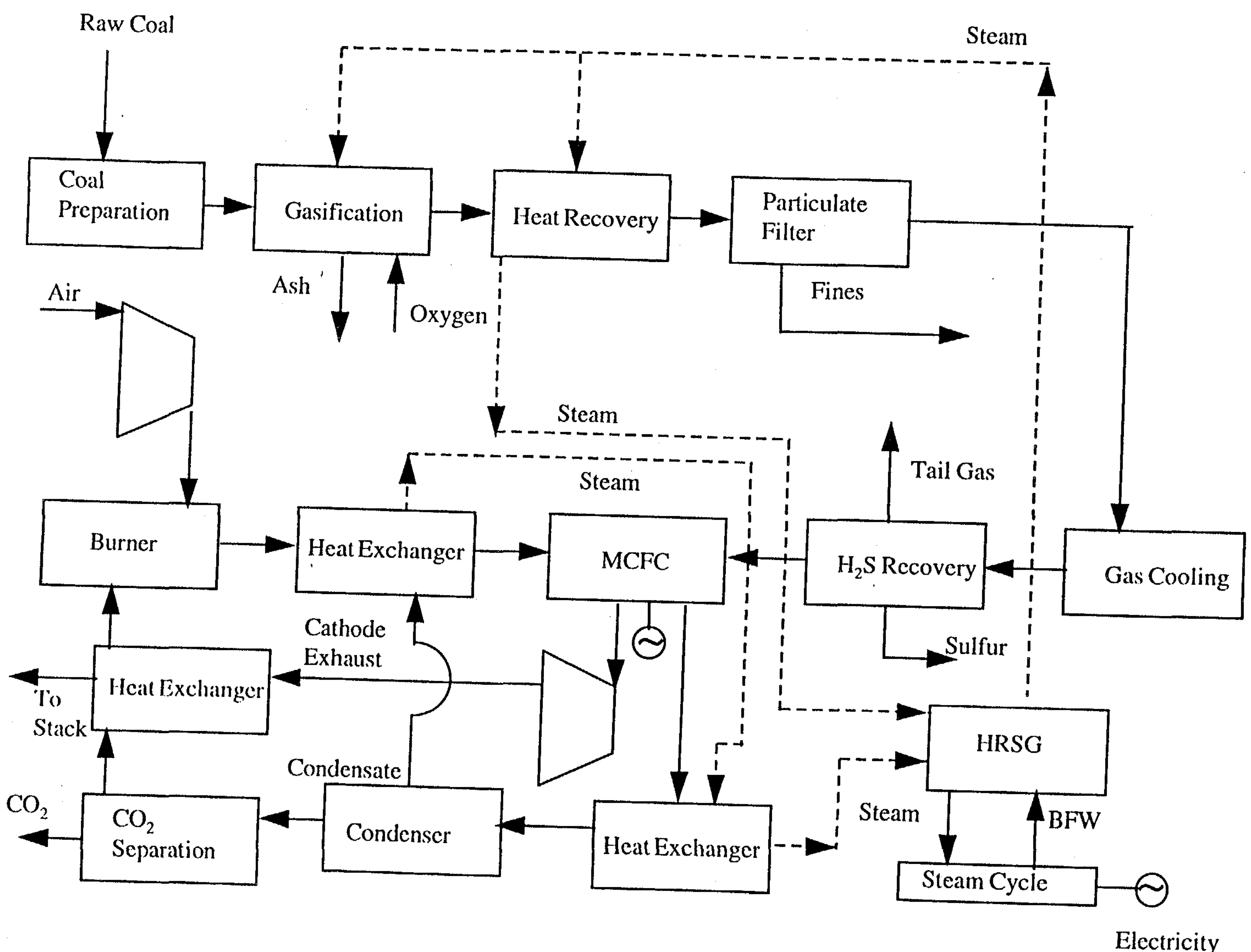

FIGURE 6.1 Block Diagram of the IGCC System with CO_2 Recovery Used in Cases 3 and 4

