated to a temperature of 80°F; and the CO₂-rich gas can be heated to 84.5°F. The overall heat transfer coefficient for this gas-gas heat exchanger is taken as 5 Btu/h/ft²/°F. The area required for this duty is calculated by using Equation 5.1, and the details are given in Table 7.5.

7.4.5 First-Stage Shift Reactor

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

7.4.6 Second-Stage Shift Reactor

The volume of the reactor is assumed to be 20% greater than the volume of the catalyst. The volume of the catalyst is estimated to be 340 ft³ from the kinetics and the equilibrium considerations.

7.4.7 Cost of Catalyst

The cost of the iron-based catalyst is assumed to be \$50/ft³ and that of the copper-based catalyst \$250/ft³.

7.4.8 Absorption Tower

The diameter of the absorption tower is calculated for 60% flooding velocity. Diameters are calculated at the top and bottom of the tower, and the larger of the two is taken as the tower diameter. The diameter of the tower is 17 ft. The height of the tower is calculated from the number of theoretical stages required and the HETP. The assumption is made that 15 theoretical

stages are required and that HETP is 3 ft. An additional 4 ft is provided for liquid and gas distributors.

7.4.9 Stripping Column

The CO₂ that is not released from the methanol, even after pressure reduction, is desorbed by heat. For this purpose, a stripping column is used. The diameter of the column is 17 ft, and the height of the column is 55 ft.

7.4.10 Flash Tank 1

Sufficient residence time is required for the gases to escape from the liquid; otherwise, these gases will be carried over along with the methanol and will be released along with CO₂ in subsequent stages. A residence time of 3 minutes is selected.

7.4.11 Recycle Gas Compressor

The less soluble gases released from the slump tank are recycled to the tower by compressing them to the tower operating pressure. The gases are compressed from 200 psia to a pressure of 250 psia. The horsepower of the compressor is estimated from Equation 5.4.

7.4.12 Flash Tank 2

The CO₂ is released from the methanol by reducing the pressure. This pressure reduction is carried out in stages. The first-stage flash tank is at a pressure of 100 psia. The flash tanks are designed to provide sufficient residence time for the gases to escape from the methanol. A residence time of 3 minutes is selected.

7.4.13 Flash Tank 3

The CO₂ is released from the methanol by reducing the pressure. This reduction is carried out in stages. The second-stage flash tank is at a pressure of 20 psia. The flash tanks are designed to provide sufficient residence time for the gases to escape from the methanol. A residence time of 3 minutes is selected.

7.4.14 Flash Gas Compressor 1

The CO₂ released at 20 psia is compressed to a pressure of 95 psia. The compressed CO₂ is combined with the CO₂ from the 100-psia flash drum 2. The horsepower of the compressor is estimated by using Equation 5.4.

7.4.15 Flash Gas Compressor 2

The CO₂ released from the stripper is at 14.7 psia and is compressed to a pressure of 95 psia. The compressed CO₂ is combined with the CO₂ from the 100-psia flash drum 2. The horsepower of the compressor is estimated by using Equation 5.4.

7.4.16 Solvent Circulation Pump

The lean methanol is pumped back to the absorption tower operating pressure by using a circulation pump. The horsepower required is calculated as flow work by using Equation 5.3.

7.4.17 Lean-Rich Methanol Heat Exchanger

The lean solvent from the stripper is at a temperature of 151°F, and the rich solvent from flash drum 3 is at -50°F. Therefore the lean methanol is cooled against the rich methanol. The heat exchanger area required is calculated by using Equation 5.1.

7.4.18 Carbon Dioxide Product Gas Compressors

The CO₂ from the stripper is at pressure of 95 psia. These gases have to be compressed to the pipeline pressure of 1,000 psia. This compression is done in three stages, with interstage cooling. The values for the horsepower of the three compressors are 2,462 hp, 1,954 hp, and 1,955 hp. The horsepower of the compressors is estimated by using Equation 5.4.

7.4.19 Solvent Refrigeration

The temperature of the methanol increases because of the heat of absorption and the heat of pumping. The methanol is cooled to the absorption tower operating temperature by using refrigeration. The amount of refrigeration is calculated by the difference in enthalpy of the solvent. The power required for the refrigeration is calculated by using Equation 6.1.

8 Integrated Gasification Combined Cycle with Hot Potassium Carbonate CO₂ Recovery

8.1 Feed Gas Preparation and Process Integration

As discussed in Section 4.3, the most significant issue in process integration is the effective use of process heat from cooling of the synthesis gas as required for the shift reaction and gas preparation for the CO₂ absorber. The most effective use of this thermal energy is preheating of the fuel gas feed to the gas turbine. With the addition of moisture to that feed, it can absorb most of the available heat. The balance is used for feedwater heating, raising steam for the steam cycle, and solvent regeneration. Because of differences in the absorber inlet gas specifications among the alternative CO₂ recovery systems, some variation exists in the amount of heat available. For the hot K₂CO₃ process, the CO₂-rich gas to the absorber must be cooled to 225°F, which is significantly warmer than the absorber inlet conditions for the other processes. The amount and allocation of thermal energy available from various processes are listed in Table 8.1.

In the hot K₂CO₃ process, solvent regeneration requires the use of a stripping column, which imposes a substantial thermal energy demand on the process. To subsequently recover this heat from the regenerated solvent is not practical. Thus, reboiler operation represents an important

TABLE 8.1 Heat Recovery and Allocation for Hot K₂CO₃ Process

Process	Total Enthalpy Available from Process (10 ⁶ Btu)	Allocation to Fuel Gas Preheating (10 ⁶ Btu)	Allocation to Steam Cycle (10 ⁶ Btu)	Allocation to Solvent Regeneration (10 ⁶ Btu)
Initial synthesis gas cooling to 1,000°F	1,299.3	329.1	970.2	0
Cooling before first-stage shift	32.9	0	32.9	0
Cooling after first-stage shift	68.8	34.9	33.9	0
Cooling after second-stage shift	640.7	640.7	0	0
Low-pressure turbine steam extraction	986.9	0	0	986.9

energy penalty, which is reflected in the net power output. The reboiler demand is met by a combination of thermal energy sources, including CO₂-rich gas cooling after the second-stage shift reaction, heat of compression of the CO₂ product, and extraction of steam from the low-pressure turbine. The net plant output, including credit for this heat recovery, is listed in Table 8.2.

8.2 Carbon Dioxide Recovery by Hot Potassium Carbonate

An overview of the hot K₂CO₃ recovery system is provided by Figure 8.1 (the system flow diagram) and by Table 8.3 (the corresponding line lists). A brief overview of the system is provided subsequently. More detailed information and assumptions for each stream are provided in Table 8.4.

The hot K₂CO₃ process uses a 40% solution of aqueous K₂CO₃ as the solvent. Absorption occurs at a relatively high temperature (near the solvent atmospheric-pressure boiling point), which reduces the cooling requirement for feed gas preparation. The CO₂ reacts with the solvent according to the following mildly exothermic reaction:

TABLE 8.2 Turbine Power Output, Plant Power Use, and Net Power Output for Hot K₂CO₃ Process

Power Variable	Power (MW)	
	Base Case	Hot K ₂ CO ₃
Power output		
Gas turbine	311.6	290.0
Steam turbine	182.2	130.3
Internal power consumption ^a		
CO ₂ recovery		
CO ₂ compression	0	(34.6)
Solvent circulation	0	(12.1)
Solvent refrigeration	0	0
Others	0	0
Gasifier system	(35.4)	(35.4)
Net power output	458.4	338.2
Energy penalty	0	120.2

^a Parentheses indicate power consumption.

FIGURE 8.1 Summary of Hot K_2CO_3 Process (numbers refer to streams)

TABLE 8.3 Stream Flows of Hot K₂CO₃ Process^a

Stream Data	Stream 1	Stream 2	Stream 3	Stream 4	Stream 5A
Description of stream	Absorber feed	Fuel gas	Rich K ₂ CO ₃ from absorber	Rich K ₂ CO ₃ to stripper	Lean K ₂ CO ₃ from stripper
Gases (lb-mol/h)					
CO	77.37	77.37	0.00	0.00	0.00
CO ₂	4,335.99	216.80	5,805.60	5,805.60	2,942.48
H ₂	5,611.86	5,611.86	0.00	0.00	0.00
H ₂ O	1,436.82	527.92	908.90	4,663.88	7,693.69
N ₂	7,306.65	7,306.65	0.00	0.00	0.00
Ar	88.60	88.60	0.00	0.00	0.00
CH ₄	128.77	128.77	0.00	0.00	0.00
NH ₃	2.99	2.99	0.00	0.00	0.00
H ₂ S	0.40	0.02	0.38	0.38	0.00
HCl	0.04	0.04	0.00	0.00	0.00
O ₂	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00
SO ₂	0.00	0.00	0.00	0.00	0.00
Total gas flow	18,989.49	13,961.01	6,714.89	10,469.86	10,636.17
Liquids (lb-mol/h)					
K ₂ CO ₃	0.00	0.00	111,403.40	107,648.43	155,087.50
Temperature (°F)	225.00	190.00	248.25	202.70	200.00
Pressure (psia)	250.00	247.00	247.00	17.50	18.50
Enthalpy of stream (Btu/h) (reference, 32°F)	5.33 × 10 ⁷	2.48 × 10 ⁷	4.97 × 10 ⁸	4.76 × 10 ⁸	5.63 × 10 ⁸

TABLE 8.3 Stream Flows of Hot K₂CO₃ Process^a (Cont.)

Stream Data	Stream 5B	Stream 5C	Stream 6	Stream 7	Stream 8
Description of stream	Lean K ₂ CO ₃ from reboiler	Recycle to stripper	Lean K ₂ CO ₃ after circulation pump	Lean K ₂ CO ₃ to absorber	Lean K ₂ CO ₃ to absorber top
Gases (lb-mol/h)					
CO	0.00	0.00	0.00	0.00	0.00
CO ₂	1,686.41	1,256.07	1,686.41	1,124.27	562.14
H ₂	0.00	0.00	0.00	0.00	0.00
H ₂ O	0.00	7,693.69	0.00	0.00	0.00
N ₂	0.00	0.00	0.00	0.00	0.00
Ar	0.00	0.00	0.00	0.00	0.00
CH ₄	0.00	0.00	0.00	0.00	0.00
NH ₃	0.00	0.00	0.00	0.00	0.00
H ₂ S	0.00	0.00	0.00	0.00	0.00
HCl	0.00	0.00	0.00	0.00	0.00
O ₂	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00
SO ₂	0.00	0.00	0.00	0.00	0.00
Total gas flow	1,686.41	8,949.76	1,686.41	1,124.27	562.14
Liquids (lb-mol/h)					
K ₂ CO ₃	111,403.40	43,684.10	111,403.40	74,268.93	37,134.47
Temperature (°F)	230.00	230.00	234.62	234.62	190.00
Pressure (psia)	18.50	18.50	250.00	250.00	250.00
Enthalpy of stream (Btu/h) (reference, 32°F)	4.44 × 10 ⁸	3.40 × 10 ⁸	4.55 × 10 ⁸	3.03 × 10 ⁸	118 × 10 ⁸

TABLE 8.3 Stream Flows of Hot K₂CO₃ Process^a (Cont.)

Stream Data	Stream 9	Stream 10	Stream 11	Stream 12	Stream 13
Description of stream	CO ₂ -rich gas from stripper	CO ₂ -rich gas after condenser	Recycle to stripper	CO ₂ -rich product	Wastewater
Gases (lb-mol/h)					
CO	0.00	0.00	0.00	0.00	0.00
CO ₂	4,119.19	4,119.19	0.00	4,119.19	0.00
H ₂	0.00	0.00	0.00	0.00	0.00
H ₂ O	12,357.57	12,357.57	11,448.66	236.53	672.37
N ₂	0.00	0.00	0.00	0.00	0.00
Ar	0.00	0.00	0.00	0.00	0.00
CH ₄	0.00	0.00	0.00	0.00	0.00
NH ₃	0.00	0.00	0.00	0.00	0.00
H ₂ S	0.38	0.38	0.00	0.38	0.00
HCl	0.00	0.00	0.00	0.00	0.00
O ₂	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00
SO ₂	0.00	0.00	0.00	0.00	0.00
Total gas flow	16,477.14	16,477.14	11,448.66	4,356.10	672.37
Liquids (lb-mol/h)					
K ₂ CO ₃	0.00	0.00	0.00	0.00	0.00
Temperature (°F)	212.00	100.00	100.00	100.00	100.00
Pressure (psia)	14.70	14.70	14.70	14.70	14.70
Enthalpy of stream (Btu/h) (reference, 32°F)	2.63 × 10 ⁸	1.76 × 10 ⁷	1.40 × 10 ⁷	6.29 × 10 ⁶	8.23 × 10 ⁵

^a Flow rates are for one of four trains.

TABLE 8.4 Descriptions of Stream Flows of Hot K_2CO_3 Process^a

Stream and Characteristic	Data	Comments on Stream Calculations
Stream 1: Gas feed to absorber		
Temperature (°F)	225	In hot K ₂ CO ₃ process, absorber operates at 230°F. Therefore, synthesis gases require cooling to 225°F. Amount of water in gas corresponds to vapor pressure of water at 225°F.
Pressure (psia)	250	
Flow rate (lb-mol/h)	18,990	
CO ₂ (mole fraction)	0.2283	
Stream 2: Fuel gas		
Temperature (°F)	190	Composition of H ₂ -rich fuel gas stream is based on 95% removal of CO ₂ . Amount of H ₂ S removed is also assumed to be 95%. Gases leave absorber at temperature equal to that of solvent at top of absorber, which is 190°F. Water content of this stream corresponds to vapor pressure of water at 190°F.
Pressure (psia)	247	
Flow rate (lb-mol/h)	13,961	
CO ₂ (mole fraction)	0.0155	
Stream 3: Rich K ₂ CO ₃ from absorber		
Temperature (°F)	248	Solvent flow rate is designed for CO ₂ concentration of 5.85 scf/gal of 40% K ₂ CO ₃ solution. Lean solvent contains CO ₂ at 2.53 scf/gal, giving net pickup of 3.32 scf/gal. Temperature of solvent increases to 248°F because of heat of reaction (261 Btu/lb of CO ₂ absorbed). Temperature is calculated by making enthalpy balance around absorber.
Pressure (psia)	247	
Flow rate (lb-mol/h)	118,118	
CO ₂ (mole fraction)	0.0492	
Stream 4: Rich K ₂ CO ₃ entering stripper		
Temperature (°F)	203	Rich solvent is flashed from 247 to 18 psia before stripper to remove as much CO ₂ as possible. Estimates of amount of CO ₂ flashed vary from one-third to two-thirds (Kohl and Riesenfeld 1985) of net CO ₂ absorbed. In present design, value of one-half is used. Along with CO ₂ , water is also flashed, bringing temperature of solvent to 203°F. Drop in temperature is estimated from heat of vaporization of water and heat of desorption of CO ₂ .
Pressure (psia)	17.5	
Flow rate (lb-mol/h)	118,118	
CO ₂ (mole fraction)	0.0492	

TABLE 8.4 (Cont.)

