6 Integrated Gasification Combined Cycle with Glycol Solvent CO_2 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 CO_2 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 CO_2 recovery systems, some variation exists in the amount of heat available. For the glycol solvent process, the CO_2-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 CO_2 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 CO_2 removal. The high-pressure

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[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 CH_4, CO_2, and H_2S, for which solubilities are given in Figure 6.3, the solubility in a glycol solvent (Selexol) for other gases, particularly steam, CO, and N_2, is needed for the process stream calculations. Solubility values for these gases were based on the solubility of CO_2, as given in Figure 6.3, and the relative solubilities listed in Table 6.2.

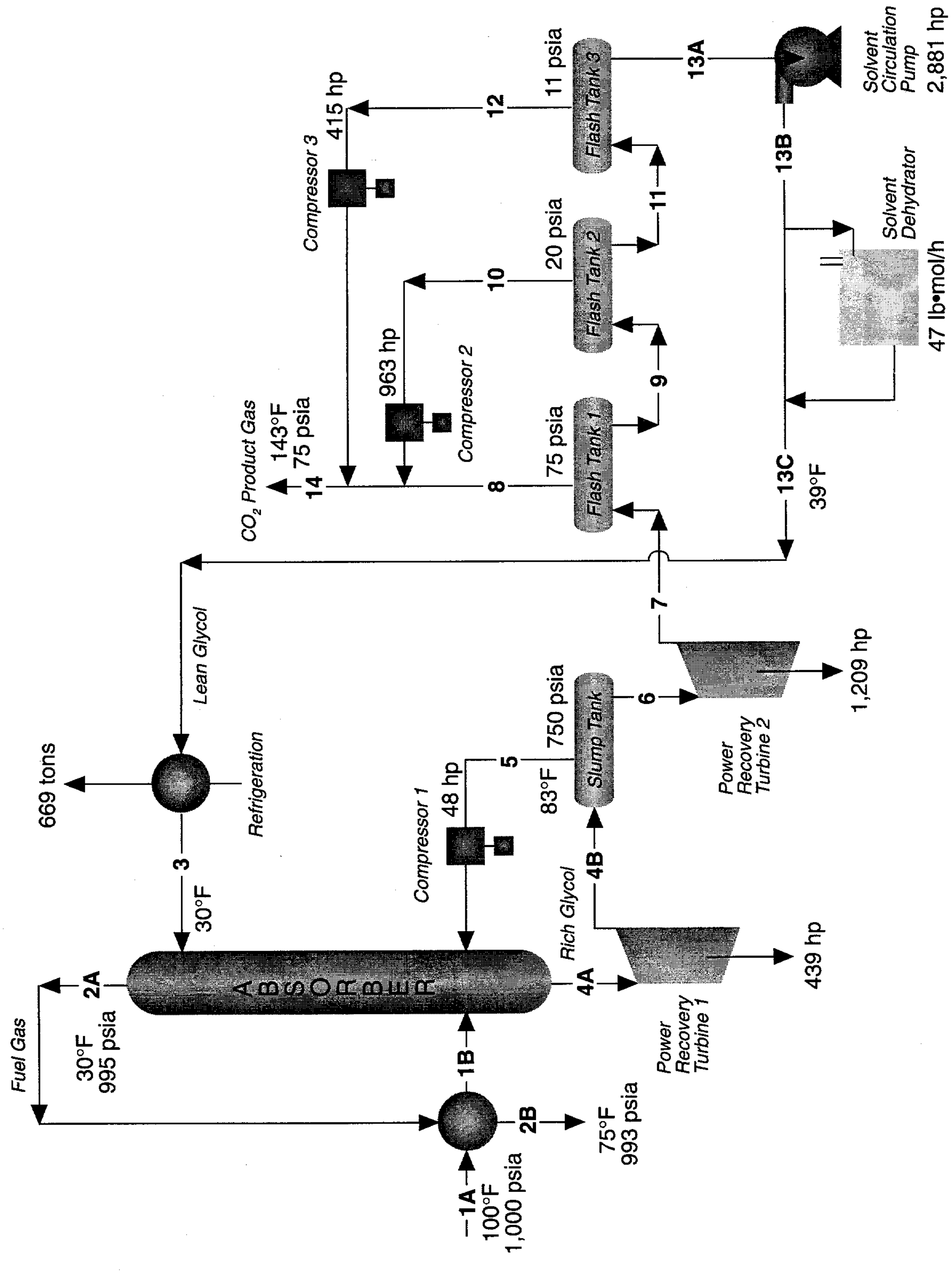

FIGURE 6.1 Summary of a Glycol Solvent Process: High-Pressure Case (numbers refer to streams)