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

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

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

6.3 Molten Carbonate Fuel Cell System

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

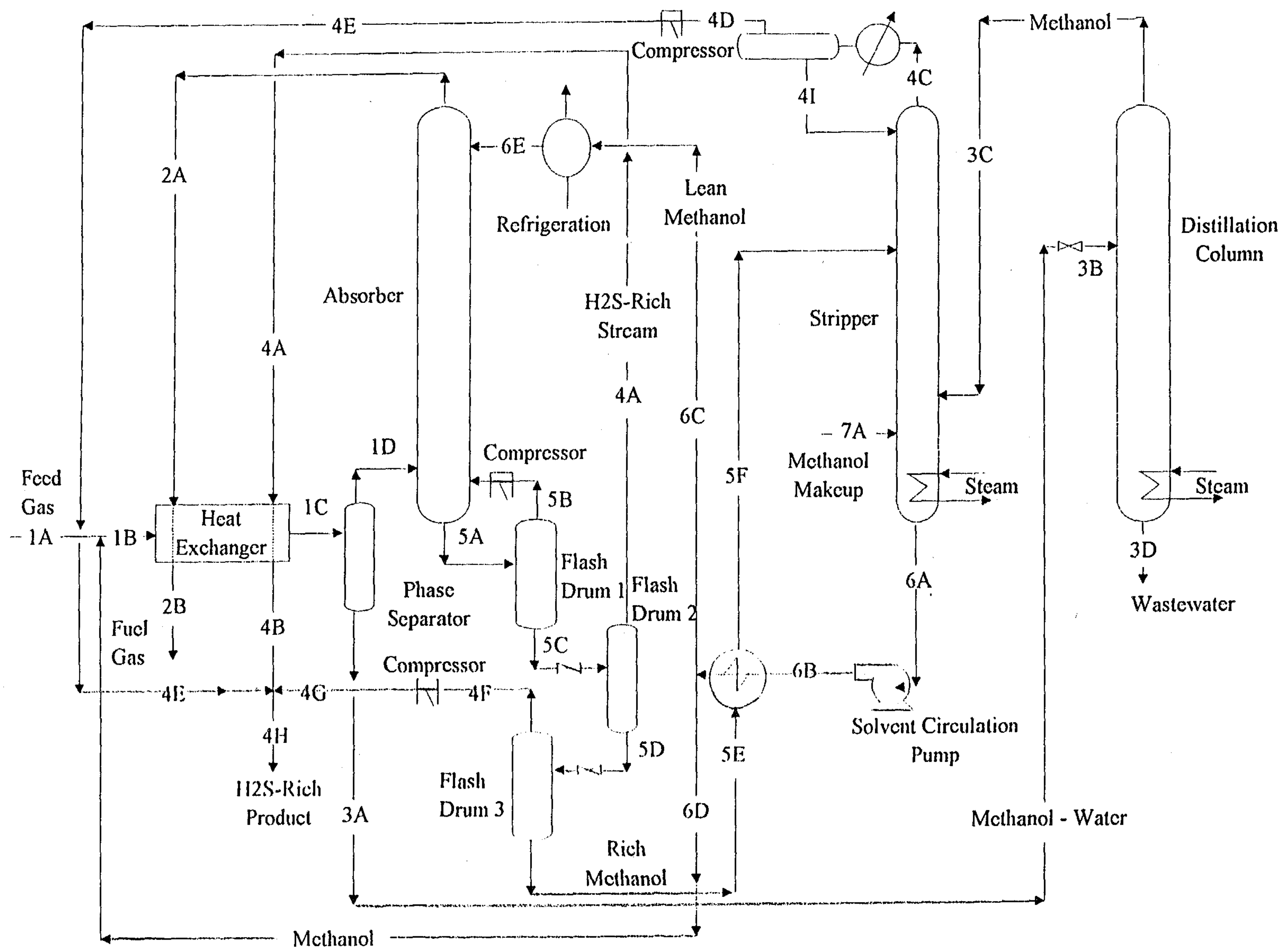

FIGURE 6.2 Flow Diagram of Chilled Methanol Process for H_2S Recovery in Case 3

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

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

TABLE 6.2 (Cont.)

Stream Data	Stream 3A	Stream 3B	Stream 3C	Stream 3D	Stream 4A	Stream 4B
Description of stream	Bottoms from phase separator	Feed to distillation column	Overhead from distillation column	Wastewater from distillation column	H_2S-rich gas from flash drum 2	H_2S-rich gas from heat exchanger
Gases (lb·mol/h)						
CO	0.00	0.00	0.00	0.00	21.48	21.48
CO_2	0.00	0.00	0.00	0.00	61.14	61.14
H_2	0.00	0.00	0.00	0.00	3.14	3.14
H_2O	19.41	19.41	0.79	18.62	0.00	0.00
N_2	0.00	0.00	0.00	0.00	0.00	0.00
Ar	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	0.00	0.00	0.00	0.00	4.05	4.05
NH_3	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	0.00	0.00	0.00	0.00	17.83	17.83
HCN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	2.35	2.35
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	19.41	19.41	0.79	18.62	109.99	109.99
Liquids (lb·mol/h) Methanol	15.07	15.07	15.07	0.00	0.00	0.00
Temperature (°F)	-34.18	-34.18	150.00	280.00	-29.88	84.50
Pressure (psia)	456.00	50.00	50.00	50.00	150.00	150.00
Enthalpy of stream (Btu/h) (reference, 32°F)	-41,190	-41,190	270,074	83,134	-55,403	48,168

TABLE 6.2 (Cont.)