Stream and Characteristic	Data	Comments on Stream Calculations
Stream 5B: Lean K ₂ CO ₃ from stripper		
Temperature (°F)	230	Reboiler operates at temperature of 230°F and pressure of 18.5 psia, corresponding to CO ₂ level of 2.53 scf/gal, in lean solvent stream.
Pressure (psia)	18.5	
Flow rate (lb-mol/h)	113,090	
CO ₂ (mole fraction)	0.0149	
Stream 6: Lean K ₂ CO ₃ after circulation pump		
Temperature (°F)	235	Solvent has to be pressurized from reboiler pressure of 19 psia to absorber operating pressure of 250 psia. Circulation pump of 885 hp, with efficiency of 75%, is used for this purpose; and temperature of solvent increases to 235°F.
Pressure (psia)	250	
Flow rate (lb-mol/h)	113,090	
CO ₂ (mole fraction)	0.0149	
Stream 7: Lean K ₂ CO ₃ fed to middle of absorber		
Temperature (°F)	235	Major portion (two-thirds) of solvent is fed hot (235°F) to absorber, without any cooling, to take advantage of higher reaction rates.
Pressure (psia)	250	
Flow rate (lb-mol/h)	75,393	
CO ₂ (mole fraction)	0.0149	
Stream 8: Lean K ₂ CO ₃ after cooling		
Temperature (°F)	190	Remaining solvent (one-third) is cooled from 235°F to 190°F and fed at top of absorber.
Pressure (psia)	250	
Flow rate (lb-mol/h)	37,697	
CO ₂ (mole fraction)	0.0149	
Stream 9: CO ₂ -rich gas from stripper		
Temperature (°F)	212	Remaining CO ₂ in solvent after flashing has to be stripped by application of heat. Reflux ratio of 1:4 is maintained in column. Reflux ratio here is defined as amount of acid gases leaving stripper divided by amount of water returned to top of stripper. Therefore, composition of this stream represents amount of CO ₂ and water flashed and CO ₂ and water vaporized by application of heat.
Pressure (psia)	14.7	
Flow rate (lb-mol/h)	16,477	
CO ₂ (mole fraction)	0.2500	

TABLE 8.4 (Cont.)

Stream and Characteristic	Data	Comments on Stream Calculations
Stream 10: CO ₂ -rich gas after condenser		
Temperature (°F)	100	Gases leaving stripper are at temperature of 212°F, and they have to be cooled and water has to be condensed so that portion can be returned to stripping column as reflux. To decrease reboiler load, gases are cooled to 100°F only.
Pressure (psia)	14.7	
Flow rate (lb-mol/h)	16,477	
CO ₂ (mole fraction)	0.2500	
Stream 11: Recycle to absorber		
Temperature (°F)	100	Water condensed is returned to stripping column to maintain reflux ratio of 1:4.
Pressure (psia)	14.7	
Flow rate (lb-mol/h)	11,449	
CO ₂ (mole fraction)	0	
Stream 12: CO ₂ -rich product		
Temperature (°F)	100	This stream is CO ₂ -rich gas stream for further compression and disposal. Water corresponds to vapor pressure of water at 100°F.
Pressure (psia)	14.7	
Flow rate (lb-mol/h)	4,356	
CO ₂ (mole fraction)	0.9456	
Stream 13: Excess water stream		
Temperature (°F)	100	Excess water has to be removed from system to maintain water balance.
Pressure (psia)	14.7	
Flow rate (lb-mol/h)	672	
CO ₂ (mole fraction)	0	

^a Flow rates are for one of four trains.

Solvent regeneration is accomplished through pressure reduction and heating. Pressure is reduced by an expansion valve on the inlet to the stripper. Most of the heat of regeneration is obtained from steam supplied to a recirculating reboiler. The sources of steam or heat for steam supply are listed in Table 8.1. The regeneration process releases CO₂, which is cooled to condense accompanying solvent vapors and to prepare CO₂ for compression to pipeline disposal conditions.

To attain a high CO₂ removal efficiency, the solvent stream has been split before it enters the absorber. A portion of the solvent stream is cooled to increase CO₂ solubility. This solvent is

fed to the top of the absorber, where the CO₂ partial pressure is lower. The balance of the solvent is fed at a lower point in the absorber and is responsible for most CO₂ absorption.

8.3 Power System Performance Effects of the Hot Potassium Carbonate Process

The fact that the synthesis gas must be cooled before processing by the CO₂ recovery system significantly affects the net power system output. This cooling substantially reduces the enthalpy of the synthesis gas, including the enthalpy contributed by the exothermic shift reaction. As discussed in Section 4.3, much of the thermal energy removed from the gas stream can be recovered, and the most effective use of that energy is preheating of humidified fuel gas. This effectiveness is shown in Table 4.7, where several alternative process integration strategies are compared. The heat demand for solvent regeneration (stripping) in the hot K₂CO₃ process cannot be met entirely by heat recovered during cooling of the synthesis gas. In fact, most of the heat for stripping is obtained by diverting steam from the low-pressure turbine, as listed in Table 8.1. The resulting reduction in power output is 54.79 MW. This reduction is due to steam extraction only. Other losses may occur. The energy penalty associated with CO₂ recovery includes both reduced output from the turbogenerator units and electricity consumption for pumping, compression, and refrigeration in the CO₂ recovery system. For the hot K₂CO₃ process, the total penalty is 120.2 MW.

8.4 Major Equipment Specification and Preliminary Cost Estimation

8.4.1 Heat Exchanger before First-Stage Shift Reactor

The temperature of the gases from the KRW process is 1,000°F, and these gases have to be cooled to a temperature of 482°F. This temperature is found to be optimum for the first-stage shift reactor. The difference between the enthalpies of gases at these two temperatures gives the heat exchanger load. Most of this heat is used for raising saturated steam from saturated water at 417.4°F and 300 psia. This steam is used in the first-stage shift reactor. By this process, the gases are cooled to 531°F, and the remaining heat is used in the steam cycle through an HRSG, thereby cooling the gases to 482°F. The heat transfer area required can be calculated by using Equation 5.1.

The area required for the heat exchanger is estimated separately because the approach temperature and the overall heat transfer coefficients are different. For raising steam from saturated water, a boiling film coefficient of 260 Btu/h/ft²/°F is used (Fraas 1982, Figure 4.5). The detailed calculations are shown in Table 8.5.

TABLE 8.5 Sizing and Cost Estimation of Major Equipment for Hot K₂CO₃ Process**1. Heat Exchanger before First Stage for Raising****Steam**

Q = Load (Btu/h)	78,838,137	
Tha = Inlet temperature of hot fluid (°F)	1,000	
Thb = Outlet temperature of hot fluid (°F)	531	
Pressure of hot gases (psia)	295	
Tca = Inlet temperature of saturated water (°F)	417	
Tcb = Outlet temperature of saturated steam (°F)	417	
ΔT_1	583	
ΔT_2	114	
Log mean temperature difference (°F)	287	
Overall heat transfer coefficient (Btu/h/ft ² /°F) ^a	260	
Heat transfer area (ft ²)	1,057	
Operating pressure (psia)	300	
Pressure factor	1.165	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$23,000	
(mild steel construction; shell and tube floating head)		
CE index for process equipment in 1987 ^b	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$96,569

2. Heat Exchanger after First Stage for Saturated Water

Q = Load (Btu/h)	31,747,489	
Tha = Inlet temperature of hot fluid (°F)	680	
Thb = Outlet temperature of hot fluid (°F)	537	
Pressure of hot gases (psia)	285	
Tca = Inlet temperature of water (°F)	100	
Tcb = Outlet saturated temperature (°F)	417	
ΔT_1	263	
ΔT_2	437	
Log mean temperature difference (°F)	342	
Overall heat transfer coefficient (Btu/h/ft ² /°F)	40	
(hot fluid: gas, cold fluid: gas) ^c		
Heat transfer area (ft ²)		
Operating pressure (psia)	2,319	
Pressure factor	285	
Module factor	1.165	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$40,000	
(mild steel construction; shell and tube floating head)		
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$167,946

TABLE 8.5 (Cont.)

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

Q = Load (Btu/h)	160,170,411	
Tha = Inlet temperature of hot fluid (°F)	480	
Thb = Outlet temperature of hot fluid (°F)	225	
Pressure of hot gases (psia)	275	
Tca = Inlet temperature of fuel gas (°F)	190	
Tcb = Outlet temperature of fuel gas (°F)	450	
ΔT_1	30	
ΔT_2	35	
Log mean temperature difference (°F)	32	
Overall heat transfer coefficient (Btu/h/ft ² /°F) ^d	5	
Heat transfer area (ft ²)	987,615	
Operating pressure (psia)	275	
Pressure factor	1.165	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$659,602	
(mild steel construction; shell and tube floating head)		
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$2,769,444

4. First-Stage Shift Reactor

Catalyst volume (ft ³)	1,000	
Reactor volume (ft ³) (1.2 times catalyst volume)	1,200	
Reactor volume (gal)	8,963	
Pressure factor	1	
Module factor	3.05	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Purchased cost of reactor in 1987	\$9,000	
Installed cost of reactor in 1993		\$30,916

5. Second-Stage Shift Reactor

Catalyst volume (ft ³)	340	
Reactor volume (ft ³) (1.2 times catalyst volume)	408	
Reactor volume (gal)	3,041	
Pressure factor	1	
Module factor	3.05	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Purchased cost of reactor in 1987	\$5,000	
Installed cost of reactor in 1993		\$17,175

TABLE 8.5 (Cont.)

6. Cost of Shift Catalyst			
Volume of catalyst in first stage (ft ³)	1,000		
Volume of catalyst in second stage (ft ³)	340		
Cost of high-temperature catalyst (\$/ft ³)	\$50		
Cost of low-temperature catalyst (\$/ft ³)	\$250		
Total cost of catalyst			\$135,000
7. CO₂ Absorption Column			
Diameter of tower (ft)	8		
HETP (ft)	3		
No. of theoretical stages	15		
Absorber tower height (ft)	49		
(4 ft for inlet and outlet and for gas and liquid distributors)			
Volume of packing (ft ³)	2,263		
Pressure factor	2		
Cost per foot of column height (\$/ft)	\$1,000		
(mild steel construction)			
Module factor	4.16		
CE index for process equipment in 1987	320		
CE index for process equipment in 1993	360.4		
1993 installed cost of absorber			\$459,150
Cost of packing (\$/ft ³)	\$63.5		
(2-in. pall rings-metal)			
Total cost of packing			\$143,691
8. CO₂ Stripping Column			
Diameter of tower (ft)	12		
HETP (ft)	3		
No. of theoretical stages	17		
Absorber tower height (ft)	55		
(4 ft for inlet and outlet and for gas and liquid distributors)			
Volume of packing (ft ³)	5,770		
Pressure factor	2		
Cost per foot of column height (\$/ft)	\$1,400		
(mild steel construction)			
Module factor	4.16		
CE index for process equipment in 1987	320		
CE index for process equipment in 1993	360.4		
1993 installed cost of absorber			\$721,521
Cost of packing (\$/ft ³)	\$63.5		
(2-in. pall rings-metal)			
Total cost of packing			\$366,413
9. Heat Exchanger for Lean Solvent Cooling			
Q = Load (Btu/h)	33,383,789		
T _{ha} = Inlet temperature of hot fluid (°F)	235		
T _{hb} = Outlet temperature of hot fluid (°F)	190		
Pressure of hot gases (psia)	250		
T _{ca} = Inlet water temperature (°F)	100		
T _{cb} = Outlet temperature of saturated water (°F)	180		

TABLE 8.5 (Cont.)

9. (Cont.)

ΔT_1	55	
ΔT_2	90	
Log mean temperature difference ($^{\circ}\text{F}$)	70.84	
Overall heat transfer coefficient ($\text{Btu/h/ft}^2/^{\circ}\text{F}$) ^c	150	
Heat transfer area (ft^2)	3,142	
Operating pressure (psia)	250	
Pressure factor	1.165	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$50,000	
(mild steel construction; shell and tube floating head)		
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$209,933

10. Stripper Overhead Condenser

Q = Load (Btu/h)	245,129,023	
Tha = Inlet temperature of hot fluid ($^{\circ}\text{F}$)	212	
Thb = Outlet temperature of hot fluid ($^{\circ}\text{F}$)	100	
Tca = Inlet water temperature ($^{\circ}\text{F}$)	70	
Tcb = Outlet water temperature ($^{\circ}\text{F}$)	165	
ΔT_1	47	
ΔT_2	30	
Log mean temperature difference ($^{\circ}\text{F}$)	37.87	
Overall heat transfer coefficient ($\text{Btu/h/ft}^2/^{\circ}\text{F}$)	40	
(hot fluid: gas, cold fluid: water) ^e		
Heat transfer area (ft^2)	161,839	
Operating pressure (psia)	14.7	
Pressure factor	1	
Module factor	3.2	
(includes all supporting equipment and connections and installation)		
Purchased cost of heat exchanger in 1987	\$730,415	
(mild steel construction; shell and tube floating head)		
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		\$2,632,416

11. Solvent Circulation Pump

Horsepower	4,076	
Size exponent	0.79	
Purchased cost of 300-hp pump in 1987	\$30,000	
(includes motor, coupling, base; cast iron, horizontal)		
Module factor	1.5	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of solvent pump in 1993		\$398,077

TABLE 8.5 (Cont.)