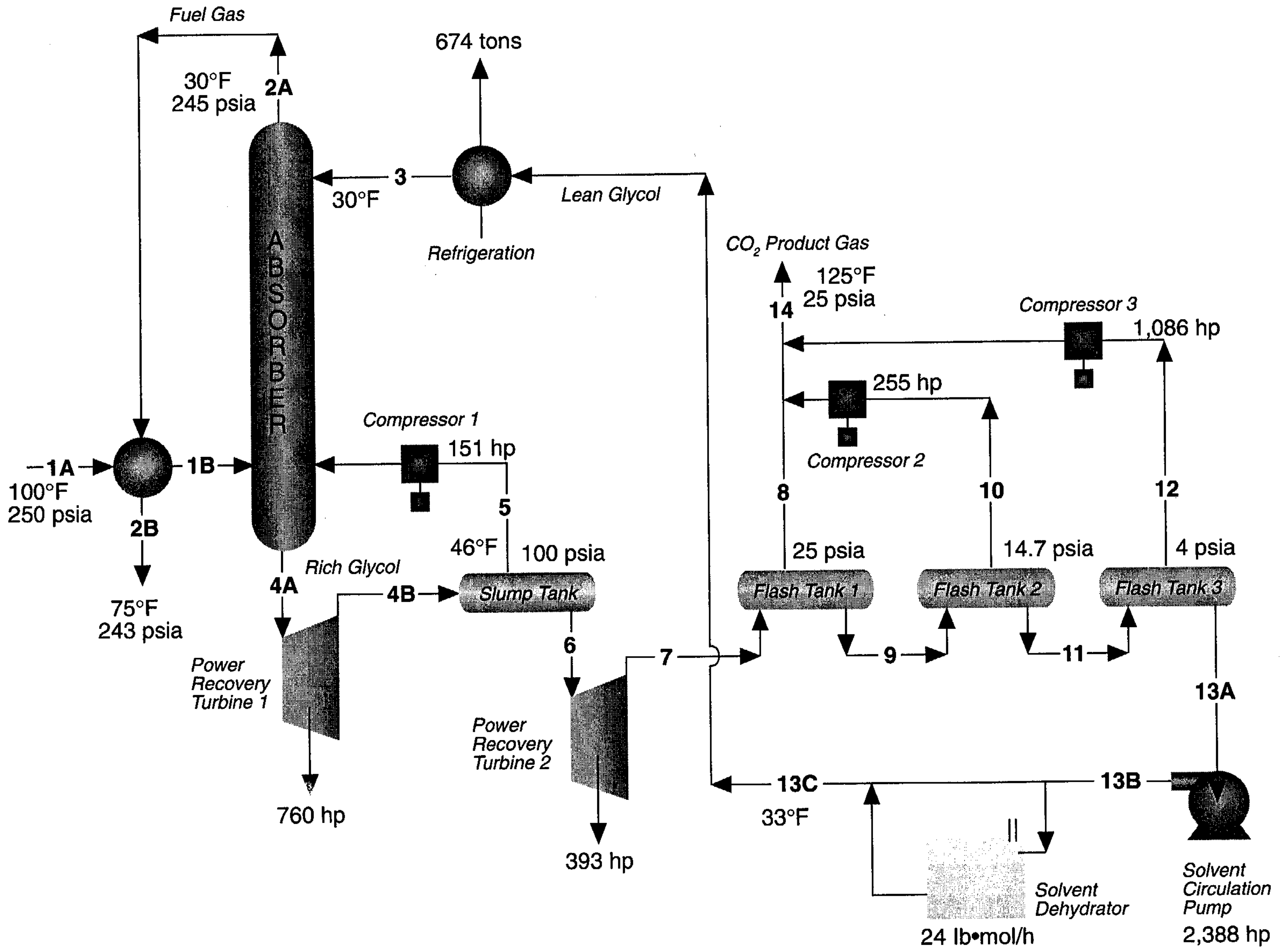

FIGURE 6.2 Summary of a Glycol Solvent Process: Low-Pressure Case (numbers refer to streams)

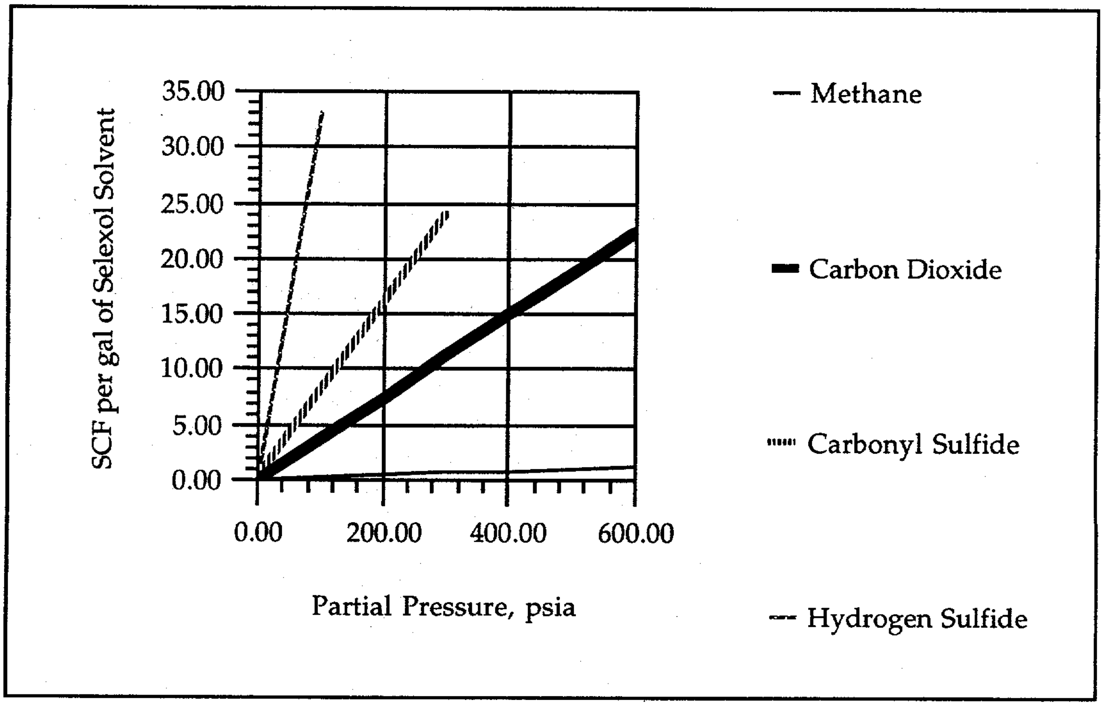

FIGURE 6.3 Solubility of Acid Gases in a Glycol Solvent (Selexol) at 70°F (SCF, standard cubic feet) (Source: Sweeny and Valentine 1970)

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, CO_2, and H_2 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_2O	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 CO_2 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 CO_2 recovery is less than with the other CO_2 recovery systems. The overall power balance for the plant is summarized in Table 6.6. The energy penalty associated with CO_2 recovery includes both reduced output from the turbogenerator units and electricity for pumping, compressing, and refrigeration in the CO_2 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 exhange	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_2O	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_2S	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_2O	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_2S	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_2O	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_2S	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 exhange	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_2O	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_2S	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_2O	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_2S	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_2O	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_2S	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_2O. 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_2O 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, $\int \upsilon dp$. 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_2O 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	

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[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)

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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 (includes all supporting equipment and connections and installation)	3.2	
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$23,000	
CE index for process equipment in 1987[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 (includes all supporting equipment and connections and installation)	3.2	
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) (hot fluid: gas, cold fluid: gas)[c]	40	
Heat transfer area (ft^2)	2,319	
Operating pressure (psia)	300	
Pressure factor	1.165	
Module factor (includes all supporting equipment and connections and installation)	3.2	
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$40,000	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		**$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 (includes all supporting equipment and connections and installation)	3.2	
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$161,742	
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 (includes all supporting equipment and connections and installation)	3.2	
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 (includes all supporting equipment and connections and installation)	3.2	
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) (hot fluid: gas, cold fluid: water)[c]	40	
Heat transfer area (ft^2)	11,056	
Operating pressure (psia)	1,000	
Pressure factor	1.23	
Module factor (includes all supporting equipment and connections and installation)	3.2	
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		**$135,000**
13. CO_2 Absorption Column		
Diameter of tower (ft)	11	
HETP (ft)	3	
No. of theoretical stages	10	
Absorber tower height (ft)	34	
(4 ft for inlet and outlet and for gas and liquid distributors)		
Volume of packing (ft^3)	2,852	
Pressure factor	4.2	
Cost per foot of column height ($/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) (2-in. pall rings-metal)	$63.5	
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. CO_2 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/ft^2/°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/ft^2/°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 ft^3 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 ft^3 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/ft^3 and that of the copper-based catalyst \$250/ft^3.