Stream Data	Stream 4C	Stream 4D	Stream 4E	Stream 4F	Stream 4G	Stream 4H
Description of stream	Overhead from stripper	H_2S-rich gas from phase separator	H_2S-rich gas after compressor	H_2S-rich gas from flash drum 3	H_2S-rich gas after compressor	H_2S-rich product
Gases (lb·mol/h)						
CO	1.79	1.79	1.79	5.37	5.37	28.64
CO_2	30.57	30.57	30.57	30.57	30.57	122.29
H_2	0.16	0.16	0.16	0.63	0.63	3.93
H_2O	0.79	0.79	0.79	0.00	0.00	0.79
N_2	0.00	0.00	0.00	0.00	0.00	0.00
Ar	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	1.35	1.35	1.35	1.35	1.35	6.75
NH_3	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	28.70	28.70	28.70	12.48	12.48	59.01
HCN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00
COS	3.42	3.42	3.42	1.64	1.64	7.42
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	66.78	66.78	66.78	52.04	52.04	228.83
Liquids (lb·mol/h) Methanol	224.97	21.45	21.45	0.00	0.00	21.45
Temperature (°F)	135.00	100.00	619.94	-33.64	286.22	318.59
Pressure (psia)	14.70	14.70	150.00	20.00	150.00	95.00
Enthalpy of stream (Btu/h) (reference, 32°F)	3,874,837	410,887	803,850	-28,354	117,525	969,544

TABLE 6.2 (Cont.)

Stream Data	Stream 4I	Stream 5A	Stream 5B	Stream 5C	Stream 5D	Stream 5E
Description of stream	Methanol reflux to stripper	Rich methanol from absorber	Gases from flash drum 1	Feed to flash drum 2	Feed to flash drum 3	Rich methanol from flash drum 3
Gases (lb·mol/h)						
CO	0.00	286.38	257.74	28.64	7.16	1.79
CO_2	0.00	152.86	30.57	122.29	61.14	30.57
H_2	0.00	39.27	35.34	3.93	0.79	0.16
H_2O	0.00	0.00	0.00	0.00	0.00	0.00
N_2	0.00	0.00	0.00	0.00	0.00	0.00
Ar	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	0.00	67.52	60.77	6.75	2.70	1.35
NH_3	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	0.00	66.02	6.60	59.42	41.60	29.12
HCN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	8.70	0.87	7.83	5.48	3.84
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	0.00	620.75	391.89	228.86	118.87	66.83
Liquids (lb·mol/h) Methanol	203.53	4,114.03	0.00	4,114.03	4,114.03	4,114.03
Temperature (°F)	100.00	-23.04	-23.04	-23.04	-29.88	-33.64
Pressure (psia)	95.00	450.00	300.00	300.00	150.00	20.00
Enthalpy of stream (Btu/h) (reference, 32°F)	250,640	-4,358,335	-153,557	-4,204,777	-4,671,853	-4,927,451

TABLE 6.2 (Cont.)

Stream Data	Stream 5F	Stream 6A	Stream 6B	Stream 6C	Stream 6D	Stream 6E	Stream 7A
Description of stream	Rich methanol to stripper	Lean methanol to circulation pump	Lean methanol from solvent circulation pump	Lean methanol to refrigeration	Methanol for feed gas injection	Lean methanol to absorber	Methanol makeup to stripper
Gases (lb·mol/h)							
CO	1.79	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	30.57	0.00	0.00	0.00	0.00	0.00	0.00
H_2	0.16	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ar	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	1.35	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	29.12	0.41	0.41	0.41	0.00	0.41	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
COS	3.84	0.41	0.41	0.41	0.00	0.41	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	66.83	0.82	0.82	0.82	0.00	0.82	0.00
Liquids (lb·mol/h) Methanol	4,114.03	4,129.14	4,129.14	4,114.06	15.07	4,114.06	21.48
Temperature (°F)	128.66	149.00	152.90	-10.00	-10.00	-70.00	70.00
Pressure (psia)	20.00	14.70	456.00	456.00	456.00	456.00	14.70
Enthalpy of stream (Btu/h) (reference, 32°F)	7,257,493	8,749,976	9,043,938	-3,129,543	-11,463	-7,600,310	14,782

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

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

TABLE 6.3 (Cont.)

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

TABLE 6.3 (Cont.)