12. CO₂ Product Gas Compressors

Compressor 1 (hp)	4,007	
Compressor 2 (hp)	3,791	
Compressor 3 (hp)	3,789	
Purchased cost of centrifugal compressor 1 in 1987	\$800,000	
Purchased cost of centrifugal compressor 2 in 1987	\$750,000	
Purchased cost of centrifugal compressor 3 in 1987	\$750,000	
(includes electric motor drive and gear reducer)		
Size factor for compressor	1	
Module factor ^f	2.6	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of compressor 1 in 1993		\$2,342,600
Installed cost of compressor 2 in 1993		\$2,196,188
Installed cost of compressor 3 in 1993		\$2,196,188
Total Cost		\$14,883,227
Total Cost of Four Trains		\$59,532,908

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

^b CE, Chemical engineering.

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

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

^e Source: Peters and Timmerhaus (1980, p. 651).

^f Source: Garrett (1989).

8.4.2 Heat Exchanger between First and Second Shift Reactors

The shift reaction is an exothermic reaction, and the heat released in the first stage has to be removed to obtain an optimum temperature for the second-stage shift reactor. The gases have to be cooled from 680°F to 460°F. The heat load of the heat exchanger is again the difference between the enthalpies of the gases. A major portion of the available heat is used in heating water from 100°F to saturated water conditions at 300 psia and 417.4°F. By this process, the gases are cooled to 537°F, and the remaining heat is used in the steam cycle through an HRSG to achieve a gas temperature of 460°F. The surface area is calculated by using Equation 5.1, and the details are given in Table 8.5.

8.4.3 Heat Exchanger after Second-Stage Shift Reactor

The shifted gases from the second stage are at a temperature of 480°F. These gases have to be cooled to a temperature of 225°F. The cooling is achieved by heating CO₂-lean fuel gas from the absorber at 190°F to a temperature of 450°F. The heat exchanger areas required for these operations are given in Table 8.5.

8.4.4 First-Stage Shift Reactor

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

8.4.5 Second-Stage Shift Reactor

The volume of the reactor is assumed to be 20% greater than the volume of the catalyst. The volume of the catalyst is estimated to be 340 ft³ from the kinetics and the equilibrium considerations.

8.4.6 Cost of Catalyst

The cost of the iron-based catalyst is assumed to be \$50/ft³ and that of the copper-based catalyst \$250/ft³.

8.4.7 Absorption Tower

The diameter of the absorption tower is calculated for 60% flooding velocity. Diameters are calculated at the top and bottom of the tower, and the larger of the two is taken as the tower diameter. The diameter of the tower is 8 ft. The height of the tower is calculated from the number of theoretical stages required and the HETP. The assumption is made that 15 theoretical stages are required and that HETP is 3 ft. An additional 4 ft is provided for liquid and gas distributors.

8.4.8 Stripping Column

The diameter again is calculated for 60% flooding velocity. The diameter of the tower is 12 ft. The number of theoretical stages required is 17, and the HETP is 3 ft. Therefore, the total height of the column is 55 ft, including 4 ft for gas and liquid distributors.

8.4.9 Lean Solvent Cooler

The lean K_2CO_3 solution from the reboiler is at a temperature of 235°F. A portion (about one-third) of this solution is cooled to 190°F by circulating cooling water. The area required for this duty is calculated by using Equation 5.1, and the details are given in Table 8.5.

8.4.10 Stripper Overhead Condenser

The stripper overhead contains CO_2 , as well as water that is evaporated. Therefore, to separate the CO_2 and water vapor, the gases have to be cooled. The condensed water will be recycled to the top of the stripper. The gases are at a temperature of 212°F and are cooled to 100°F by using cooling circulating water.

8.4.11 Solvent Circulation Pump

The lean K_2CO_3 solution from the stripper is at 14.7 psia and is pumped back to the absorber operating pressure of 250 psia. The horsepower of this circulating pump is estimated as flow work by using Equation 5.3.

8.4.12 Carbon Dioxide Product Gas Compressors

The CO_2 from the stripper is at 14.7 psia. These gases have to be compressed to the pipeline pressure of 1,000 psia. This compression is done in three stages, with interstage cooling. The values for the horsepower of the three compressors are 4,007 hp, 3,791 hp, and 3,789 hp. The horsepower of the compressors is estimated by using Equation 5.4.

9 Integrated Gasification Combined Cycle with Calcium Oxide CO₂ Recovery

9.1 Background

Louisiana State University (LSU) has ongoing studies to develop a separation technique for coal-derived gases that will operate at moderate to high temperatures by using calcium-based sorbents (Harrison et al. 1993). Carbon dioxide is absorbed in the exothermic but reversible noncatalytic reaction:

Carbon dioxide, CO, H₂, and water all participate in the shift reaction system. Interactions of IGCC gas mixtures with calcium oxide (CaO) must take into account the fact that shift chemistry, with the absorption of CO₂ product, will operate during this process.

9.2 Sorbents

Nine combinations of calcium sorbents and techniques of physical preparation were investigated at LSU. On the basis of this screening, two favorable sorbents for this reaction scheme appear to be dolomite limestone [CaMg(CO₃)₂] and calcium acetate.

Dolomite contains magnesium carbonate, which appears to be unaffected by the temperatures and pressures used to calcine and return the calcium carbonate (CaCO₃) to its original reactive oxide form (Harrison et al. 1991, p. 3). Because of this low reactivity, the magnesium carbonate apparently stabilizes the physical structure of the sorbent, maintaining pore structure and surface area. Dolomite is also less costly than calcium acetate.

Calcium acetate appears to show good promise because it also maintains a very open structure. Electron microscopic studies on these crystals as received show them to be needle-like and subject to shattering if handled at all roughly (Harrison et al. 1991, p. 117). Results from the five-cycle tests of this compound show that the needle-like morphology is gradually replaced by grain-like crystals that maintain a very open structure.

9.3 Optimum Conditions for Reaction

The reaction for CO₂ capture in Equation 9.1 is designated as "carbonation," while the reverse of this reaction is designated as "calcination." The optimum conditions for reaction (Harrison et al. 1991, p. 46) appear to be the following:

- Carbonation temperature of 650-750°C;
- Carbonation pressure equal to gasifier pressure;
- Calcination temperature of 750°C; and
- Calcination pressure of 15 atm (Harrison et al. 1991, p. 76).

The LSU system of CO₂ capture should be advantageous when compared with the energy losses that accompany commercially available processes that require cooling the gases. To maximize the capture of CO₂ and CO, the temperature should be lowered to 500°C from 650°C.

9.4 Carbon Dioxide Capture: Carbonation

Over a wide range of operating conditions, CO₂ removals greater than 95% are possible. At lower temperatures ($550^{\circ}\text{C} < T < 650^{\circ}\text{C}$), the partial pressure of the CO₂ is low, and the reaction can proceed to even higher levels of removal; however, these lower temperatures are not favored because the ability of the sorbents to keep their capacity for removal is not well maintained (Harrison et al. 1991, p. 38).

Between 1 atm and 15 atm, the reactivity of the sorbent was found to decrease 56-79% with increasing pressure, possibly as an artifact of using the electrobalance reactor (Harrison et al. 1991, p. 40). This finding leads to the conclusion that transport resistances are important in establishing reactivity. At the same time, the stability of the sorbent appears to be unaffected by pressure; hence, the operation of this system will continue at typical operating pressures for gasification near 15 atm.

9.5 Carbon Dioxide Release: Calcination

Two-cycle tests of sorbent carbonation-calcination all employed a pure nitrogen sweep as the calcination gas (Harrison et al. 1991, p. 8). Five-cycle tests of the sorbent employed the same strategy by using a pure nitrogen sweep at 1-15 atm (Harrison et al. 1991, p. 51). The option of *soft-burning* the sorbent between 980°C and 1,100°C to produce a pure CO₂ atmosphere was examined by LSU but degrades the performance of the sorbent.

The strategy of nitrogen sweeps presents an economic challenge in integrating the LSU process with an IGCC system. The net consequence will be the necessity of having the CO₂ absorbed and separated from the nitrogen-carrier gas in a second system that would look very similar to commercial CO₂ separation systems that have been investigated for the recovery of CO₂ from the flue gas from pulverized-coal-fired boilers.

9.6 Macroscale Physical Properties of the Sorbent, Support Geometry, and Bed Depth

At present, excellent contact between the gas and sorbent has been maintained by loosely packing test sorbent in two different types of reactors. Initially, a high-temperature and high-pressure electrobalance reactor was employed by LSU, while current studies are conducted in a fixed-bed reactor of 5/8-in. diameter packed with approximately 2 in. of sorbent supported by quartz wool. This second design should hold approximately 10 g of test sorbent.

Further scale-up testing appears to be necessary to ascertain what geometry, support structure, and bed depth will be effective for maintaining the necessary efficiency of contact in a commercial-scale system; however, that a fixed-bed design would be employed is unlikely. Researchers at LSU currently believe that the favored system would be a dual fluidized bed, with the first bed for carbonation and the second for calcination. Testing of such a system may be proposed to the U.S. Department of Energy (DOE) in the future.

9.7 Louisiana State University Carbonation as a Shift Reactor

The current thrust of research in the LSU program is pursuing the potential application of this sorbent reaction system to promote shift (Harrison et al. 1994):

Regardless of whether this research initiative results in a viable alternative to current shift catalysts, the LSU work signifies that the integration of this process, either as an alternative to shift or after shift, will require considerable care so that feed compositions and temperatures are compatible and so that the further reactions of these gases are understood.

9.8 Process Performance with Trace Gases

The product gas from the shift reactors appears in Table 4.4 (stream 5). The performance of the LSU system when trace gases are present needs to be established for purposes of design. Gases that will interact in the shift reaction (CO₂, CO, H₂, and H₂O) have already been alluded to in Section 9.7. The presence of trace amounts of CH₄ and heavier hydrocarbons and of the trace sulfur species, carbonyl sulfide (COS) and H₂S, needs to be understood in greater detail.

10 Integrated Gasification Combined Cycle with Low-Temperature Polymer Membrane CO₂ Recovery

10.1 Gas Pretreatment and Operating Conditions

Polymeric membranes generally show their greatest advantage under a combination of the following three conditions: (1) mixtures of similar chemical and physical properties, (2) mixtures of structural or position isomers, and (3) thermally unstable compounds (Perry and Chilton 1973, p. 17-37). None of these three conditions characterize the separation of the shift product gas. At the same time, the need to operate polymer membranes at high pressures generally limits their range of operating temperatures to the ambient. For the case of IGCC, this limitation means that the thermodynamic penalties for cooling will once again be significant.

10.2 Polymer Membrane Selectivity

The recovery of CO₂ from IGCC shift product streams by using both commercially available gas-separation polymer membranes and developmental gas-absorption polymer membranes was recently investigated by the TNO-Institute of Environmental and Energy Technology, Apeldoorn, the Netherlands (Feron et al. 1992). This study concluded that three stages of separation are required for the operation of the membrane CO₂ process. First, the nitrogen must be separated from the CO₂. Second, the CO₂ must be separated from H₂ in two stages. The operation of all of these stages assumes pressurization to 77 atm (1,132 lb/in.²) and cooling of the gas to ambient conditions.

Among the commercially available polymeric materials for CO₂-N₂ selectivity, cellulose acetate, with a CO₂-N₂ selectivity of 66.5:1, shows the best performance; however, the work by Feron et al. (1992) finds that this selectivity is still lower by a factor of 3 than what would be required for a competitive polymeric membrane process: "CO₂-N₂-selectivities in excess of 200 will be competitive with conventional amine absorption technology" (p. 423).

The highest H₂-CO₂ selectivity commercially available is 10:1 by using polyimide. Hence, even though a high driving force is available, the production of a pipeline CO₂ mixture of 95% is not possible in a single stage. Feron et al. (1992, p. 425) find that: "In order to meet the requirements regarding the coal gasification process, gas selectivities need to be increased to at least 150." Even after two stages of separation, leaving 5% H₂ in the permeate going to the CO₂ pipeline may represent an unacceptable loss of fuel, as well as a materials problem, because the pipeline would now be subject to H₂ embrittlement.

10.3 Membrane Separation: Conclusions

The competitive success of membrane separation technology will require that the performance of commercially available materials be improved as follows:

- $\text{CO}_2:\text{N}_2$ by a factor of 3, and
- $\text{CO}_2:\text{H}_2$ by a factor of 15.

At the same time, research work on polymeric materials will, by its nature, continue to be a low-temperature high-pressure process. The conclusions from the study by Feron et al. (1992) were that the application of commercially available gas-separation membranes for CO_2 recovery was technically feasible, but unattractive as a consequence of the high energy use for compression and cooling. The estimated cost for a two-stage gas-separation polymer membrane system producing pipeline-quality (95%) CO_2 were "at least double the cost of a conventional MEA-absorption process" (Feron et al. 1992, p. 423). Research efforts to develop high-temperature membrane separation using nonpolymeric materials are under way and may prove to be of interest in the future.

11 Transport of CO₂ by Pipeline

11.1 Commercial Considerations of CO₂ Separation Systems for Pipeline Gas

Before entering the pipeline, CO₂ and any contained gases must be purified, dried, and compressed. Purification is necessary because contaminants, such as H₂S and SO₂, can cause corrosion problems in the pipeline. The water must be removed to prevent the formation of acids because any trace amounts of water vapor in the pipeline would combine with the CO₂, SO_x, and NO_x to produce harmful acids. This gas preparation has taken place in the CO₂ recovery system; nevertheless, some discussion of separation processes is warranted in this section (i.e., from the perspective of how each process affects pipeline considerations). Four basic methods of separation will briefly be discussed: (1) chemical solvent methods, (2) physical solvent methods, (3) membrane systems, and (4) cryogenics.