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 CO_2 is desorbed from the glycol solvent; and the other less soluble gases, such as H_2, CH_4, 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

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

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 CO_2 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 CO_2 in subsequent stages. A residence time of 3 minutes is selected on the basis of the literature (Woelfer et al. 1977):

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

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 CO_2 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 CO_2 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 CO_2 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 CO_2 released at 20 psia is compressed to a pressure of 75 psia. The compressed CO_2 is combined with the CO_2 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 CO_2 released at 11 psia is compressed to a pressure of 75 psia. The compressed CO_2 is combined with the CO_2 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 CO_2 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:

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

The energy efficiency ratio (EER) is defined as

$$EER = 9.0 + \frac{T_{evap}}{10} \quad . \tag{6.4}$$

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 (includes all supporting equipment and connections and installation)	3.2	
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$23,000	
CE index for process equipment in 1987[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 (includes all supporting equipment and connections and installation)	3.2	
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 6.8 (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) (hot fluid: gas, cold fluid: water)[d]	40	
Heat transfer area (ft^2)	2,319	
Operating pressure (psia)	285	
Pressure factor	1.165	
Module factor (includes all supporting equipment and connections and installation)	3.2	
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$40,000	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		**$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 (includes all supporting equipment and connections and installation)	3.2	
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$161,455	
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) (hot fluid: gas, cold fluid: water)[e]	5	
Heat transfer area (ft^2)	473,251	
Operating pressure (psia)	275	
Pressure factor	1.165	
Module factor (includes all supporting equipment and connections and installation)	3.2	
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$495,084	
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 (includes all supporting equipment and connections and installation)	3.2	
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$161,814	
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		**$135,000**
10. CO_2 Absorption Column		
Diameter of tower (ft)	17	
HETP (ft)	3	
No. of theoretical stages	12	
Absorber tower height (ft) (4 ft for inlet and outlet and for gas and liquid distributors)	40	
Volume of packing (ft^3)	8,175	
Pressure factor	2	
Cost per foot of column height ($/ft) (mild steel construction)	$1,800	
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) (2-in. pall rings-metal)	$63.5	
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[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		**$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[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		**$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. CO_2 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[g]	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/ft^2/°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 ft^3 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 ft^3 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 $50/$ft^3$ and that of the copper-based catalyst $250/$ft^3$.

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 CO_2 is desorbed from the glycol solvent; and the other less soluble gases, such as H_2, CH_4, 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 CO_2 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 CO_2 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 CO_2 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 CO_2 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 CO_2 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 CO_2 released at 14.7 psia is compressed to a pressure of 25 psia. The compressed CO_2 is combined with the CO_2 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 CO_2 released at 4 psia is compressed to a pressure of 25 psia. The compressed CO_2 is combined with the CO_2 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 CO_2 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 CO_2 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 CO_2 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 CO_2 recovery systems, some variation exists in the amount of heat available. For the chilled methanol process, the CO_2-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 (K_2CO_3).

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.

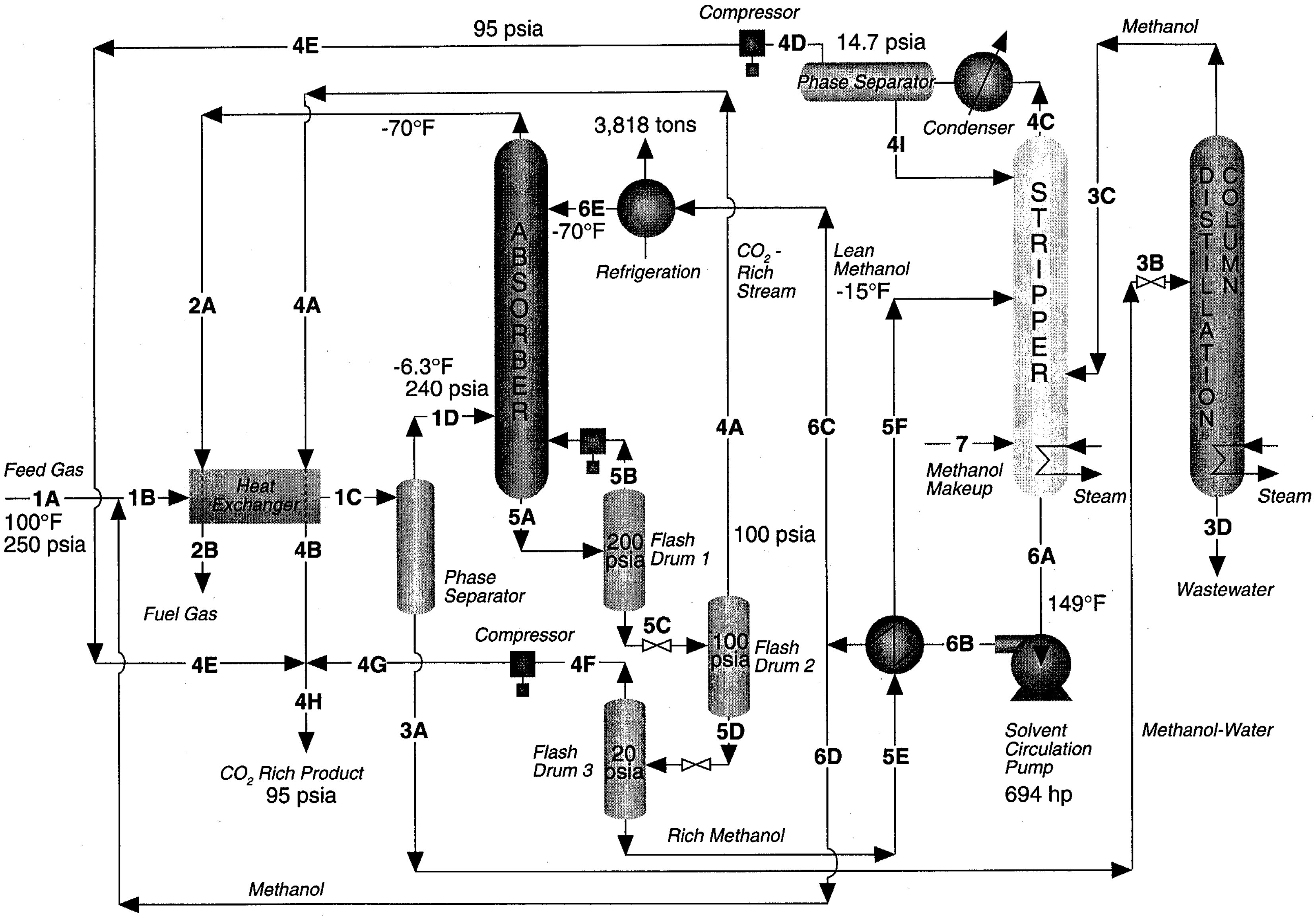

FIGURE 7.1 Summary of Chilled Methanol Process (numbers refer to streams)

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 CO_2 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 CO_2. The first flash drum in the series is used to recover combustible gases, which are recycled to the absorber. The remaining flash drums recover CO_2. Partial recovery of CO_2 in this manner reduces the compressor work required for the CO_2 product.