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

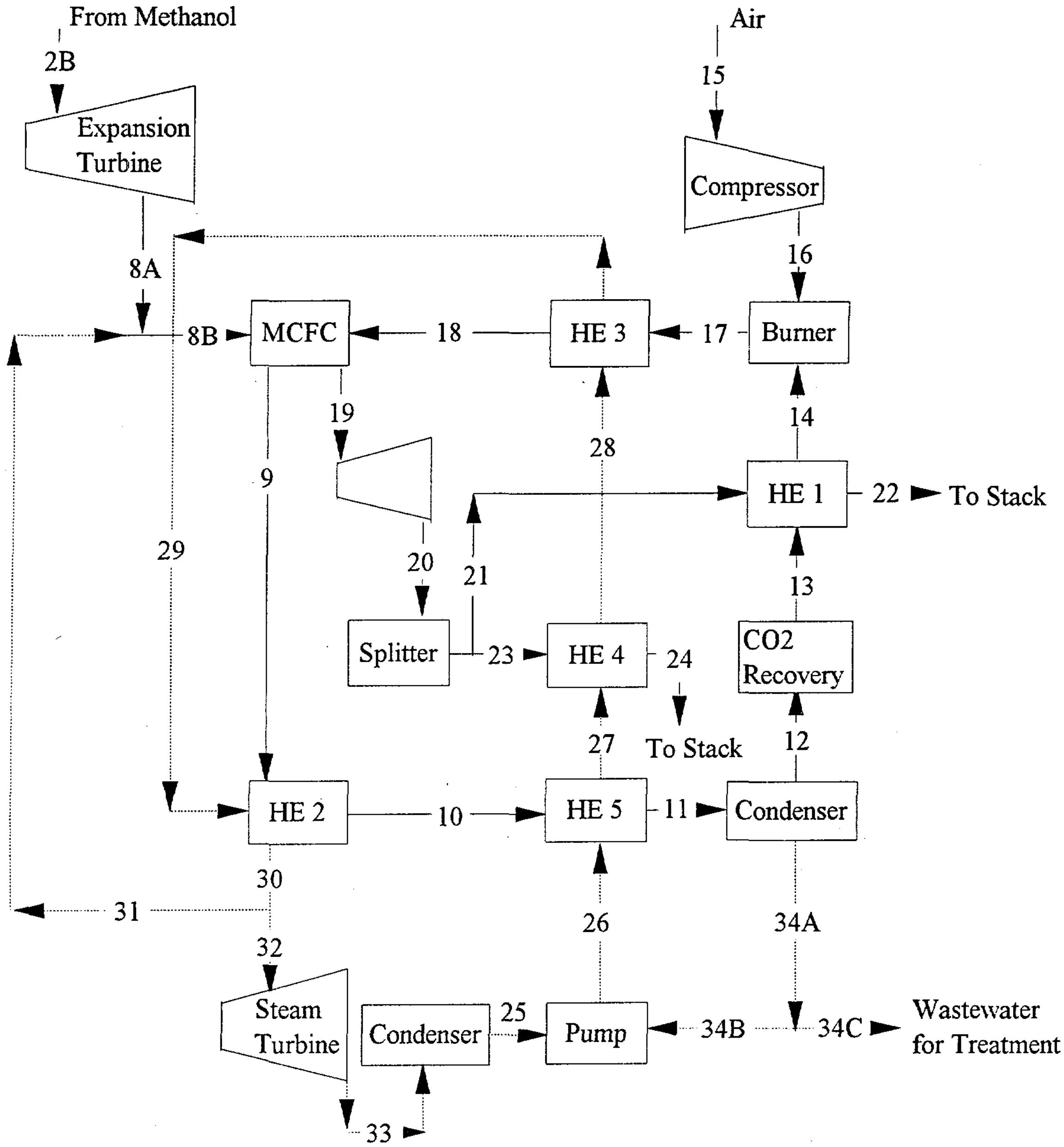

FIGURE 6.3 Flow Diagram of Fuel Cell System and Associated Heat Recovery in Case 3

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

Stream Data	Stream 2B	Stream 8A	Stream 8B	Stream 9	Stream 10	Stream 11
Description of stream	Feed gas from methanol	Fuel gas from expansion turbine	Fuel gas to fuel cell	Fuel cell anode exhaust	Gases from heat exchanger 2	Gases from heat exchanger 5
Gases (lb·mol/h)						
CO	4,530.65	4,530.65	4,530.65	1,812.26	1,812.26	1,812.26
CO_2	267.08	267.08	267.08	8,724.66	8,724.66	8,724.66
H_2	2,311.51	2,311.51	2,311.51	1,693.50	1,693.50	1,693.50
H_2O	0.00	0.00	12,000.00	13,579.13	13,579.13	13,579.13
N_2	36.44	36.44	36.44	36.44	36.44	36.44
Ar	72.73	72.73	72.73	72.73	72.73	72.73
CH_4	480.56	480.56	480.56	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	0.8 ppm	0.8 ppm	0.8 ppm	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.1 ppm	0.1 ppm	0.1 ppm	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	7,698.98	7,698.98	19,698.97	25,918.72	25,918.72	25,918.72
Liquids (lb·mol/h) H_2O	0.00	0.00	0.00	0.00	0.00	0.00
Temperature (°F)	80.00	-28.67	502.23	1,300.00	450.00	150.00
Pressure (psia)	450.00	150.00	150.00	150.00	150.00	150.00
Enthalpy of stream (Btu/h) (reference, 32°F)	2,596,057	-3,248,936	302,507,570	591,791,666	351,871,758	41,118,662

TABLE 6.4 (Cont.)