Chemical solvent methods are used when the partial pressure of CO₂ is low. These systems extensively use alkanolamines such as MEA, diethanolamine (DEA), triethanolamine (TEA), diglycolamine (DGA), and methyl diethanolamine (MDEA) for acid gas cleanup or separation (corrosion is a problem). Dow has developed a proprietary solvent formulation (FS-1) for use as a corrosion inhibitor in amine systems recovering CO₂ from flue gases. An alternative is hot K₂CO₃ as a solvent. Solvent regeneration is accomplished by steam stripping. The advantage or disadvantage of these methods is the necessity of removing SO_x and NO_x before attempting separation; otherwise, the regeneration of the alkanolamine would be too difficult. Because this preconditioning is also a requirement for the combustion turbine, such preconditioning is an essential operation in the IGCC system.

Physical solvent systems absorb and desorb acid gases by means of pressure swings. Flashing is used in place of steam stripping. These systems usually operate above 400 psig. Available processes (and their solvents) include *N*-methylpyrrolidone (Purisol), methanol (Rectisol), and dimethyl ether of polyethylene glycol (Selexol). An advantage of this system with respect to IGCC is that the need to compress the gas to approximately 400 psig from atmospheric pressure is not a penalty because gas emerges from hot-gas cleanup in a compressed state.

Membrane systems must operate at modest temperatures to protect the polymers from which they are formed. To effect the separations required to prepare this gas mixture for feeding to a CO₂ pipeline, at least two banks of high-pressure membranes would be required to produce a gas composition that is of lower purity than would be obtained by any of the other technologies.

Cryogenic distillation can be used to separate CO₂ from CH₄ (also from higher hydrocarbons and H₂S). This process requires that the process stream be compressed and cooled (in stages) to condense the CO₂, as a means of separating it from the noncondensable hydrogen. Again, an advantage of this method is the possibility of recovering the energy cost of compression

because the CO₂ is already in compression, and the final separated product must be compressed to supercritical conditions before introduction into the pipeline.

11.2 Supercritical Compression for CO₂ Pipeline

The CO₂ gas product from the various recovery systems that were investigated was put on a common basis by having the gas compressed and delivered to the pipeline compression station at 1,000 psia. This compression has been done in three stages. At the pipeline station, trim cooling takes place to lower the inlet temperature of the gas to 122°F for its final stage of compression.

The compressibility factor for CO₂ above 1,400 psig and 60°F is less than 0.3. Manufacturers of compressors desire a minimum compressibility factor of 0.5 at 1,400 psig. Therefore, a compressor inlet temperature of 130°F solves this requirement for a minimum compressibility factor. Thus, a preheater is provided to heat the CO₂ before each compressor. (The number of pipeline booster compressors required is based on pipeline pressure, pressure drop, flow rate, and pipeline length.)

The base case chosen for this study will employ a pipeline operating at 2,100 psia, a supercritical pressure. The compressibility factor for CO₂ was based on previous work in this area (Farris 1983), and consideration for maintaining a compressibility factor above 0.5 led to adopting an inlet temperature of 122°F (Figure 11.1). Because the pipeline compression conditions are somewhat unusual, details of the calculations are included here and follow a published method (Lynch 1985):

$$T_2 = T_1(P_2/P_1)^{(k-1)/k} , \quad (11.1)$$

where

$$k = C_p/C_v = 7.08/5.09 = 1.39 \text{ (at } 100^\circ\text{C),}$$

$$P_1 = \text{pressure at compressor inlet (1,000 psia),}$$

$$P_2 = \text{pressure at compressor outlet (2,100 psia),}$$

$$T_1 = \text{temperature at compressor inlet (122°F), and}$$

$$T_2 = \text{temperature at compressor outlet, calculated to be } 254^\circ\text{F.}$$

FIGURE 11.1 Compressibility Factor for CO₂ versus Pressure at Various Temperatures (adapted from Figure 1 of Farris [1983]) (Source: Lynch 1985).

Then, with reference to Figure 11.1, the average dimensionless compressibility is $Z = 0.67$, and the compressor brake horsepower may be directly calculated from the following equation (Segelar 1974, p. 8/52):

$$bhp = 3.0325 \times 10^6 Q_s P_s Z T_1 / T_s (k/k - 1) [(P_2/P_1)^{(k-1)/k} - 1] / \eta , \quad (11.2)$$

where

bhp = compressor brake horsepower (hp), calculated to be 4,986 hp;

P_1 = pressure at compressor inlet (1,000 psia);

P_2 = pressure at compressor outlet (2,100 psia);

P_s = standard pressure (14.7 psia);

Q_s = flow rate at standard temperature and pressure (144.6 ft³/d);

T_1 = temperature at compressor inlet (122°F = 582°R);

T_s = standard temperature (520°R);

k = $C_p/C_v = 7.08/5.09 = 1.39$ (dimensionless at 100°C); and

η = compressor efficiency (dimensionless) (80%).

The compression ratio for the final stage of compression to pipeline pressures is not equal to the ratios used in the previous cases, and further optimization of the system is possible. The calculated brake horsepower directly converts to a power consumption of 3.72 MW. Because the power for this compressor comes from within the IGCC plant, this power consumption has been added to the sum of in-plant power use. The CO₂ emissions resulting from the use of this power are those coming from the IGCC system.

11.3 Substation Recompression

The length of the CO₂ pipeline is 500 km (310 mi). For practical reasons of access, substations will be located at 80-mi intervals (Segelar 1974, p. 8/29; Wolsky et al. 1985, p. 35). Hence, recompression stations will be at 80, 160, and 240 mi; and an access and control station will be at 310 mi, located in the disposal field. Because the pipeline pressure at this point (1,850 psia) is considerably higher than that required for charging the reservoir with CO₂, no further compression will be assumed after the third recompression station.

The allowed drop in pressure for a given flow rate balances pipe diameter and maximum allowable pressure against compressor operating costs. In the study by Wolsky et al. (1985), this pressure drop was found to vary from 2.6 to 4.6 psi/mi. For the purposes of these calculations, a typical value of 3.6 psi/mi was used. Hence, starting at 2,100 psia, the pressure at the first recompression station 80 mi away is 288 psi lower. Including substation handling and gas conditioning at 12 psi, the total pressure drop becomes 300 psi for recompression. Gas inlet temperature will be controlled at 100°C. Hence, by using the method outlined previously, the compression requirement for each substation is 924 hp, and the total power requirement for three substations is 2.066 MW, with an equivalent CO₂ emission of 2.066 kg/h.

11.4 Total Power Consumption for CO₂ Pipeline

The compression costs from the CO₂ recovery system through the booster stations to the sequestering zone differ only by the compression load that is required from the operating pressure of the CO₂ recovery system to the pipeline operating pressure of 2,100 psia. A summary of the power consumption appears in Table 11.1. The outputs of the CO₂ recovery system were normalized to a CO₂ value of 322,000 kg/h (the calculated CO₂ recoveries differed from this by a maximum of 3.5%). The low power consumption for the chilled methanol system is a consequence of its high operating pressure and low temperature.

11.5 Costs for CO₂ Pipeline: Compression and Dehydration

The costs for pipelines that could carry CO₂ from the IGCC plant have been developed for previous DOE studies (Hare et al. 1978; Sparrow et al. 1988) concerned with supplying CO₂ for enhanced oil recovery. New pipelines should consider supercritical operation because higher gas densities can be achieved and because two-phase flow is eliminated. The pressure-temperature combination selected for transmission is a function of pressurization and repressurization cost, the separation between compressors, and the pipeline's diameter, which, in turn, are functions of the quantity of CO₂ transported and the cost of energy.

The data for initial compression and dehydration are shown in Table 11.2. Costs for pipeline compression of the CO₂ are insensitive to the size of the plant, even over the order of magnitude from 24×10^6 to 240×10^6 standard cubic feet per day (scfd); here the economy of scale exhibits a decrease of only 8%. The CO₂ recovery from the IGCC system falls comfortably into this range at 142×10^6 scfd.

TABLE 11.1 Power Consumption for CO₂ Pipeline^a

Source	Consumption of Electricity (MW)			
	Glycol (Low Pressure)	Amine	Chilled Methanol	K ₂ CO ₃
CO ₂ recovery (2,100 psia)	33.04	38.63	21.96	38.38
Pipeline booster stations	2.07	2.07	2.07	2.07
Total	35.11	40.70	24.03	40.45

^a Pipeline/sequestering summary; basis, CO₂ sequestered in reservoir at 322,000 kg/h.

TABLE 11.2 Equipment for CO₂ Compression

Variable	Flow Rate of CO ₂ (10 ⁶ scfd)		
	240	48	24
Capital costs (10 ³ \$)			
Compressor	27.9	5.9	3.4
Dehydrator	4.4	1.6	1.1
Total	32.2	7.5	4.5
Horsepower (10 ³ hp)	110	22	11
Cost of CO ₂ (\$/10 ³ scf)	0.475	0.489	0.514

The projected costs for pipelines as a function of their anticipated capacity and lifetime appear in Table 11.3, again with consideration of the same range of flow rates. Here the economies of scale are very significant. The higher costs of pipe for the cases with higher flow rates are relatively minor costs in comparison with the fixed costs of the excavation. Table 11.4 is a summary of pipeline costs by mode (either supercritical or subcritical), capacity, and length.

The procedure for computing costs is described subsequently. Costs reported in Table 11.4 have been adjusted from 1978 to 1993 by using a 3%/yr escalation factor. Costs were computed for a natural supply of CO₂. The four types of reservoirs are as follows:

- T1 — 98% CO₂, 2% CH₄, and traces of H₂S;
- T2 — 96% CO₂, 25% CH₄, and 2% H₂S;
- T3 — 50% CO₂, 50% CH₄, and traces of H₂S; and
- T4 — 48% CO₂, 50% CH₄, and 25% H₂S.

The wellhead pressure is about 250 psig. The cost of recovering CO₂ is estimated at \$0.11/10³ scf for T3 and \$0.47/10³ scf for T4. The credit from the CH₄ by-product gas is \$1.75/10³ scf. The cost to remove trace amounts of H₂S is estimated to be \$0.35/10³ scf. The cost of compression and transmission (100 mi) at supercritical pressure of 2,200 psig ranges from a low of \$0.72/10³ scf (T1) to a high of \$0.89/10³ scf (T4).

TABLE 11.3 Pipeline Costs
by Capacity and Duration

Flow Rate (10 ⁶ scfd)	Cost (\$/[10 ³ scf·100 mi ⁻¹])	
	20 yr	10 yr
500	0.17	0.19
250	0.25	0.30
125	0.36	0.41
50	0.56	0.67

A sample transportation cost calculation follows:

$$\text{cost} = R/G/340 \times \$1,000/10^3 \text{ scf CO}_2 ,$$

and

$$R = P (C - W) + 48/52 \times P (C - W) + 0.05 (C + N) ,$$

where

P = annual percent return on rate base (%/yr),

C = total capital requirement (\$),

W = working capital (\$),

N = total net annual operating cost (\$/yr),

R = annual gas revenue requirement (\$/yr), and

G = CO₂ flow rate (scfd).

The assumptions are a 20-year project life, straight-line depreciation on total capital requirement excluding working capital, a 48% income tax rate, 100% equity capital (no debt), a new pipeline system, operating time of 340 d/yr, and a 3% investment per year for state and local taxes and insurance.

TABLE 11.4 Pipeline Costs by Mode, Capacity, and Length

Mode and Capacity (10 ⁶ scfd)	Length (mi)	Cost (\$/10 ³ scf) by Discounted Cash Flow			
		10%	20%	30%	40%
Supercritical					
50	50	0.19	0.26	0.34	0.42
	100	0.33	0.45	0.59	0.75
	300	0.95	1.29	1.58	2.34
125	50	0.11	0.14	0.19	0.23
	100	0.22	0.30	0.37	0.47
	300	0.61	0.81	1.04	1.29
	500	0.98	1.31	1.68	2.10
250	50	0.08	0.11	0.14	0.17
	100	0.16	0.30	0.37	0.47
	300	0.39	0.55	0.70	0.89
	500	0.67	0.92	1.18	1.50
500	50	0.06	0.08	0.11	0.13
	100	0.13	0.16	0.20	0.25
	300	0.37	0.48	0.61	0.75
	500	0.67	0.92	1.18	1.50
Subcritical					
50	50	0.36	0.56	0.56	0.69
	100	0.64	0.81	1.03	1.26
	300	1.63	2.23	2.90	3.66
125	50	0.25	0.31	0.37	0.45
	300	0.89	1.18	1.51	1.89
Liquid					
50	100	1.43	1.79	2.21	2.68

Operating costs are based on the following:

- Utility cost
 - Electric power at \$0.35/kWh (note that \$0.60/kWh would be the current basis),
 - Cooling water at \$0.02/1,000 gal,

- Fuel gas at \$2.00/10⁶ Btu,
- Glycol solvent (Selexol) at \$1.00/lb, and
- Steam (900 psig and 900°F) at \$2.80/1,000 lb (on basis of coal price of \$0.05/10⁶ Btu [lower heating value]);
- Indirect cost
 - Operating labor of one man per shift at \$9.00/h for 24 years,
 - Operating supplies at 30% of operating labor,
 - Supervision and overhead at 100% of operating labor,
 - Maintenance and labor at 3.5% of total plant investment cost, and
 - State and local taxes at 3.0% of total plant investment cost and insurance; and
- Total operating cost = utility cost + indirect cost.

11.6 Commercial Pipelines

In some cases, existing commercial pipelines could be used to transport CO₂, perhaps in a shared mode with other products. In other cases, construction of a new pipeline would be required. Costs for pipeline construction and use vary greatly on a regional basis within the United States.

Table 11.5 is a summary of the capital cost estimate for equipment for the CO₂ disposal pipeline. Remember that crossing a mountain range, crossing a river, or needing to pass through a metropolitan area with highways could substantially increase this cost. The lifetime of 10-20 years for a pipeline is considerably lower than for the IGCC plant (Wolsky et al. 1985). For this reason, a higher rate of capital recovery should be considered for these systems.