Final desorption of the CO_2 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 CO_2, which is cooled to condense accompanying solvent vapors and to prepare the CO_2 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_2O	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_2S	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_2O	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_2S	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_2O	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_2S	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_2O	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_2S	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_2O	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ar	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	0.40	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

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[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		
Temperature (°F)	-70	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.
Pressure (psia)	235	
Flow rate (lb·mol/h)	13,325	
CO_2 (mole fraction)	0.0120	
Stream 3A: Methanol-water mixture from phase separator		
Temperature (°F)	-6	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.
Pressure (psia)	240	
Flow rate (lb·mol/h)	110	
CO_2 (mole fraction)	0	
Stream 3B: Methanol-water mixture to distillation column		
Temperature (°F)	-6	Methanol is separated from methanol-water mixture in distillation column.
Pressure (psia)	50	
Flow rate (lb·mol/h)	110	
CO_2 (mole fraction)	0	
Stream 3C: Methanol from distillation column to stripper		
Temperature (°F)	150	Methanol from distillation column is sent to stripper.
Pressure (psia)	50	
Flow rate (lb·mol/h)	51	
CO_2 (mole fraction)	0	
Stream 3D: Wastewater from distillation column		
Temperature (°F)	280	Water from distillation column is removed from bottom of column for disposal.
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	

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[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 CO_2 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 CO_2 recovery system reflects the refrigeration demand. The energy penalty associated with CO_2 recovery includes both reduced output from the turbogenerator units and electricity consumption for pumping, compression, and refrigeration in the CO_2 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/ft^2/°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/ft^2/°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 CO_2-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 (includes all supporting equipment and connections and installation)	3.2	
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$23,000	
CE index for process equipment in 1987[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 (includes all supporting equipment and connections and installation)	3.2	
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$131,716	
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) (hot fluid: gas, cold fluid: water)[d]	40	
Heat transfer area (ft^2)		
Operating pressure (psia)	2,319	
Pressure factor	275	
Module factor (includes all supporting equipment and connections and installation)	3.2	
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$40,000	
CE index for process equipment in 1987	320	
CE index for process equipment in 1993	360.4	
Installed cost of heat exchanger in 1993		**$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) (hot fluid: gas, cold fluid: fuel gas)[c]	5	
Heat transfer area (ft^2)	37,361	
Operating pressure (psia)	275	
Pressure factor	1.165	
Module factor (includes all supporting equipment and connections and installation)	3.2	
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$183,923	
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) (hot fluid: gas, cold fluid: fuel gas)[c]	5	
Heat transfer area (ft^2)	305,556	
Operating pressure (psia)	275	
Pressure factor	1.165	
Module factor (includes all supporting equipment and connections and installation)	3.2	
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$417,425	
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 CO$_2$-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 (includes all supporting equipment and connections and installation)	3.2	
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$119,696	
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 CO_2-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) (hot fluid: gas, cold fluid: gas)[e]	5	
	Heat transfer area (ft^2)	79,086	
	Operating pressure (psia)	275	
	Pressure factor	1.165	
	Module factor (includes all supporting equipment and connections and installation)	3.2	
	Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$246,407	
	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		**$135,000**
10. CO_2 Absorption Column		
Diameter of tower (ft)	17	
HETP (ft)	3	
No. of theoretical stages	15	
Absorber tower height (ft) (4 ft for inlet and outlet and for gas and liquid distributors)	49	
Volume of packing (ft^3)	10,218	
Pressure factor	2	
Cost per foot of column height ($/ft) (mild steel construction)	$1,800	
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$) (2-in. pall rings-metal)	$63.5	
Total cost of packing		**$648,857**
11. CO_2 Stripping Column		
Diameter of tower (ft)	17	
HETP (ft)	3	
No. of theoretical stages	17	
Absorber tower height (ft) (4 ft for inlet and outlet and for gas and liquid distributors)	55	
Volume of packing (ft^3)	11,581	
Pressure factor	2	
Cost per foot of column height ($/ft) (mild steel construction)	$1,800	
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$) (2-in. pall rings-metal)	$63.5	
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 (includes all supporting equipment and connections and installation)	3.2	
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$209,837	
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. CO_2 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 (includes electric motor drive and gear reducer)	$450,000	
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 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**

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[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/ft^2/°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 CO_2-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 CO_2-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/ft^2/°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 ft^3 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 ft^3 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/ft^3 and that of the copper-based catalyst \$250/ft^3.

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 CO_2 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 CO_2 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 CO_2 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 CO_2 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 CO_2 released at 20 psia is compressed to a pressure of 95 psia. The compressed CO_2 is combined with the CO_2 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 CO_2 released from the stripper is at 14.7 psia and is compressed to a pressure of 95 psia. The compressed CO_2 is combined with the CO_2 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 CO_2 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 CO_2 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 CO_2 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 CO_2 recovery systems, some variation exists in the amount of heat available. For the hot K_2CO_3 process, the CO_2-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 K_2CO_3 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 K_2CO_3 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 CO_2-rich gas cooling after the second-stage shift reaction, heat of compression of the CO_2 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 K_2CO_3 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 K_2CO_3 process uses a 40% solution of aqueous K_2CO_3 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 CO_2 reacts with the solvent according to the following mildly exothermic reaction:

$$K_2CO_3 + CO_2 + H_2O \Leftrightarrow 2\ KHCO_3\ . \tag{8.1}$$

TABLE 8.2 Turbine Power Output, Plant Power Use, and Net Power Output for Hot K_2CO_3 Process

Power Variable	Power (MW)	
	Base Case	Hot K_2CO_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.