Stream Data	Stream 12	Stream 13	Stream 14	Stream 15	Stream 16	Stream 17
Description of stream	Gases to glycol process	Gases from glycol process	Gases from heat exchanger 1	Air to compressor	Air from compressor	Gases from burner
Gases (lb·mol/h)						
CO	1,812.26	1,794.14	1,794.14	0.00	0.00	0.00
CO_2	8,724.66	3,926.10	3,926.10	0.00	0.00	5,720.23
H_2	1,693.50	1,688.08	1,688.08	0.00	0.00	0.00
H_2O	78.96	0.00	0.00	0.00	0.00	1,688.08
N_2	36.44	35.71	35.71	44,202.24	44,202.24	44,237.95
Ar	72.73	72.73	72.73	543.05	543.05	615.78
CH_4	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	11,822.71	11,822.71	10,081.60
COS	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	12,418.55	7,516.76	7,516.76	56,568.00	56,568.00	62,343.64
Liquids (lb·mol/h) H_2O	0.00	0.00	0.00	0.00	0.00	0.00
Temperature (°F)	70.00	56.14	600.00	81.00	713.05	1,411.87
Pressure (psia)	150.00	145.00	145.00	14.70	150.00	150.00
Enthalpy of stream (Btu/h) (reference, 32°F)	3,933,703	1,441,106	36,784,141	19,042,871	275,242,646	706,519,291

TABLE 6.4 (Cont.)

Stream Data	Stream 18	Stream 19	Stream 20	Stream 21	Stream 22	Stream 23
Description of stream	Gases from heat exchanger 3	Fuel cell cathode exhaust	Gases from expansion turbine	Gases from splitter to heat exchanger 1	Gases from heat exchanger 1	Gases from splitter to heat exchanger 4
Gases (lb·mol/h)						
CO	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	5,720.23	461.60	461.60	73.82	73.82	387.78
H_2	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	1,688.08	1,688.08	1,688.08	269.95	269.95	1,418.13
N_2	44,237.95	44,237.95	44,237.95	7,074.39	7,074.39	37,163.55
Ar	615.78	615.78	615.78	98.47	98.47	517.31
CH_4	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	10,081.60	7,452.29	7,452.29	1,191.75	1,191.75	6,260.54
COS	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	62,343.64	54,455.70	54,455.70	8,708.38	8,708.38	45,747.31
Liquids (lb·mol/h) H_2O	0.00	0.00	0.00	0.00	0.00	0.00
Temperature (°F)	980.33	1,300.00	667.23	667.23	100.00	667.23
Pressure (psia)	150.00	150.00	14.70	14.70	14.70	14.70
Enthalpy of stream (Btu/h) (reference, 32°F)	482,306,421	546,008,492	280,696,393	44,888,086	9,545,050	235,808,307

TABLE 6.4 (Cont.)

Stream Data	Stream 24	Stream 25	Stream 26	Stream 27	Stream 28	Stream 29
Description of stream	Gases from heat exchanger 4	Water from condenser	Water from pump	Steam from heat exchanger 5	Steam from heat exchanger 4	Steam from heat exchanger 3
Gases (lb·mol/h)						
CO	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	387.78	0.00	0.00	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	1,418.13	0.00	0.00	9,917.77	15,668.16	30,055.55
N_2	37,163.55	0.00	0.00	0.00	0.00	0.00
Ar	517.31	0.00	0.00	0.00	0.00	0.00
CH_4	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	6,260.54	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	45,747.31	0.00	0.00	9,917.77	15,668.16	30,055.55
Liquids (lb·mol/h) H_2O	0.00	24,215.42	36,215.42	26,297.65	20,547.25	6,159.87
Temperature (°F)	400.00	121.36	121.36	356.77	356.77	356.77
Pressure (psia)	14.70	1.76	146.96	146.96	146.96	146.96
Enthalpy of stream (Btu/h) (reference, 32°F)	146,194,205	38,996,150	58,258,290	368,957,384	458,571,485	682,784,353

TABLE 6.4 (Cont.)

Stream Data	Stream 30	Stream 31	Stream 32	Stream 33	Stream 34A	Stream 34B	Stream 34C
Description of stream	Steam from heat exchanger 2	Steam for heating feed to fuel cell	Steam to steam turbine	Steam turbine exhaust	Water from condenser	Makeup water to pump	Wastewater for treatment
Gases (lb·mol/h)							
CO	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	36,215.42	12,000.00	24,215.42	22,813.57	0.00	0.00	0.00
N_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ar	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	36,215.42	12,000.00	24,215.42	22,813.57	0.00	0.00	0.00
Liquids (lb·mol/h) H_2O	0.00	0.00	0.00	1,401.85	13,500.17	12,000.00	1,500.17
Temperature (°F)	775.00	775.00	775.00	121.36	70.00	70.00	70.00
Pressure (psia)	146.96	146.96	146.96	1.76	150.00	150.00	150.00
Enthalpy of stream (Btu/h) (reference, 32°F)	922,758,266	305,756,514	617,001,752	459,798,748	9,234,116	8,208,000	1,026,116

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

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

TABLE 6.5 (Cont.)