TABLE 11.5 Capital Cost Estimate for Equipment
for the 500-km CO₂ Disposal Pipeline

Equipment	Cost (\$)	Size (brake hp)
Pipeline feed compressor (1,000-2,100 psia)	2,883,289	4,986
Booster compressor 1 (at 80 mi)	622,122	924
Booster compressor 2 (at 160 mi)	622,122	924
Booster compressor 3 (at 240 mi)	622,122	924
Ancillary equipment	474,966	- ^a
Pipeline (500 km [310 mi])	140,100,000	-
Sum of capital	145,324,621	-
Sum of consumables	0	-
Total	145,324,621	-

^a Not applicable.

12 Sequestering of CO₂

12.1 Overview of Geologic Processes: The Dominance of Ocean Sequestering

The following distribution for the carbon on the earth (see Figure 12.1) was taken from a report by Dunsmore and was cited in a DOE study from the Energy Laboratory of Massachusetts Institute of Technology (DOE 1993, p. 2). The distribution of carbon demonstrates that, throughout geologic time, the sequestering cycle of significance continues to be that of the biological activity in the oceans to form carbonate rocks. Ocean activity has always dominated this process and continues to do so today. The overall sequence of sequestering can be represented as

If we take all of the carbonate rocks on earth to be the result of this oceanic biological activity and add in all of the sedimentary and fossil fuels as also being sequestered by biological activity, the average rate of sequestering since the appearance of life in the oceans is

$$\text{planetary sequestering rate of C} = 75 \times 10^6 \text{ Gt} / 3.5 \times 10^9 \text{ yr} = 0.02 \text{ Gt/yr} . \quad (12.2)$$

Current anthropogenic activity releases 5.9 gigatonnes (Gt) of carbon per year as CO₂ at a rate of 21.8 Gt/yr (Figure 12.2). Hence, today's rate of anthropogenic release of carbon exceeds the historical planetary sequestering rate of carbon by a factor of 280 and leads directly to increases in atmospheric CO₂.

The base-case IGCC system releases CO₂ at 379,000 kg/h. At a utilization rate of 65%, this amount comes to a carbon release of 0.59×10^9 kg/yr (from a CO₂ release of 2.16×10^9 kg/yr). By assuming that an acceptable offset to this impact will be approximately 90% of this value, carbon needs to be sequestered each year at a rate of 0.5×10^9 kg/yr. Over a 30-year life for the IGCC power station, this amount comes to 15×10^9 kg of carbon.

12.2 Terrestrial Biological Sequestering

Data for sequestering by tropical rain forests report that they sequester between 76 and 250 kg of carbon per hectare (1 ha = 2.47 acres = 0.00386 mi²) (Houghton 1986, p. 186). By assuming an average inventory of carbon for tropical rain forests of 163×10^3 kg/ha, the area of the rain forest that must be maintained in perpetuity is 355.2 mi². Washington, D.C., encompasses an area of approximately 61 mi². By using the Washington Monument as the center of a circle, a radius that could contain this rain forest would be just slightly larger than that enclosed by the famous I-495 beltway surrounding the city.

FIGURE 12.1 Distribution of Carbon on Earth

FIGURE 12.2 Current Release of Carbon and CO₂ Caused by Anthropogenic Activity

Microalgae have been proposed as a possible method for CO₂ capture. The use of ponds and today's technology would permit the sequestering of carbon at 66,000 kg/ha/yr (DOE 1993, p. 4-14). Potentially, this inventory of carbon could be doubled. Bear in mind that the pond must be sized for the daily emission rate during plant operations (i.e., the plant's emission rate for sequestering on a daily basis must not be averaged over the entire year; hence, rate = 0.5×10^9 kg C/yr/ $0.65 = 0.77 \times 10^9$ kg C/yr). Using a figure for carbon of 100,000 kg/ha/yr (an average of the current design limit and the theoretical limit) results in a pond of 30 mi² for the IGCC power station. Unless the biomass were to be returned to the process as a fuel, no scenario exists under which the dewatering and processing of wet microalgae would not result in a daily tonnage of material for disposal that exceeds the daily weight of coal consumed by the IGCC. If the fuel value of the microalgae were recovered, we would then be discussing a renewable biomass system, rather than a fossil system. The operating and annualized costs for this system were projected at \$376/10³ kg of carbon. (Some of the participants at the DOE conference that served as the basis for the reference observed that even the current generation of solar cells would prove to be a cost-effective competitor to a combined power plant-microalgae system.)

12.3 Ocean Sequestering

Ocean disposal is a serious competitor to terrestrial disposal. The huge size of the ocean is the major attraction of this concept because, if this option proves feasible, finding available capacity would no longer be a concern. If CO₂ were inert and the oceans were well mixed, predicting the consequence of ocean sequestering might be a relatively simple task (Baes 1982). Carbon dioxide is both chemically and biologically active in the oceans, and the oceans are stratified and not well mixed. The oceans receive practically all of their heat at the surface; and this effect, in combination with evaporation and precipitation, causes stable density gradients. Most water in the deep ocean moves along isopycnics by advection and, less significantly, in a vertical direction by turbulence.

Although, conceptually, a portion of the sequestered CO₂ would eventually, in time, reach the surface and escape into the atmosphere, the fraction released is unknown, as is the duration of time for the migration. The process is complicated; and attempts to theorize and model the process, especially on a global scale, are proving challenging.

With respect to depth, the ocean is essentially structured into three layers (Concorelli et al. 1991). Extending from the surface to about 100 m is a small mixed layer, where the density is nearly constant, the temperature varies linearly with depth, and atmospheric gas concentrations are at their equilibrium value. The thermocline layer (100 m to about 1,000 m) exhibits large temperature and density gradients, which effectively prevent vertical mixing. The deep ocean is at constant temperature. Seasonal variations in temperature only affect the mixed layer.

Horizontal convection in the deep ocean is a slow process. Convection completely around the globe would require centuries. Vertical convection is also a slow process and is further constricted by the aforementioned barriers between ocean layers. Estimates of the time scale for

CO₂ to migrate from a deep ocean sequestering to gradual release into the atmosphere range between 100 and 1,000 years (Baes 1982). To mitigate significant upward convection, it would be necessary to release CO₂ below 3,000 m, where it would be heavier than water, would sink to the bottom of the ocean, and would eventually dissolve in the water and thus start its upward transport journey from the ocean floor. Also, formation of a layer of liquid CO₂ will affect the bicarbonate ion formation occurring on the ocean floor, a location rich in CaCO₃. Because this reaction requires carbonates, CO₂, and water, whether the displacement of water by CO₂ on the ocean floor will promote or inhibit this reaction is not clear.

The injection of CO₂ into the mixed layer (near the shoreline) would be the best option — technically, economically, and politically (DOE 1993). Unfortunately, this technique is not effective because CO₂ in the surface waters would retard the natural absorption of CO₂ from the atmosphere. Also, solubility is very low; therefore, large amounts of water would be needed.

Below 500 m, a propensity exists for CO₂ and water to form hydrates (which are heavier than water and would therefore sink) (DOE 1993). The process of formation occurs at nonequilibrium conditions, and the effect of hydrodynamic mixing plays a significant role in the process of hydrate formation. Currently, reasonably predicting the extent of hydrate formation is not possible, even under ideally controlled conditions.

Basically, four ways exist to introduce CO₂ into the ocean (DOE 1993). Carbon dioxide can be mixed with seawater until dissolved, and the dissolved liquid can be pumped into the ocean, with this denser liquid presumably sinking. Carbon dioxide can be introduced at a relatively shallow depth (<500 m) as a gas, dissolve in the seawater as the CO₂ migrates toward the surface, and sink as heavier water (with dissolved CO₂). Carbon dioxide can be introduced as a solid and sink to the bottom, while any surface losses either take the form of hydrates or become dissolved in the surrounding water. (Making dry ice is expensive.) The preferred method is for CO₂ to be introduced as a liquid. Depending on conditions, having the liquid jet break up into tiny droplets that can readily be dissolved in water would be conceptually desirable. Presumably, a supercritical fluid is somehow included in this group. Even if the dumping could be ecologically localized, two significant impediments should be noted: (1) the effect on marine life is unknown, and (2) any action would require international agreements. That the ocean is not the domain of any nation is widely recognized (a similar belief with respect to our atmosphere can be argued, but historical precedents prevent nations from relinquishing claims to their air). Ocean disposal is not a near-term option; first, serious technical, economic, environmental, and political concerns need to be addressed.

12.4 Geologic Sequestering: Aquifers

An aquifer can be defined as a geologic formation with sufficient porosity and permeability and a large capability for natural containment. Considerable variation exists in the estimates of how much of the available volume in an aquifer CO₂ can occupy (van der Meer 1992). Also, the mobility ratio (the ratio of the permeability to the viscosity of CO₂ to that of water) varies greatly,

and this ratio can significantly affect the available pore volume. Regardless, the formation of acidic conditions caused by the interaction of CO₂ and freshwater is inevitable. The consequential damage as a result of this geochemical action is not well known.

The basic theory of storing CO₂ in underground repositories is based on fluid displacements in semiconfined regions. When large amounts of CO₂ are injected, some of it will dissolve with the water, while the remaining gas becomes immiscible with respect to the water. Buoyancy forces the gas to migrate upward; this movement is further enhanced by the increasing fluid pressure. Capillary pressure offers resistance to the CO₂ migration, and this effect tends to trap the gas under a rock cover.

Pressure buildup in aquifers can cause a shifting of the earth's surface above the damaged reservoir (van der Meer 1992). Chemical reaction between the acidic water and the geologic material could promote shifting. Also, leaks through new fissures, through now weakened but originally intact caps, and through other weak points in the encapsulation structure could lead to release of CO₂ into the atmosphere or groundwater.

Estimates of CO₂ terrestrial sedimentary basin storage capacity exceed 300×10^9 tons of CO₂ (Koide et al. 1992). Compared with the world total annual CO₂ emissions of just over 20×10^9 tons, sufficient storage capacity exists.

12.5 Geologic Sequestering: Depleted Oil and Gas Reservoirs

In analyzing depleted gas reservoirs, selecting a method of analysis that was simple, yet comprehensive enough to explain the technical aspects of this option, seemed appropriate. Computation of the extraction of gas from a gas reservoir requires some assumptions: the geometry of the reservoir, boundary conditions, the number of wells, injection sites, etc. To simplify the analysis, a finite cylindrical reservoir was assumed, with no-flow boundary conditions and with constant volumetric flow out of the reservoir and a single well in the center of the reservoir. The scoping calculations, which are detailed subsequently, show that this typical reservoir has ample capacity for the IGCC plant described in this report.

12.5.1 Natural Gas Outflow Calculations

For constant flow rate, and by ignoring well damage and overpressure drop, the reservoir dimensionless equations are as follows (Slider 1976):

$$\Delta(P^2) = \frac{1.424 \bar{\mu} \bar{z} T_f q_g}{h k} \left\{ \frac{2t_D}{r_{De}^2} + \ln r_{De} - \frac{3}{4} \right\}, \quad (12.3)$$

where

$$t_D = \frac{6.33kt}{\phi\mu cr_w^2} \quad (12.4)$$

and

$$r_{De} = \frac{r_e}{r_w} \quad (12.5)$$

Assume the following values for a typical gas reservoir (cylindrical geometry):

$$T_f = 600^\circ\text{R} \quad (\text{formation temperature});$$

$$k = 75 \text{ md} \quad (\text{permeability});$$

$$h = 200 \text{ ft} \quad (\text{cylindrical height});$$

$$\phi = 15\% \quad (\text{porosity});$$

$$r_w = 0.3 \text{ ft} \quad (\text{well radius});$$

$$r_e = 6,000 \text{ ft} \quad (\text{reservoir external radius}); \text{ and}$$

$$q_g = 5 \text{ MMscfd} \quad (\text{gas volumetric flow rate}).$$

By substituting into Equations 12.4 and 12.5, respectively, the typical values yield the following:

$$r_{De} = \frac{6,000}{0.3} = 2 \times 10^4$$

and

$$t_D = \frac{6.33(75 \times 10^{-3})}{(0.15)\mu c (0.3)^2} t = \frac{35.17t}{\mu c}.$$

With the further assumption that the initial reservoir pressure is 2,500 psi, the gas properties can now be computed (Reid et al. 1977). For gas composed of 80% CH₄, 15% ethane, and 5% butane, the critical gas constants are as follows:

$$T_c = 0.8 (190.7) + 0.15 (305.4) + 0.05 (425.2) = 219.63 \text{ K},$$

$$M_w = 0.8 (16.04) + 0.15 (30.07) + 0.05 (58.12) = 20.25 \text{ g/g-mol},$$

$$V_c = 0.8 (99.5) + 0.15 (148) + 0.05 (255) = 114.55 \text{ cm}^3/\text{g-mol},$$

$$z_c = 0.8 (0.290) + 0.15 (0.285) + 0.05 (0.274) = 0.289,$$

$$\omega = 0.8 (0.013) + 0.15 (0.105) + 0.05 (0.201) = 0.036,$$

$$P_c = 0.289 (82.0567) (219.63)/114.55 = 45.72 \text{ atm, and}$$

$$\xi = (219.63)^{1/6}/(20.25)^{1/2}/(45.72)^{2/3} = 0.0427,$$

where

T_c is the critical temperature,

M_w is the molecular weight,

V_c is the critical volume,

z_c is the critical compressibility factor,

ω is the acentric factor,

P_c is the critical pressure, and

ξ is a dimensionless parameter used in the estimation of viscosity.