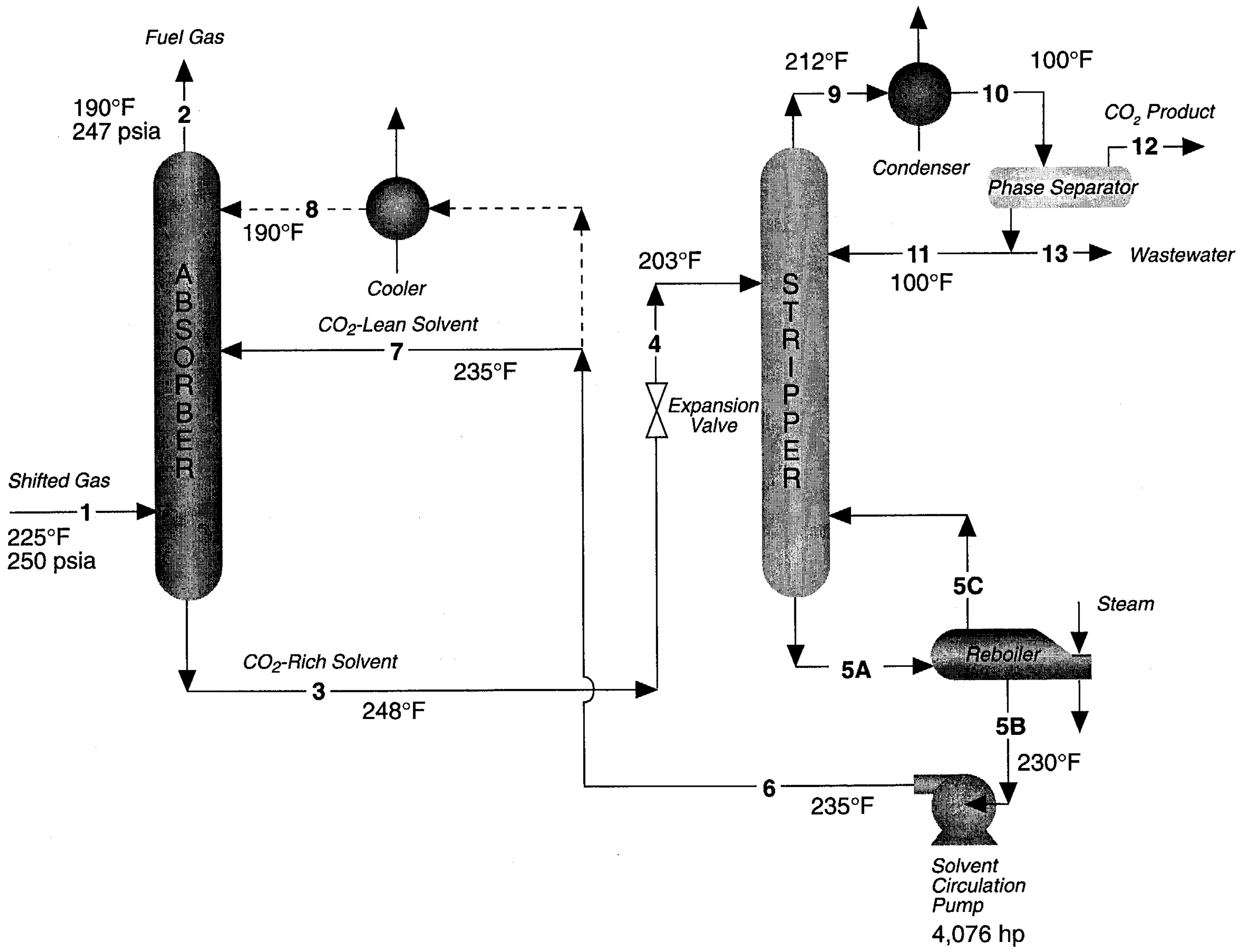

FIGURE 8.1 Summary of Hot K_2CO_3 Process (numbers refer to streams)

TABLE 8.3 Stream Flows of Hot K_2CO_3 Process[a]

Stream Data	Stream 1	Stream 2	Stream 3	Stream 4	Stream 5A
Description of stream	Absorber feed	Fuel gas	Rich K_2CO_3 from absorber	Rich K_2CO_3 to stripper	Lean K_2CO_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_2O	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_2S	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_2CO_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 K_2CO_3 Process[a] (Cont.)

Stream Data	Stream 5B	Stream 5C	Stream 6	Stream 7	Stream 8
Description of stream	Lean K_2CO_3 from reboiler	Recycle to stripper	Lean K_2CO_3 after circulation pump	Lean K_2CO_3 to absorber	Lean K_2CO_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_2O	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_2S	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_2CO_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 K_2CO_3 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_2O	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_2S	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_2CO_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_2CO_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_2S 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_2CO_3 from absorber		
Temperature (°F)	248	Solvent flow rate is designed for CO_2 concentration of 5.85 scf/gal of 40% K_2CO_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_2CO_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_2CO_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_2CO_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_2CO_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_2CO_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 CO_2, which is cooled to condense accompanying solvent vapors and to prepare CO_2 for compression to pipeline disposal conditions.

To attain a high CO_2 removal efficiency, the solvent stream has been split before it enters the absorber. A portion of the solvent stream is cooled to increase CO_2 solubility. This solvent is fed to the top of the absorber, where the CO_2 partial pressure is lower. The balance of the solvent is fed at a lower point in the absorber and is responsible for most CO_2 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 CO_2 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 K_2CO_3 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 CO_2 recovery includes both reduced output from the turbogenerator units and electricity consumption for pumping, compression, and refrigeration in the CO_2 recovery system. For the hot K_2CO_3 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 $Btu/h/ft^2/°F$ 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 K_2CO_3 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)	3.2	
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 (includes all supporting equipment and connections and installation)	3.2	
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$659,602	
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		**$135,000**
7.	**CO_2 Absorption Column**		
	Diameter of tower (ft)	8	
	HETP (ft)	3	
	No. of theoretical stages	15	
	Absorber tower height (ft) (4 ft for inlet and outlet and for gas and liquid distributors)	49	
	Volume of packing (ft^3)	2,263	
	Pressure factor	2	
	Cost per foot of column height ($/ft) (mild steel construction)	$1,000	
	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$) (2-in. pall rings-metal)	$63.5	
	Total cost of packing		**$143,691**
8.	**CO_2 Stripping Column**		
	Diameter of tower (ft)	12	
	HETP (ft)	3	
	No. of theoretical stages	17	
	Absorber tower height (ft) (4 ft for inlet and outlet and for gas and liquid distributors)	55	
	Volume of packing (ft^3)	5,770	
	Pressure factor	2	
	Cost per foot of column height ($/ft) (mild steel construction)	$1,400	
	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$) (2-in. pall rings-metal)	$63.5	
	Total cost of packing		**$366,413**
9.	**Heat Exchanger for Lean Solvent Cooling**		
	Q = Load (Btu/h)	33,383,789	
	Tha = Inlet temperature of hot fluid (°F)	235	
	Thb = Outlet temperature of hot fluid (°F)	190	
	Pressure of hot gases (psia)	250	
	Tca = Inlet water temperature (°F)	100	
	Tcb = Outlet temperature of saturated water (°F)	180	

TABLE 8.5 (Cont.)

9. (Cont.)		
ΔT_1	55	
ΔT_2	90	
Log mean temperature difference (°F)	70.84	
Overall heat transfer coefficient (Btu/h/ft^2/°F)[c]	150	
Heat transfer area (ft^2)	3,142	
Operating pressure (psia)	250	
Pressure factor	1.165	
Module factor (includes all supporting equipment and connections and installation)	3.2	
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$50,000	
CE index for process equipment in 1987	320	
CE index for process equipment in 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 (°F)	212	
Thb = Outlet temperature of hot fluid (°F)	100	
Tca = Inlet water temperature (°F)	70	
Tcb = Outlet water temperature (°F)	165	
ΔT_1	47	
ΔT_2	30	
Log mean temperature difference (°F)	37.87	
Overall heat transfer coefficient (Btu/h/ft^2/°F) (hot fluid: gas, cold fluid: water)[e]	40	
Heat transfer area (ft^2)	161,839	
Operating pressure (psia)	14.7	
Pressure factor	1	
Module factor (includes all supporting equipment and connections and installation)	3.2	
Purchased cost of heat exchanger in 1987 (mild steel construction; shell and tube floating head)	$730,415	
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 (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		**$398,077**

TABLE 8.5 (Cont.)