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

TABLE 6.5 (Cont.)

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

TABLE 6.5 (Cont.)

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

TABLE 6.5 (Cont.)

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

TABLE 6.5 (Cont.)

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

6.4 Glycol Process for CO_2 Recovery

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

6.5 Fuel Cell, Steam Cycle, and Plant Performance

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

6.6 Economics

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

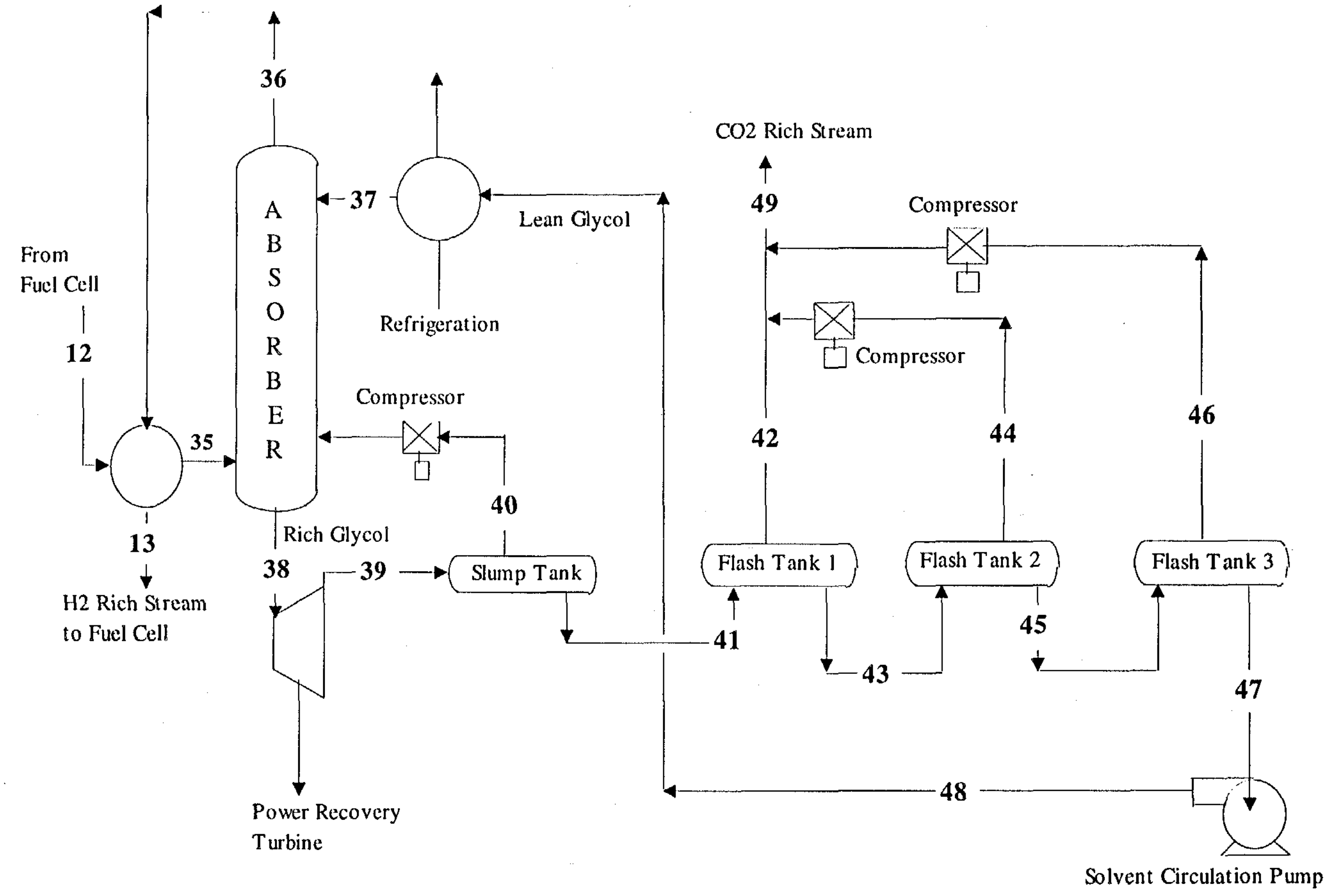

FIGURE 6.4 Flow Diagram of Glycol Process for CO_2 Recovery and Chilled Methanol Process for H_2S Recovery with Fuel Cell Topping Cycle in Case 3