The reduced variables (indicated by subscript r) are the following:

$$T_r = (140 + 460)/219.63/1.8 = 1.52;$$

$$P_r = 2,500/45.72/14.7 = 3.72;$$

$$z^0 = 0.79; z^1 = 0.2; z = 0.79 + 0.036(0.2) = 0.797;$$

$$V = (0.797) (82.0567) (600/1.8)/(2,500/14.7) = 128.18 \text{ cm}^3/\text{g-mol};$$

$$\rho_r = V_c/V = 0.894;$$

$$\mu^0 = 34 \times 10^{-5} (1.52)^{8/9}/0.0427 = 1.1537 \times 10^{-2} \text{ cp};$$

$$\Delta\mu = 10.8 \times 10^{-5} ((e^{(1.439(0.894))}) - (e^{-(1.111(0.894)^{1.858}})))/\xi \text{ cp}$$

$$= 10.8 \times 10^{-5} (3.62 - 0.41)/0.0427 = 8.130 \times 10^{-3} \text{ cp; and}$$

$$\mu = \mu^0 + \Delta\mu = 1.98 \times 10^{-2} \text{ cp},$$

where μ is the viscosity.

Compute the compressibility, c , from Figure B.5 of Slider (1976), as follows:

$$c_r = 0.25 \rightarrow c = \frac{0.25}{(45.72)(14.7)} = 3.7 \times 10^{-4} \text{ psi}^{-1} .$$

From Equation 12.4,

$$t_D = \frac{35.17t}{(1.98 \times 10^{-2})(3.7 \times 10^{-4})} = 4.825 \times 10^6 t .$$

Finally, from Equation 12.3,

$$\begin{aligned} (P^2) &= \frac{1.424 \bar{\mu} \bar{z} (600) (5 \times 10^3)}{(200) (75 \times 10^3)} \{0.024t + 9.15\} \\ &= 2.848 \times 10^5 \bar{\mu} \bar{z} \{0.024t + 9.15\} . \end{aligned}$$

To compute the average values of the viscosity and compressibility factor, assume that the reservoir is depleted at $P = 500$ psi, and compute the time to depletion, as follows:

$$\Delta(2,500^2 - 500^2) = 2.848 \times 10^5 (1.61 \times 10^{-2}) (0.86) \{0.024t + 9.15\}$$

$$6 \times 10^6 = 3.94 \times 10^3 \{0.024t + 9.15\}$$

$$t = 5,964 \text{ d (or 16.3 yr)} .$$

For this time, the total outflow is

$$Q = 5 \times 10^6 \text{ scfd (5,964 d)} = 2.982 \times 10^{10} \text{ scf} .$$

Now compute the final pressure required to fill this depleted reservoir with CO_2 as delivered from the pipeline.

12.5.2 Inflow Calculations

Inflow calculations were as follows:

$$R' = 1,545.33/144/44.01 = 0.244 \text{ psia-ft}^3/\text{lb}_m^\circ\text{R},$$

$$\rho = P/(zR'T) = 1,250/(0.80 \cdot 0.244 \cdot 530) = 8.06 \text{ lb}_m/\text{ft}^3,$$

$$q'_g = \dot{m}/\rho = 8 \times 10^5/(24)/8.06 = 2.4 \times 10^6 \text{ scfd},$$

$$Q' = 2.4 \times 10^6 (30) (365) = 2.63 \times 10^{10} \text{ scf},$$

$$T_c = 304.2 \text{ K},$$

$$M_w = 44.01 \text{ g/g-mol},$$

$$V_c = 94.0 \text{ cm}^3/\text{g-mol},$$

$$z_c = 0.274,$$

$$\omega = 0.420,$$

$$P_c = 0.274 (82.0567) (304.2)/94.0 = 72.76 \text{ atm, and}$$

$$\xi = (304.2)^{1/6}/(44.01)^{1/2}/(72.76)^{2/3} = 0.02243.$$

Assume that CO_2 is cooled to reservoir conditions:

$$T_r = 600/304.2/1.8 = 1.096;$$

$$P_r = 1,250/72.76/14.7 = 1.17;$$

$$z^0 = 0.618; z^1 = 0.083; z = 0.61 + 0.420 (0.083) = 0.64;$$

$$V = 0.64 (82.0567) (530/1.8)/(1,250/14.7) = 182 \text{ cm}^3/\text{g-mol};$$

$$\rho_r = 94.0/182 = 0.516;$$

$$\mu^0 = 34 \times 10^{-5} (1.096)^{8/9}/0.02243 = 1.64 \times 10^{-2} \text{ cp};$$

$$\Delta\mu = 10.8 \times 10^{-5} ((e^{(1.439(0.516))}) - (e^{-(1.111(0.516)^{1.858}})))/0.02243 = 6.6 \times 10^{-3} \text{ cp; and}$$

$$\mu = 2.3 \times 10^{-2} \text{ cp}.$$

Compute the properties at $P = 500$:

$$P_r = 500/72.76/14.7 = 0.467;$$

$$z^0 = 0.8737; z^1 = 0.006; z = 0.88;$$

$$V = 0.88 (82.0567) (530/1.8)/(500/14.7) = 625 \text{ cm}^3/\text{g-mol};$$

$$\rho_r = 94/625 = 0.15;$$

$$\Delta\mu = 1.3 \times 10^{-3} \text{ cp};$$

$$\mu = 1.77 \times 10^{-2} \text{ cp};$$

$$c_g = \frac{1}{P} - \frac{\Delta \cdot \dot{z}/z}{\Delta P}$$

$$= \frac{1}{1,250} - \frac{0.24/0.64}{(-750)} = 1.3 \times 10^{-3} \text{ psi}^{-1};$$

$$\bar{z} = \frac{1}{2} (z_{1,250} + z_{500}) = 0.76;$$

$$\bar{\mu} = \frac{1}{2} (\mu_{1,250} + \mu_{500}) = 2.035 \times 10^{-2} \text{ cp};$$

$$t_D = \frac{6.33 (75 \times 10^{-3}) (30) (365)}{(0.15) (2.035 \times 10^{-2}) (1.3 \times 10^{-3}) (0.3)^2} = 1.45 \times 10^{10}; \text{ and}$$

$$\Delta(P^2) = \frac{1.424 (2.035 \times 10^{-2}) (0.76) (530) (2.4 \times 10^3)}{(200) (75 \times 10^{-3})} \left\{ \frac{1.45 \times 10^{10} (2)}{(2 \times 10^4)^2} + 9.15 \right\}$$

$$= 1.87 \times 10^3 \{72.69 + 9.15\}$$

$$\Delta(P^2) = 1.53 \times 10^5 = P^2 - 500^2$$

$$P = 635 \text{ psi},$$

where c_g is gas compressibility.

Hence, an additional compressor is not needed; however, if, for another case, additional compression power was required, the power requirements would be very modest because the pressure ratio ($P/1,250$) would be very low.

12.6 Conclusions

Many options exist for sequestering CO₂ off-gas. All options seem to have problems; for example, deep ocean dumping is questionable with regard to maintaining the integrity of the sequestering (especially over a reasonable but lengthy period of time [e.g., two generations]), as well as the political difficulty in obtaining the international cooperation necessary to pursue this option. Sequestering in aquifers is questionable because of concerns about water contamination. Sequestering in depleted gas reservoirs is technically, environmentally, and politically feasible, given our present understanding of the long-range ramifications of each option. The major disadvantage of this option has been the limited availability of appropriate depositories; however, for the purposes of our study, and given the modest requirements for additional capacity (new power plants), this option seems to be the best choice.

13 Conclusions

13.1 Summary of Energy Consumption and CO₂ Emissions

13.1.1 Base-Case IGCC

The IGCC base case without CO₂ controls yields the energy consumption and CO₂ emissions appearing in Table 13.1. The IGCC plant delivers 493.8 MW of power (gross). The fuel-cycle power use yields a net power of 454.4 MW at a CO₂ emission rate of 0.835 kg/kWh. This rate is significantly below the typical CO₂ emission rate of 1 kg/kWh for the installed electric power generating capacity used throughout the study. To summarize the ranking of the CO₂ recovery systems with respect to favorable fuel-cycle CO₂ sequestering costs: glycol > K₂CO₃ > amine > chilled methanol.

13.1.2 IGCC/Shift/Amine CO₂ Recovery Case

Table 13.2 surveys the base-case IGCC system with shift and conventional amine-based CO₂ recovery. The IGCC/CO₂ recovery plant delivers 395.5 MW of power (gross). Both the gas turbine and the steam turbine have been significantly derated, largely as a consequence of the high steam demands for amine regeneration. Additionally, the in-plant use results in a fuel-cycle net power yield of 310.5 MW at a CO₂ emission rate of 0.19 kg/kWh. This rate is a 77% reduction in emission rates from the base-case IGCC system, accomplished with a net derating of nearly 144 MW. With the design set of assumptions, the energy costs of CO₂ sequestering come to 0.449 kWh/kg of CO₂, a value derived by dividing this net derating of 144 MW by the net decrease in CO₂ released to the atmosphere in the base case. The impacts of purchasing makeup power so as to match the base-case net power of 454.4 MW do not appear in this table.

13.1.3 IGCC/Shift/Glycol CO₂ Recovery Case

Table 13.3 surveys the base-case IGCC system with shift and low-pressure glycol CO₂ recovery. This case represents the best CO₂ recovery performance observed in this survey. The IGCC/CO₂ recovery plant delivers 473.0 MW of power (gross). The gas turbine has undergone modest derating, while the steam turbine has actually increased its production with respect to the base case. In-plant use results in a fuel-cycle net power yield of 380.8 MW at a CO₂ emission rate of 0.156 kg/kWh. This yield is accomplished with a net derating of 73.6 MW. The energy costs for CO₂ sequestering come to 0.23 kWh/kg of CO₂. Once again, the impacts of purchasing makeup power so as to match the base-case net power of 454.4 MW do not appear in this table.

TABLE 13.1 Energy Consumption and CO₂ Emissions
for KRW IGCC Base Case

Source	Electricity (MW)	CO ₂ Release (kg/h)
Mining and transport		
Raw coal in mine	2.42	2,418
Coal rail transport	0.10	637
Limestone mining	0.26	257
Limestone rail transport	0.02	140
Subtotal	2.80	3,452
IGCC plant		
Coal preparation	2.35	13,099
Limestone preparation	1.23	- ^a
Gasifier island	21.66	158
Power island	11.39	362,820
Subtotal	36.63	376,077
Pipeline/sequester	0.00	0
Total	39.43	379,529
Power from gas turbine	311.60	-
Power from steam turbine	182.20	-
Gross power	493.80	-
Net cycle energy	454.37	-
CO ₂ emission rate	-	0.8353 ^b
CO ₂ sequestering	N/A ^c	-

^a Not applicable.

^b 0.8353 kg/kWh.

^c N/A, Not available.

13.1.4 IGCC/Shift/Methanol CO₂ Recovery Case

Table 13.4 surveys the base-case IGCC system with shift and chilled methanol CO₂ recovery. The chilled methanol recovery is a highly effective technique for removing even trace concentrations of contaminants from the gasifier product. For this reason, this technique has been chosen as the gas cleanup system for the Great Plains Gasification Project (Beulah, North Dakota), where high cleaning standards are necessary to protect costly methanation catalyst. In this survey, the methanol technique represents the costliest system. The IGCC/CO₂ recovery plant still delivers 439.1 MW of power (gross), with the gas turbine significantly derated, while the steam turbine has undergone slight derating. In-plant use of energy is exceptionally high to provide for the

TABLE 13.2 Energy Consumption and CO₂ Emissions
for KRW IGCC with Amine CO₂ Recovery

Source	Electricity (MW)	CO ₂ Release (kg/h)
Mining and transport		
Raw coal in mine	2.42	2,418
Coal rail transport	0.10	637
Limestone mining	0.26	257
Limestone rail transport	0.02	140
Subtotal	2.80	3,452
IGCC plant		
Coal preparation	2.35	13,099
Limestone preparation	1.23	- ^a
Gasifier island	21.66	158
Power island	11.39	362,820
CO ₂ recovery	4.83	-328,919
CO ₂ compression (2,100 psi)	38.67	-
Subtotal	80.13	47,158
Pipeline/sequester		
Pipeline CO ₂	-	328,919
Pipeline booster stations	2.07	2,071
Geologic reservoir (2% loss)	0.00	-322,341
Subtotal	2.07	8,649
Total	85.00	59,259
Power from gas turbine	282.10	-
Power from steam turbine	113.40	-
Gross power	395.50	-
Net cycle energy	310.50	-
Decrease from base KRW	-143.87	320,270
CO ₂ emission rate	-	0.1909 ^b
CO ₂ sequestering	0.4492 ^c	-

^a Not applicable.

^b 0.1909 kg/kWh.

^c 0.4492 kWh/kg of CO₂.

TABLE 13.3 Energy Consumption and CO₂ Emissions
for KRW IGCC with Low-Pressure Glycol CO₂ Recovery

Source	Electricity (MW)	CO ₂ Release (kg/h)
Mining and transport		
Raw coal in mine	2.42	2,418
Coal rail transport	0.10	637
Limestone mining	0.26	257
Limestone rail transport	0.02	140
Subtotal	2.80	3,452
IGCC plant		
Coal preparation	2.35	13,099
Limestone preparation	1.23	- ^a
Gasifier island	21.66	158
Power island	11.39	362,820
CO ₂ recovery	17.66	-328,810
CO ₂ compression (2,100 psi)	33.06	-
Subtotal	87.35	47,267
Pipeline/sequester		
Pipeline CO ₂	-	328,810
Pipeline booster stations	2.07	2,066
Geologic reservoir (2% loss)	0.00	-322,234
Subtotal	2.07	8,642
Total	92.22	59,361
Power from gas turbine	281.60	-
Power from steam turbine	191.40	-
Gross power	473.00	-
Net cycle energy	380.78	-
Decrease from base KRW	-73.59	320,168
CO ₂ emission rate	-	0.1559 ^b
CO ₂ sequestering	0.2298 ^c	-

^a Not applicable.

^b 0.1559 kg/kWh.

^c 0.2298 kWh/kg of CO₂.