12. CO_2 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 (includes electric motor drive and gear reducer)	$750,000	
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 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 CO_2-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 ft^3 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 ft^3 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/ft^3 and that of the copper-based catalyst \$250/ft^3.

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 CO_2 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:

$$CO_2(g) + CaO(s) \Longrightarrow CaCO_3(s) \ . \tag{9.1}$$

Carbon dioxide, CO, H_2, 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 CO_2 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 [$CaMg(CO_3)_2$] 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 ($CaCO_3$) 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 CO_2 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 CO_2 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 CO_2 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, CO_2 removals greater than 95% are possible. At lower temperatures (550°C < T < 650°C), the partial pressure of the CO_2 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 CO_2 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 CO_2 absorbed and separated from the nitrogen-carrier gas in a second system that would look very similar to commercial CO_2 separation systems that have been investigated for the recovery of CO_2 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):

$$CO + H_2O <==> CO_2 + H_2 \ . \qquad (9.2)$$

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 (CO_2, CO, H_2, and H_2O) have already been alluded to in Section 9.7. The presence of trace amounts of CH_4 and heavier hydrocarbons and of the trace sulfur species, carbonyl sulfide (COS) and H_2S, needs to be understood in greater detail.

10 Integrated Gasification Combined Cycle with Low-Temperature Polymer Membrane CO_2 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 CO_2 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 CO_2 process. First, the nitrogen must be separated from the CO_2. Second, the CO_2 must be separated from H_2 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 CO_2-N_2 selectivity, cellulose acetate, with a CO_2-N_2 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: "CO_2-N_2-selectivities in excess of 200 will be competitive with conventional amine absorption technology" (p. 423).

The highest H_2-CO_2 selectivity commercially available is 10:1 by using polyimide. Hence, even though a high driving force is available, the production of a pipeline CO_2 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% H_2 in the permeate going to the CO_2 pipeline may represent an unacceptable loss of fuel, as well as a materials problem, because the pipeline would now be subject to H_2 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:

- CO_2:N_2 by a factor of 3, and
- CO_2: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 CO_2 by Pipeline

11.1 Commercial Considerations of CO_2 Separation Systems for Pipeline Gas

Before entering the pipeline, CO_2 and any contained gases must be purified, dried, and compressed. Purification is necessary because contaminants, such as H_2S and SO_2, 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 CO_2, SO_x, and NO_x to produce harmful acids. This gas preparation has taken place in the CO_2 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 CO_2 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 CO_2 from flue gases. An alternative is hot K_2CO_3 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 CO_2 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 CO_2 from CH_4 (also from higher hydrocarbons and H_2S). This process requires that the process stream be compressed and cooled (in stages) to condense the CO_2, 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 CO_2 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 CO_2 Pipeline

The CO_2 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 CO_2 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 CO_2 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 CO_2 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):

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

where

k = Cp/Cv = 7.08/5.09 = 1.39 (at 100°C),

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

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

T_1 = temperature at compressor inlet (122°F), and

T_2 = temperature at compressor outlet, calculated to be 254°F.

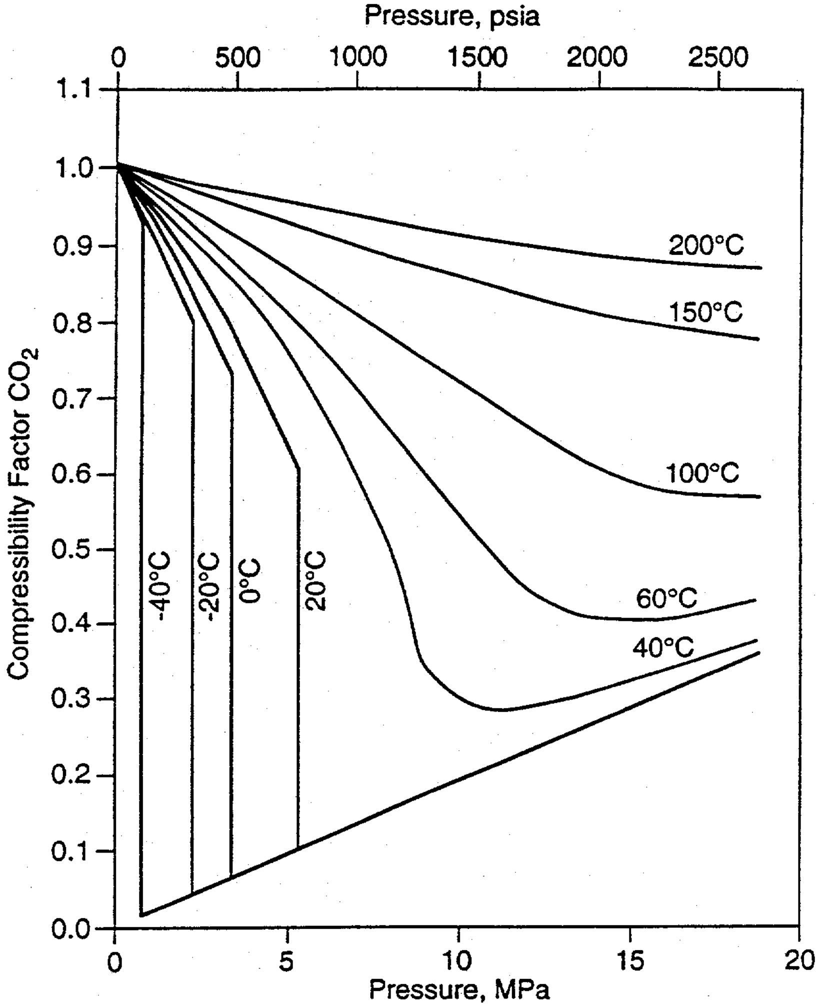

FIGURE 11.1 Compressibility Factor for CO_2 versus Pressure at Various Temperatures (adapted from Figure 1 of Farris [1983]) (Source: Lynch 1985).

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):

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

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 ft^3/d);

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

T_s = standard temperature (520°R);

k = Cp/Cv = 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 CO_2 emissions resulting from the use of this power are those coming from the IGCC system.

11.3 Substation Recompression

The length of the CO_2 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 CO_2, 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 CO_2 emission of 2.066 kg/h.