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

Stream Data	Stream 12	Stream 13	Stream 35	Stream 36	Stream 37	Stream 38
Description of stream	Feed gas from fuel cell system	H_2-rich gas after heat exchanger	Absorber feed	Clean fuel gas from absorber	Lean glycol solvent to absorber	Rich glycol from absorber
Gases (lb·mol/h)						
CO	1,812.16	1,794.14	1,812.16	1,794.14	0.00	181.23
CO_2	8,724.66	3,926.10	8,724.66	3,926.10	210.13	5,110.91
	1,693.50	1,688.08	1,693.50	1,688.08	10.41	158.18
H_2O	78.96	0.00	78.96	0.00	0.00	79.76
N_2	36.44	35.71	36.44	35.71	0.00	7.29
Ar	72.73	72.73	72.73	72.73	0.00	0.00
CH_4	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	0.01	0.00	0.01	0.00	0.00	0.01
HCN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	12,418.46	7,516.76	12,418.46	7,516.76	220.54	5,537.38
Liquids (lb·mol/h) Glycol solvent	0.00	0.00	0.00	0.00	20,802.84	20,802.84
Temperature (°F)	70.00	56.14	55.00	30.00	30.00	50.36
Pressure (psia)	150.00	145.00	150.00	145.00	150.00	145.00
Enthalpy of stream (Btu/h) (reference, 32°F)	3,933,489	1,441,041	2,373,591	-119,000	-5,712,163	53,306,378

TABLE 6.6 (Cont.)

Stream Data	Stream 39	Stream 40	Stream 41	Stream 42	Stream 43	Stream 44
Description of stream	Rich glycol solvent after turbine 1	Gases from slump tank	Rich glycol to flash tank 1	CO_2-rich gas from flash tank 1	Rich glycol to flash tank 2	CO_2-rich gas from flash tank 2
Gases (lb·mol/h)						
CO	181.23	163.10	18.12	18.12	0.00	0.00
CO_2	5,110.91	102.22	5,008.69	3,756.52	1,252.17	876.52
H_2	158.18	142.36	15.82	1.58	14.24	2.14
H_2O	79.76	0.80	78.96	78.96	0.00	0.00
N_2	7.29	6.56	0.73	0.73	0.00	0.00
Ar	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	0.01	0.00	0.01	0.00	0.01	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	5,537.38	415.04	5,122.33	3,855.91	1,266.42	878.66
Liquids (lb·mol/h) Glycol solvent	20,802.84	0.00	20,802.84	0.00	20,802.84	0.00
Temperature (°F)	49.97	49.57	49.57	34.81	34.81	31.33
Pressure (psia)	50.00	50.00	50.00	25.00	25.00	14.70
Enthalpy of stream (Btu/h) (reference, 32°F)	52,160,172	53,779	50,942,277	95,822	8,065,266	-5,201

TABLE 6.6 (Cont.)

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

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

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

TABLE 6.7 (Cont.)

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

TABLE 6.7 (Cont.)

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

TABLE 6.7 (Cont.)

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

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

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

[a] Includes H_2S recovery system energy use.

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

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

TABLE 6.9 (Cont.)

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

TABLE 6.9 (Cont.)

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

TABLE 6.9 (Cont.)

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

TABLE 6.9 (Cont.)

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

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

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

TABLE 6.10 (Cont.)

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

TABLE 6.10 (Cont.)

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

TABLE 6.10 (Cont.)

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

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

TABLE 6.10 (Cont.)

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

TABLE 6.10 (Cont.)

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

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

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

TABLE 6.11 (Cont.)

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

TABLE 6.11 (Cont.)

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

TABLE 6.11 (Cont.)

9.	**Solvent Circulation Pump**		
	Horsepower	1,282	
	Size exponent	0.79	
	Purchased cost of 300-hp pump in 1987 (includes motor, coupling, base; cast iron, horizontal)	$30,000	
	Module factor	1.5	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1993	360.4	
	Installed cost of solvent pump in 1993		**$165,631**
10.	**Compressor 1 for CO_2**		
	Inlet pressure (psia)	14.70	
	Outlet pressure (psia)	50.00	
	Compressor size (hp)	600.41	
	Purchased cost of centrifugal compressor in 1987 (includes electric motor drive and gear reducer)	$85,000	
	Size factor	1	
	Materials correction factor	1	
	Module factor	2.6	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1993	360.4	
	Installed cost of compressor in 1993		**$258,225**
11.	**Compressor 2 for CO_2**		
	Inlet pressure (psia)	4.00	
	Outlet pressure (psia)	50.00	
	Compressor size (hp)	120.54	
	Purchased cost of centrifugal compressor in 1987 (includes electric motor drive and gear reducer)	$50,000	
	Size factor	1	
	Materials correction factor	1	
	Module factor	2.6	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1993	360.4	
	Installed cost of compressor in 1993		**$151,897**
12.	**Solvent Refrigeration**		
	Refrigeration (tons)	434.53	
	Purchased cost in 1987	$230,000	
	Temperature correction factor	1.25	
	Module factor	1.46	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1993	360.4	
	Installed cost of solvent refrigeration in 1993		**$490,452**

TABLE 6.11 (Cont.)