TABLE 13.4 Energy Consumption and CO₂ Emissions
for KRW IGCC with Chilled Methanol CO₂ Recovery

Source	Electricity (MW)	CO ₂ Release (kg/h)
Mining and transport		
Raw coal in mine	2.42	2,418
Coal rail transport	0.10	637
Limestone mining	0.26	257
Limestone rail transport	0.02	140
Subtotal	2.80	3,452
IGCC plant		
Coal preparation	2.35	13,099
Limestone preparation	1.23	- ^a
Gasifier island	21.66	158
Power island	11.39	362,820
CO ₂ recovery	105.60	-
CO ₂ compression (2,100 psi)	22.75	-333,462
Subtotal	164.98	42,615
Pipeline/sequester		
Pipeline CO ₂	-	333,462
Pipeline booster stations	2.10	2,099
Geologic reservoir (2% loss)	0.00	-326,793
Subtotal	2.10	8,769
Total	169.88	54,836
Power from gas turbine	269.00	-
Power from steam turbine	170.10	-
Gross power	439.10	-
Net cycle energy	269.22	-
Decrease from base KRW	-185.15	324,693
CO ₂ emission rate	-	0.2037 ^b
CO ₂ sequestering	0.5702 ^c	-

^a Not applicable.

^b 0.2037 kg/kWh.

^c 0.5702 kWh/kg of CO₂.

refrigeration of the methanol to -40°C . Consequently, the in-plant use results in a fuel-cycle net power yield of 29.2 MW at a CO_2 emission rate of 0.204 kg/kWh. This emission rate is nearly identical to that in the glycol case. This net derating of 185 MW is the highest observed. With the design set of assumptions, the energy costs of CO_2 sequestering come to 0.57 kWh/kg of CO_2 . The impacts of purchasing makeup power so as to match the base-case net power of 454.4 MW do not appear in this table.

13.1.5 IGCC/Shift/Carbonate CO_2 Recovery Case

Table 13.5 surveys the base-case IGCC system with shift and hot potassium carbonate CO_2 recovery. The K_2CO_3 system lives up to its promise of being a very energy-competitive system for CO_2 recovery; however, an analysis of the total energy cycle shows that this system has a performance lower than that of the glycol case. The IGCC/ CO_2 recovery system now delivers 420.3 MW of power (gross). The gas turbine has undergone modest derating, while the steam turbine undergoes nearly 30% derating with respect to the base case. In-plant use results in a fuel-cycle net power yield of 328.2 MW at a CO_2 emission rate of 0.18 kg/kWh; hence, the net derating is 126 MW. The energy costs for CO_2 sequestering come to 0.39 kWh/kg of CO_2 . Once again, the impacts of purchasing makeup power so as to match the base-case net power of 454.4 MW do not appear in this table.

13.2 Summary of Economics

13.2.1 Direct Costs for IGCC/Shift/ CO_2 Recovery

Summaries of the economics for the IGCC CO_2 recovery systems appear in Tables 13.6 through 13.9. A levelized fixed charge of 20% is taken for the capital on the CO_2 recovery systems; and in a similar way, labor and maintenance have been included as fixed percentages of the capital. A somewhat higher levelized fixed charge of 23% was considered appropriate for the pipeline. To this levelized fixed charge are added the annual costs of consumables and chemicals. Costs for power are taken consistently at 60 mills/kWh during the operation of the plant. The on-stream factor for the IGCC CO_2 recovery system was taken to be 65%, as in the base case. Hence, no derating of plant availability is assumed as a consequence of the addition of shift and CO_2 recovery. The four systems surveyed fell in the range of \$19-\$46/tonne (t) (metric ton) of CO_2 . The amine CO_2 recovery system represents the lowest cost option, while the chilled methanol represents the upper range of capital and operating costs. (Note that the amine system is not the lowest cost option when the cost of replacement power is considered.)

TABLE 13.5 Energy Consumption and CO₂ Emissions
for KRW IGCC with Hot Potassium Carbonate
CO₂ Recovery

Source	Electricity (MW)	CO ₂ Release (kg/h)
Mining and transport		
Raw coal in mine	2.42	2,418
Coal rail transport	0.10	637
Limestone mining	0.26	257
Limestone rail transport	0.02	140
Subtotal	2.80	3,452
IGCC plant		
Coal preparation	2.35	13,099
Limestone preparation	1.23	- ^a
Gasifier island	21.66	158
Power island	11.39	362,820
CO ₂ recovery	12.14	-328,919
CO ₂ compression (2,100 psi)	38.42	-
Subtotal	87.19	47,158
Pipeline/sequester		
Pipeline CO ₂	-	328,919
Pipeline booster stations	2.07	2,071
Geologic reservoir	0.00	-322,341
Subtotal	2.07	8,649
Total	92.06	59,259
Power from gas turbine	290.00	-
Power from steam turbine	130.30	-
Gross power	420.30	-
Net cycle energy	328.24	-
Decrease from base KRW	-126.13	320,270
CO ₂ emission rate	-	0.1805 ^b
CO ₂ sequestering	0.3938 ^c	-

^a Not applicable.

^b 0.1805 kg/kWh.

^c 0.3938 kWh/kg of CO₂.

TABLE 13.6 Economics of Shift/MEA CO₂ Recovery

Factor	No. ^a	Cost (\$)	Change in Power (MW)	CO ₂ Emissions (kg/h)	CO ₂ Sequestering (kg/h)
Shift reactor (1 of 4 trains)					
Equipment					
Heat exchanger before 1st stage for steam	1	96,569	- ^b	-	-
Heat exchanger before 1st stage for reboiler	2	79,775	-	-	-
Heat exchanger after 1st stage for water	3	167,946	-	-	-
Heat exchanger after 1st stage for reboiler	4	134,357	-	-	-
Heat exchanger after 2nd stage for reboiler	5	377,879	-	-	-
Heat exchanger after 2nd stage for fuel gas	6	1,762,202	-	-	-
Heat exchanger after 2nd stage: gas-liquid	7	251,920	-	-	-
1st-stage shift reactor	8	30,916	-	-	-
2nd-stage shift reactor	9	17,175	-	-	-
Sum of capital		2,918,739	-	-	-
Consumables					
Shift catalyst	10	134,647	-	-	-
Sum of consumables	-	134,647	-	-	-
MEA process (1 of 4 trains)					
Equipment					
CO ₂ absorption column	11a	524,742	-	-	-
CO ₂ stripping column	12a	757,597	-	-	-
Heat exchanger for lean solvent cooling	13	870,740	-	-	-
Interstage cooler for absorber	14	288,320	-	-	-
Lean-rich MEA heat exchanger	15	396,440	-	-	-
Stripper overhead condenser	16	324,360	-	-	-
Stripper reboiler	17	747,186	-	-	-
Solvent circulation pump	18	191,700	-	-	-
CO ₂ product: stage 1 compressor	19a	2,342,600	-	-	-
CO ₂ product: stage 2 compressor	19b	2,196,188	-	-	-
CO ₂ product: stage 3 compressor	19c	2,196,188	-	-	-
Sum of capital	-	10,836,061	-	-	-
Consumables					
Packing	11b	258,645	-	-	-
Packing	12b	505,165	-	-	-
Sum of consumables	-	763,810	-	-	-

TABLE 13.6 (Cont.)

Factor	No. ^a	Cost (\$)	Change in Power (MW)	CO ₂ Emissions (kg/h)	CO ₂ Sequestering (kg/h)
Summary of shift/MEA (1 of 4 trains)					
Sum of capital	-	13,754,800	-	-	-
Sum of consumables	-	898,456	-	-	-
Total	-	14,653,256	-	-	-
Summary of shift/MEA (4 of 4 trains)					
Sum of capital	-	55,019,200	-	-	-
Sum of consumables	-	3,593,824	-	-	-
Total	-	58,613,024	-	-	-
Pipeline (500 km)					
Pipeline feed compressor (1,000-2,100 psia) ^c	-	2,883,289	-	-	-
Booster compressor 1 (at 80 mi) ^d	-	622,122	-	-	-
Booster compressor 2 (at 160 mi) ^d	-	622,122	-	-	-
Booster compressor 3 (at 240 mi) ^d	-	622,122	-	-	-
Ancillary equipment	-	474,966	-	-	-
Pipeline (500 km [310 mi])	-	140,100,000	-	-	-
Sum of capital	-	145,324,621	-	-	-
Sum of consumables	-	0	-	-	-
Total	-	145,324,621	-	-	-
Cost summary on annual basis					
Shift/MEA CO ₂ recovery					
Levelized fixed charge at 20%	-	11,003,840	-	-	-
Consumables	-	3,593,824	-	-	-
Amine solution	-	28,708	-	-	-
Power at 60 mills/kWh	-	13,590,439	-	-	-
Labor (4%)	-	2,200,768	-	-	-
Maintenance (7%)	-	3,851,344	-	-	-
Subtotal	-	34,268,924	-	-	-
Cost (\$/t of CO ₂)	-	19	-	-	-
Cost (\$/1,000 scf of CO ₂)	-	1.05	-	-	-
Cost (\$/t of C)	-	69	-	-	-
Pipeline (500 km)					
Levelized fixed charge at 23%	-	33,424,663	-	-	-
Consumables	-	0	-	-	-
Power at 60 mills/kWh	-	1,976,729	-	-	-
Labor (4%)	-	5,812,985	-	-	-
Maintenance (7%)	-	10,172,724	-	-	-
Subtotal	-	51,387,100	-	-	-
Cost (\$/t of CO ₂)	-	28	-	-	-
Cost (\$/1,000 scf of CO ₂)	-	1.57	-	-	-
Cost (\$/t of C)	-	103	-	-	-

TABLE 13.6 (Cont.)

Factor	No. ^a	Cost (\$)	Change in Power (MW)	CO ₂ Emissions (kg/h)	CO ₂ Sequestering (kg/h)
Replacement power at 60 mills/kWh					
IGCC with shift/MEA-net life cycle	-	-	310.5	59,259	322,341
Cost: base case - derated production	-	-	143.87	143,870	0
Sum	-	49,151,747	454.37	203,129	322,341
Cost (\$/t of CO ₂)	-	27	-	-	-
Cost (\$/1,000 scf of CO ₂)	-	1.50	-	-	-
Cost (\$/t of C)	-	99	-	-	-
Total cost	-	134,807,771	-	-	-
Total cost (\$/t of CO ₂)	-	73	-	-	-
Cost (\$/1,000 scf of CO ₂)	-	4.11	-	-	-
Total cost (\$/t of C)	-	269	-	-	-

^a Numbers refer to Table 5.5.

^b Not applicable.

^c Size of compressor, 4,986 brake hp.

^d Size of compressor, 924 brake hp.

13.2.2 Direct Costs for Pipeline

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

TABLE 13.7 Economics of Shift/Low-Pressure Glycol CO₂ Recovery

Factor	No. ^a	Cost (\$)	Change in Power (MW)	CO ₂ Emissions (kg/h)	CO ₂ Sequestering (kg/h)
Shift reactor (1 of 4 trains)					
Equipment					
Heat exchanger before 1st stage for steam	1	96,569	- ^b	-	-
Heat exchanger before 1st stage for fuel gas	2	514,823	-	-	-
Heat exchanger after 1st stage for water	3	167,946	-	-	-
Heat exchanger after 1st stage for fuel gas	4	677,897	-	-	-
Heat exchanger after 2nd stage for fuel gas	5	2,078,687	-	-	-
Heat exchanger for raw gas cooling	6	679,401	-	-	-
1st-stage shift reactor	7	30,916	-	-	-
2nd-stage shift reactor	8	17,175	-	-	-
Sum of capital	-	4,263,414	-	-	-
Consumables					
Shift catalyst	9	135,000	-	-	-
Sum of consumables	-	135,000	-	-	-
Glycol process (1 of 4 trains)					
Equipment					
CO ₂ absorption column	10a	674,669	-	-	-
Power recovery turbine 1	11	253,406	-	-	-
Power recovery turbine 2	12	197,094	-	-	-
Slump tank	13	163,982	-	-	-
Recycle compressor	14	169,839	-	-	-
Flash tank 1	15	163,982	-	-	-
Flash tank 2	16	163,982	-	-	-
Flash tank 3	17	163,982	-	-	-
Flash gas compressor 1	18	248,901	-	-	-
Flash gas compressor 2	19	819,910	-	-	-
Solvent circulation pump	20	260,988	-	-	-
CO ₂ product: stage 1 compressor	21a	2,049,775	-	-	-
CO ₂ product: stage 2 compressor	21b	1,903,363	-	-	-
CO ₂ product: stage 3 compressor	21c	1,903,363	-	-	-
Solvent refrigeration	22	616,622	-	-	-
Sum of capital	-	9,753,838	-	-	-
Consumables					
Packing	10b	519,085	-	-	-
Sum of consumables	-	519,085	-	-	-

TABLE 13.7 (Cont.)

Factor	No. ^a	Cost (\$)	Change in Power (MW)	CO ₂ Emissions (kg/h)	CO ₂ Sequestering (kg/h)
Summary of shift/glycol (1 of 4 trains)					
Sum of capital	-	14,016,917	-	-	-
Sum of consumables	-	654,085	-	-	-
Total	-	14,671,002	-	-	-
Summary of shift/glycol (4 of 4 trains)					
Sum of capital	-	56,067,660	-	-	-
Sum of consumables	-	2,616,340	-	-	-
Total	-	58,684,000	-	-	-
Pipeline (500 km)					
Pipeline feed compressor (1,000-2,100 psia) ^c	-	2,883,289	-	-	-
Booster compressor 1 (at 80 mi) ^d	-	622,122	-	-	-
Booster compressor 2 (at 160 mi) ^d	-	622,122	-	-	-
Booster compressor 3 (at 240 mi) ^d	-	622,122	-	-	-
Ancillary equipment	-	474,966	-	-	-
Pipeline (500 km [310 mi])	-	140,100,000	-	-	-
Sum of capital	-	145,324,621	-	-	-
Sum of consumables	-	0	-	-	-
Total	-	145,324,621	-	-	-
Cost summary on annual basis					
Shift/low-pressure glycol CO ₂ recovery					
Levelized fixed charge at 20%	-	11,213,532	-	-	-
Consumables	-	2,616,340	-	-	-
Glycol solution	-	170,949	-	-	-
Power at 60 mills/kWh	-	16,057,080	-	-	-
Labor (4%)	-	2,251,773	-	-	-
Maintenance (7%)	-	3,924,736	-	-	-
Subtotal	-	36,225,343	-	-	-
Cost (\$/t of CO ₂)	-	20	-	-	-
Cost (\$/1,000 scf of CO ₂)	-	1.11	-	-	-
Cost (\$/t of C)	-	73	-	-	-

TABLE 13.7 (Cont.)