11.4 Total Power Consumption for CO_2 Pipeline

The compression costs from the CO_2 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 CO_2 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 CO_2 recovery system were normalized to a CO_2 value of 322,000 kg/h (the calculated CO_2 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 CO_2 Pipeline: Compression and Dehydration

The costs for pipelines that could carry CO_2 from the IGCC plant have been developed for previous DOE studies (Hare et al. 1978; Sparrow et al. 1988) concerned with supplying CO_2 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 CO_2 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 CO_2 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 CO_2 recovery from the IGCC system falls comfortably into this range at 142×10^6 scfd.

TABLE 11.1 Power Consumption for CO_2 Pipeline[a]

	Consumption of Electricity (MW)			
Source	Glycol (Low Pressure)	Amine	Chilled Methanol	K_2CO_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, CO_2 sequestered in reservoir at 322,000 kg/h.

TABLE 11.2 Equipment for CO_2 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 CO_2. The four types of reservoirs are as follows:

- T1 — 98% CO_2, 2% CH_4, and traces of H_2S;
- T2 — 96% CO_2, 25% CH_4, and 2% H_2S;
- T3 — 50% CO_2, 50% CH_4, and traces of H_2S; and
- T4 — 48% CO_2, 50% CH_4, and 25% H_2S.

The wellhead pressure is about 250 psig. The cost of recovering CO_2 is estimated at \$0.11/10^3 scf for T3 and \$0.47/10^3 scf for T4. The credit from the CH_4 by-product gas is \$1.75/10^3 scf. The cost to remove trace amounts of H_2S is estimated to be \$0.35/10^3 scf. The cost of compression and transmission (100 mi) at supercritical pressure of 2,200 psig ranges from a low of \$0.72/10^3 scf (T1) to a high of \$0.89/10^3 scf (T4).

TABLE 11.3 Pipeline Costs by Capacity and Duration

Flow Rate (10^6 scfd)	Cost ($\$/[10^3\ \text{scf} \cdot 100\ \text{mi}^{-1}]$) 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 = CO_2 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/$10^6$ 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/$10^6$ 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 CO_2, 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 CO_2 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 that 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 CO_2 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 CO_2

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

$$CO_2 + \text{Ca silicates (volcanic)} \Longrightarrow CaCO_3 + \text{sand} \ . \qquad (12.1)$$

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

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

Current anthropogenic activity releases 5.9 gigatonnes (Gt) of carbon per year as CO_2 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 CO_2.

The base-case IGCC system releases CO_2 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 CO_2 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 mi^2) (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 mi^2. Washington, D.C., encompasses an area of approximately 61 mi^2. 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.