Factor	No. ^a	Cost (\$)	Change in Power (MW)	CO ₂ Emissions (kg/h)	CO ₂ Sequestering (kg/h)
Pipeline (500 km)					
Levelized fixed charge at 23%	-	33,424,663	-	-	-
Consumables	-	0	-	-	-
Power at 60 mills/kWh	-	1,976,729	-	-	-
Labor (4%)	-	5,812,985	-	-	-
Maintenance (7%)	-	10,172,724	-	-	-
Subtotal	-	51,387,100	-	-	-
Cost (\$/t of CO ₂)	-	28	-	-	-
Cost (\$/1,000 scf of CO ₂)	-	1.57	-	-	-
Cost (\$/t of C)	-	103	-	-	-
Replacement power at 60 mills/kWh					
IGCC with shift/glycol-net life cycle	-	-	380.78	59,361	322,264
Cost: base case - derated production	-	-	73.59	73,590	0
Sum	-	25,141,288	454.37	132,951	322,264
Cost (\$/t of CO ₂)	-	14	-	-	-
Cost (\$/1,000 scf of CO ₂)	-	0.77	-	-	-
Cost (\$/t of C)	-	51	-	-	-
Total cost	-	112,753,731	-	-	-
Total cost (\$/t of CO ₂)	-	61	-	-	-
Cost (\$/1,000 scf of CO ₂)	-	3.44	-	-	-
Total cost (\$/t of C)	-	225	-	-	-

^a Numbers refer to Table 6.8.

^b Not applicable.

^c Size of compressor, 4,986 brake hp.

^d Size of compressor, 924 brake hp.

13.2.3 Costs for Replacement Electricity

The costs of replacement electricity have been included and charged at a rate of 60 mills/kWh. This replacement cost is taken as the fuel-cycle derating of the IGCC CO₂ recovery system. This derating is simply the difference between the base-case fuel-cycle electric production of 454.4 MW and the fuel-cycle costs for each process. Note that the power consumption already counted against the shift/CO₂ recovery system, as well as pipeline power use, appears here. When replacement costs for electricity are considered, the chilled methanol system, with a cost of \$35/t of

TABLE 13.8 Economics of Shift/Chilled Methanol CO₂ Recovery

Factor	No. ^a	Cost (\$)	Change in Power (MW)	CO ₂ Emissions (kg/h)	CO ₂ Sequestering (kg/h)
Shift reactor (1 of 4 trains)					
Equipment					
Heat exchanger before 1st stage for steam	1	96,569	- ^b	-	-
Heat exchanger before 1st stage for fuel gas	2	553,029	-	-	-
Heat exchanger after 1st stage for water	3	167,946	-	-	-
Heat exchanger after 1st stage for fuel gas	4	772,229	-	-	-
Heat exchanger after 2nd stage for fuel gas	5	1,752,624	-	-	-
Heat exchanger for raw gas cooling (CO ₂ rich)	6a	502,562	-	-	-
Heat exchanger for raw gas cooling (CO ₂ lean)	6b	1,034,578	-	-	-
1st-stage shift reactor	7	30,916	-	-	-
2nd-stage shift reactor	8	17,175	-	-	-
Sum of capital	-	4,927,628	-	-	-
Consumables					
Shift catalyst	9	134,647	-	-	-
Sum of consumables	-	134,647	-	-	-
Methanol process (1 of 4 trains)					
Equipment					
CO ₂ absorption column	10a	826,469	-	-	-
CO ₂ stripping column	11a	927,670	-	-	-
Flash tank 1	12	65,593	-	-	-
Recycle compressor	13	234,260	-	-	-
Flash tank 2	14	65,593	-	-	-
Flash tank 3	15	65,593	-	-	-
Flash gas compressor 1	16	937,040	-	-	-
Flash gas compressor 2	17	1,581,255	-	-	-
Solvent circulation pump	18	98,263	-	-	-
Lean-rich solvent heat exchanger	19	881,033	-	-	-
CO ₂ product: stage 1 compressor	20a	1,581,233	-	-	-
CO ₂ product: stage 2 compressor	20b	1,317,713	-	-	-
CO ₂ product: stage 3 compressor	20c	1,317,713	-	-	-
Solvent refrigeration	21	6,330,651	-	-	-
Sum of capital	-	16,230,079	-	-	-
Consumables					
Absorption column packing	10b	648,857	-	-	-
Stripping column packing	11b	735,371	-	-	-
Sum of consumables	-	1,384,228	-	-	-

TABLE 13.8 (Cont.)

Factor	No. ^a	Cost (\$)	Change in Power (MW)	CO ₂ Emissions (kg/h)	CO ₂ Sequestering (kg/h)
Summary of shift/methanol (1 of 4 trains)					
Sum of capital	-	21,157,727	-	-	-
Sum of consumables	-	1,518,875	-	-	-
Total	-	22,676,602	-	-	-
Summary of shift/methanol (4 of 4 trains)					
Sum of capital	-	84,630,906	-	-	-
Sum of consumables	-	6,075,500	-	-	-
Total	-	90,706,407	-	-	-
Pipeline (500 km)					
Pipeline feed compressor (1,000-2,100 psia) ^c	-	2,883,289	-	-	-
Booster compressor 1 (at 80 mi) ^d	-	622,122	-	-	-
Booster compressor 2 (at 160 mi) ^d	-	622,122	-	-	-
Booster compressor 3 (at 240 mi) ^d	-	622,122	-	-	-
Ancillary equipment	-	474,966	-	-	-
Pipeline (500 km [310 mi])	-	140,100,000	-	-	-
Sum of capital	-	145,324,621	-	-	-
Sum of consumables	-	0	-	-	-
Total	-	145,324,621	-	-	-
Cost summary on annual basis					
Shift/chilled methanol CO ₂ recovery					
Levelized fixed charge at 20%	-	16,926,181	-	-	-
Consumables	-	6,075,500	-	-	-
Methanol	-	51,079	-	-	-
Power at 60 mills/kWh	-	53,873,212	-	-	-
Labor (4%)	-	3,385,236	-	-	-
Maintenance (7%)	-	5,924,163	-	-	-
Subtotal	-	86,235,371	-	-	-
Cost (\$/t of CO ₂)	-	47	-	-	-
Cost (\$/1,000 scf of CO ₂)	-	2.63	-	-	-
Cost (\$/t of C)	-	173	-	-	-

TABLE 13.8 (Cont.)

Factor	No. ^a	Cost (\$)	Change in Power (MW)	CO ₂ Emissions (kg/h)	CO ₂ Sequestering (kg/h)
Pipeline (500 km)					
Levelized fixed charge at 23%	-	33,424,663	-	-	-
Consumables	-	0	-	-	-
Power at 60 mills/kWh	-	1,976,729	-	-	-
Labor (4%)	-	5,812,985	-	-	-
Maintenance (7%)	-	10,172,724	-	-	-
Subtotal	-	51,387,100	-	-	-
Cost (\$/t of CO ₂)	-	28	-	-	-
Cost (\$/1,000 scf of CO ₂)	-	1.57	-	-	-
Cost (\$/t of C)	-	103	-	-	-
Replacement power at 60 mills/kWh					
IGCC with shift/methanol-net life cycle	-	-	269.22	54,836	326,793
Cost: base case - derated production	-	-	185.15	185,150	0
Sum	-	63,254,646	454.37	239,986	326,793
Cost (\$/t of CO ₂)	-	35	-	-	-
Cost (\$/1,000 scf of CO ₂)	-	1.93	-	-	-
Cost (\$/t of C)	-	127	-	-	-
Total cost	-	200,877,118	-	-	-
Total cost (\$/t of CO ₂)	-	108	-	-	-
Cost (\$/1,000 scf of CO ₂)	-	6.13	-	-	-
Total cost (\$/t of C)	-	396	-	-	-

^a Numbers refer to Table 7.5.

^b Not applicable.

^c Size of compressor, 4,986 brake hp.

^d Size of compressor, 924 brake hp.

TABLE 13.9 Economics of Shift/Hot Potassium Carbonate CO₂ Recovery

Factor	No. ^a	Cost (\$)	Change in Power (MW)	CO ₂ Emissions (kg/h)	CO ₂ Sequestering (kg/h)
Shift reactor (1 of 4 trains)					
Equipment					
Heat exchanger before 1st stage for steam	1	96,569	- ^b	-	-
Heat exchanger after 1st stage for water	2	167,946	-	-	-
Heat exchanger after 2nd stage for fuel gas	3	2,769,444	-	-	-
1st-stage shift reactor	4	30,916	-	-	-
2nd-stage shift reactor	5	17,175	-	-	-
Sum of capital	-	3,082,050	-	-	-
Consumables					
Shift catalyst	6	134,647	-	-	-
Sum of consumables	-	134,647	-	-	-
Carbonate process (1 of 4 trains)					
Equipment					
CO ₂ absorption column	7a	459,150	-	-	-
CO ₂ stripping column	8a	721,521	-	-	-
Heat exchanger for lean solvent cooling	9	209,933	-	-	-
Stripper overhead condenser	10	2,632,416	-	-	-
Solvent circulation pump	11	398,077	-	-	-
CO ₂ product: stage 1 compressor	12a	2,342,600	-	-	-
CO ₂ product: stage 2 compressor	12b	2,196,188	-	-	-
CO ₂ product: stage 3 compressor	12c	2,196,188	-	-	-
Sum of capital	-	11,156,073	-	-	-
Consumables					
Absorption column packing	7b	143,691	-	-	-
Stripping column packing	8b	366,413	-	-	-
Sum of consumables	-	510,104	-	-	-
Summary of shift/carbonate (1 of 4 trains)					
Sum of capital	-	14,238,122	-	-	-
Sum of consumables	-	644,751	-	-	-
Total	-	14,882,873	-	-	-
Summary of shift/carbonate (4 of 4 trains)					
Sum of capital	-	56,952,488	-	-	-
Sum of consumables	-	2,579,004	-	-	-
Total	-	59,531,493	-	-	-

TABLE 13.9 (Cont.)

Factor	No. ^a	Cost (\$)	Change in Power (MW)	CO ₂ Emissions (kg/h)	CO ₂ Sequestering (kg/h)
Pipeline (500 km)					
Pipeline feed compressor (1,000-2,100 psia) ^c	1	2,883,289	-	-	-
Booster compressor 1 (at 80 mi) ^d	2	622,122	-	-	-
Booster compressor 2 (at 160 mi) ^d	3	622,122	-	-	-
Booster compressor 3 (at 240 mi) ^d	4	622,122	-	-	-
Ancillary equipment	5	474,966	-	-	-
Pipeline (500 km [310 mi])	6	140,100,000	-	-	-
Sum of capital	-	145,324,621	-	-	-
Sum of consumables	-	0	-	-	-
Total	-	145,324,621	-	-	-
Cost summary on annual basis					
Shift/carbonate CO ₂ recovery					
Levelized fixed charge at 20%	-	11,390,498	-	-	-
Consumables	-	2,579,004	-	-	-
Potassium carbonate	-	50,000	-	-	-
Power at 60 mills/kWh	-	27,297,036	-	-	-
Labor (4%)	-	2,278,100	-	-	-
Maintenance (7%)	-	3,986,674	-	-	-
Subtotal	-	47,581,311	-	-	-
Cost (\$/t of CO ₂)	-	26	-	-	-
Cost (\$/1,000 scf of CO ₂)	-	1.45	-	-	-
Cost (\$/t of C)	-	96	-	-	-
Pipeline (500 km)					
Levelized fixed charge at 23%	-	33,424,663	-	-	-
Consumables	-	0	-	-	-
Power at 60 mills/kWh	-	1,976,729	-	-	-
Labor (4%)	-	5,812,985	-	-	-
Maintenance (7%)	-	10,172,724	-	-	-
Subtotal	-	51,387,100	-	-	-
Cost (\$/t of CO ₂)	-	28	-	-	-
Cost (\$/1,000 scf of CO ₂)	-	1.57	-	-	-
Cost (\$/t of C)	-	103	-	-	-

TABLE 13.9 (Cont.)

Factor	No. ^a	Cost (\$)	Change in Power (MW)	CO ₂ Emissions (kg/h)	CO ₂ Sequestering (kg/h)
Replacement power at 60 mills/kWh IGCC with shift/carbonate-net life cycle	-	-	420.3	59,259	322,341
Cost: base case - derated production	-	-	126.13	126,130	0
Sum	-	43,091,053	546.43	185,389	322,341
Cost (\$/t of CO ₂)	-	24	-	-	-
Cost (\$/1,000 scf of CO ₂)	-	1.31	-	-	-
Cost (\$/t of C)	-	87	-	-	-
Total cost	-	142,059,465	-	-	-
Total cost (\$/t of CO ₂)	-	77	-	-	-
Cost (\$/1,000 scf of CO ₂)	-	4.33	-	-	-
Total cost (\$/t of C)	-	284	-	-	-

^a Numbers refer to Table 8.5.

^b Not applicable.

^c Size of compressor, 4,986 brake hp.

^d Size of compressor, 924 brake hp.

CO₂ is clearly at a disadvantage in comparison with the other technologies. The low-pressure glycol system shows a distinct advantage, with a replacement power cost of \$14/t of CO₂.

13.2.4 Summation of Capital, Operating, and Replacement-Electricity Costs

The summation of the capital, operating, and replacement-electricity costs for the IGCC/shift/CO₂ recovery system and the pipeline costs appears in each of the tables. These costs range from \$61 to \$108/t of CO₂. The ranking of systems from most favorable total cost to highest cost is as follows: glycol > amine > hot K₂CO₃ > chilled methanol. Were it to be included, the membrane separation technology would have a cost of nearly twice that of the amine recovery, making membrane separation technology the highest cost option.

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