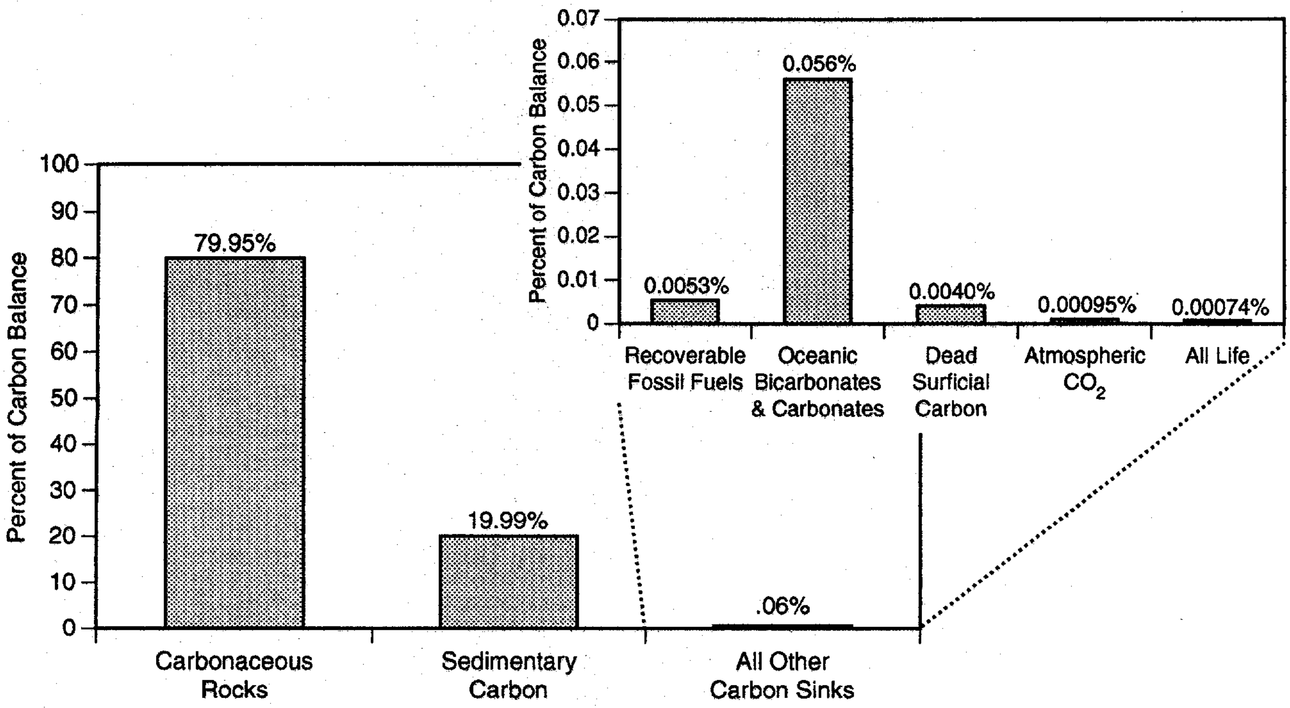

FIGURE 12.1 Distribution of Carbon on Earth

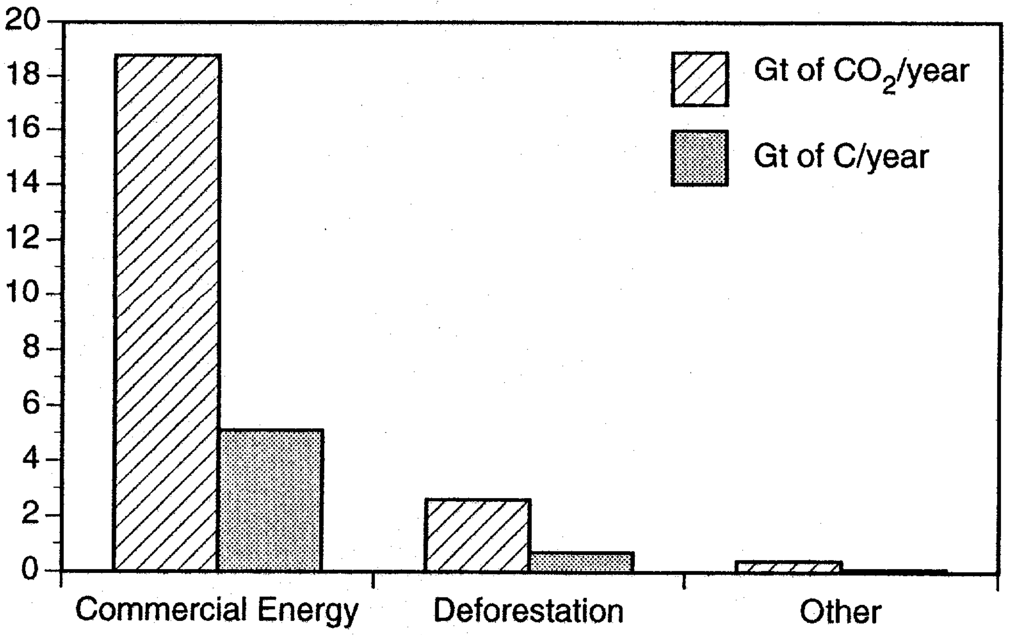

FIGURE 12.2 Current Release of Carbon and CO_2 Caused by Anthropogenic Activity

Microalgae have been proposed as a possible method for CO_2 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 = 0.77×10^9 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 mi^2 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 $\$376/10^3$ 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 CO_2 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 CO_2 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 (Condorelli 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

CO_2 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 CO_2 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 CO_2 will affect the bicarbonate ion formation occurring on the ocean floor, a location rich in $CaCO_3$. Because this reaction requires carbonates, CO_2, and water, whether the displacement of water by CO_2 on the ocean floor will promote or inhibit this reaction is not clear.

The injection of CO_2 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 CO_2 in the surface waters would retard the natural absorption of CO_2 from the atmosphere. Also, solubility is very low; therefore, large amounts of water would be needed.

Below 500 m, a propensity exists for CO_2 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 CO_2 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 CO_2 migrates toward the surface, and sink as heavier water (with dissolved CO_2). 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 CO_2 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 CO_2 can occupy (van der Meer 1992). Also, the mobility ratio (the ratio of the permeability to the viscosity of CO_2 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 CO_2 and freshwater is inevitable. The consequential damage as a result of this geochemical action is not well known.

The basic theory of storing CO_2 in underground repositories is based on fluid displacements in semiconfined regions. When large amounts of CO_2 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 CO_2 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 CO_2 into the atmosphere or groundwater.

Estimates of CO_2 terrestrial sedimentary basin storage capacity exceed 300×10^9 tons of CO_2 (Koide et al. 1992). Compared with the world total annual CO_2 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):

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

where

$$t_D = \frac{6.33kt}{\phi \mu c {r_w}^2} \tag{12.4}$$

and

$$r_{De} = \frac{r_e}{r_w} \quad . \tag{12.5}$$

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

T_f = 600°R (formation temperature);

k = 75 md (permeability);

h = 200 ft (cylindrical height);

ϕ = 15% (porosity);

r_w = 0.3 ft (well radius);

r_e = 6,000 ft (reservoir external radius); and

q_g = 5 MMscfd (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

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

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% CH_4, 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:

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

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

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

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

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

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

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

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

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

where μ is the viscosity.

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

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

From Equation 12.4,

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

Finally, from Equation 12.3,

$$\left(P^2\right) = \frac{1.424\overline{\mu z}\,(600)\left(5 \times 10^3\right)}{(200)\left(75 \times 10^{-3}\right)}\{0.024t + 9.15\}$$

$$= 2.848 \times 10^5\overline{\mu z}\,\{0.024t + 9.15\}\ .$$

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:

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

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

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

For this time, the total outflow is

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

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:

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

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

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

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

$T_c = 304.2$ K,

M_w = 44.01 g/g-mol,

V_c = 94.0 cm^3/g-mol,

z_c = 0.274,

ω = 0.420,

$P_c = 0.274\ (82.0567)\ (304.2)/94.0 = 72.76$ 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:

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

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

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

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

$\rho_r = 94.0/182 = 0.516;$

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

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

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

Compute the properties at $P = 500$:

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

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

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

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

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

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

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

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

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

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

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

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

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

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

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

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 CO_2 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 CO_2 Emissions

13.1.1 Base-Case IGCC

The IGCC base case without CO_2 controls yields the energy consumption and CO_2 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 CO_2 emission rate of 0.835 kg/kWh. This rate is significantly below the typical CO_2 emission rate of 1 kg/kWh for the installed electric power generating capacity used throughout the study. To summarize the ranking of the CO_2 recovery systems with respect to favorable fuel-cycle CO_2 sequestering costs: glycol > K_2CO_3 > amine > chilled methanol.

13.1.2 IGCC/Shift/Amine CO_2 Recovery Case

Table 13.2 surveys the base-case IGCC system with shift and conventional amine-based CO_2 recovery. The IGCC/CO_2 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 CO_2 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 CO_2 sequestering come to 0.449 kWh/kg of CO_2, a value derived by dividing this net derating of 144 MW by the net decrease in CO_2 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 CO_2 Recovery Case

Table 13.3 surveys the base-case IGCC system with shift and low-pressure glycol CO_2 recovery. This case represents the best CO_2 recovery performance observed in this survey. The IGCC/CO_2 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 CO_2 emission rate of 0.156 kg/kWh. This yield is accomplished with a net derating of 73.6 MW. The energy costs for CO_2 sequestering come to 0.23 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.

TABLE 13.1 Energy Consumption and CO_2 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 CO_2 Recovery Case

Table 13.4 surveys the base-case IGCC system with shift and chilled methanol CO_2 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/CO_2 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 CO_2 Emissions for KRW IGCC with Amine CO_2 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 CO_2.

TABLE 13.3 Energy Consumption and CO_2 Emissions for KRW IGCC with Low-Pressure Glycol CO_2 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 CO_2.

TABLE 13.4 Energy Consumption and CO_2 Emissions for KRW IGCC with Chilled Methanol CO_2 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 CO_2.

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 CO_2 Emissions for KRW IGCC with Hot Potassium Carbonate CO_2 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 CO_2.

TABLE 13.6 Economics of Shift/MEA CO_2 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 CO_2 recovery systems. While these costs of $28/t of CO_2 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 CO_2 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/CO_2 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 CO_2 that operate in the natural gas distribution network, the pipeline cost of $1.57/scf of CO_2 could come into line with the distribution costs for a natural gas pipeline of approximately $0.50/scf of CO_2.

TABLE 13.7 Economics of Shift/Low-Pressure Glycol CO_2 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 CO_2 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/CO_2 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 CO_2 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 CO_2 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.

CO_2, 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 CO_2.

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/CO_2 recovery system and the pipeline costs appears in each of the tables. These costs range from \$61 to \$108/t of CO_2. The ranking of systems from most favorable total cost to highest cost is as follows: glycol > amine > hot K_2CO_3 > 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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