13.	**CO_2 Product Gas Compressors**		
	Compressor 1 (hp)	2,913.98	
	Compressor 2 (hp)	2,913.98	
	Compressor 3 (hp)	2,913.98	
	Purchased cost of centrifugal compressor 1 in 1987	$750,000	
	Purchased cost of centrifugal compressor 2 in 1987	$750,000	
	Purchased cost of centrifugal compressor 3 in 1987 (includes electric motor drive and gear reducer)	$750,000	
	Size factor	1	
	Module factor	2.6	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1993	360.4	
	Installed cost of compressor 1 in 1993		**2,278,453**
	Installed cost of compressor 2 in 1993		**2,278,453**
	Installed cost of compressor 3 in 1993		**2,278,453**
	Total Direct Cost		**$10,518,378**
	Total Direct Cost for Three Trains		**$31,555,133**

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

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

7.1 Design Basis

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

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

7.2 Chilled Methanol Process for H_2S Recovery

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

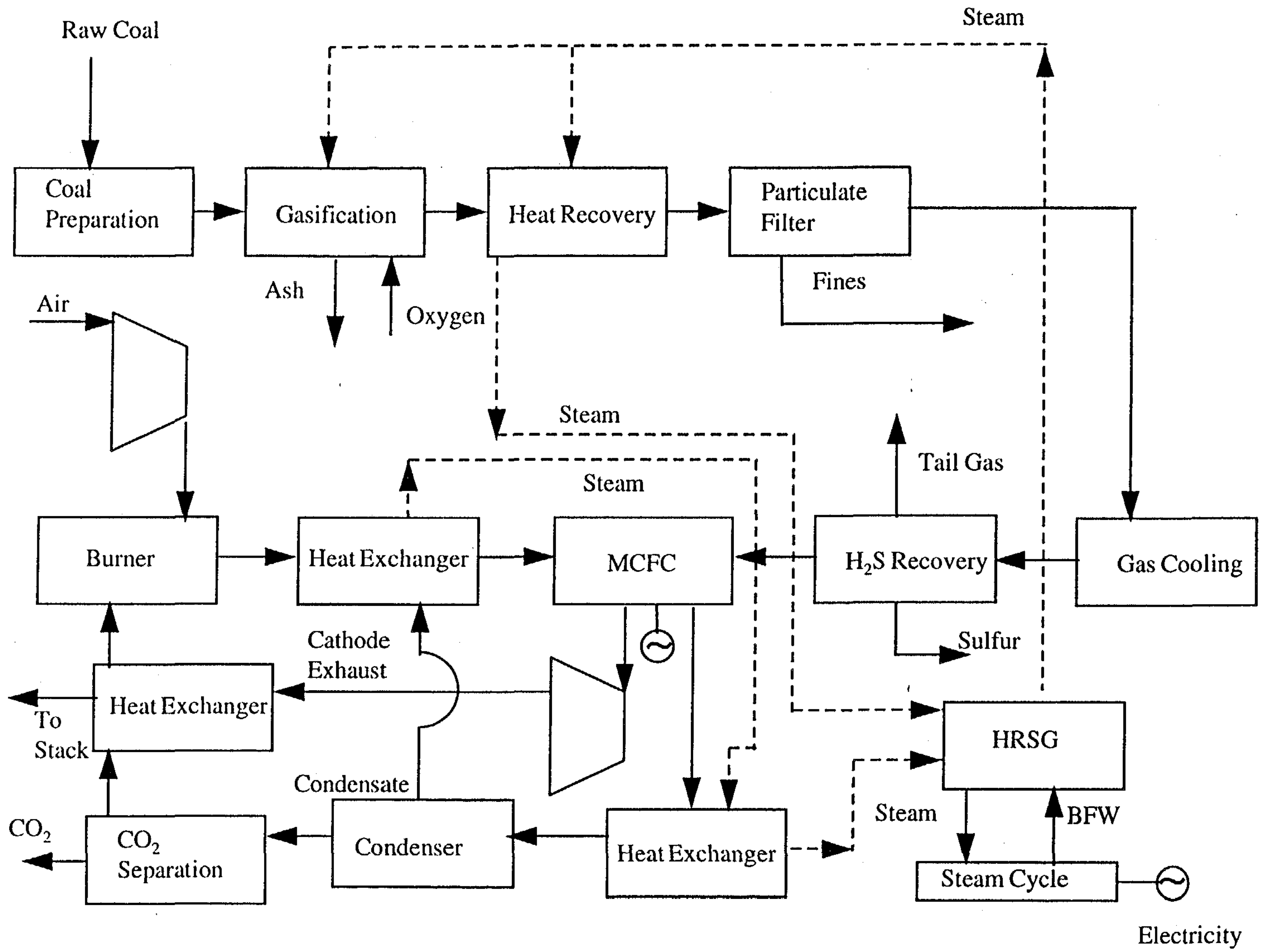

FIGURE 7.1 Block Diagram of the IGCC System with CO_2 Recovery Used in Cases 3 and 4

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

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

7.3 Molten Carbonate Fuel Cell System

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

7.4 Membrane System for CO_2 Recovery

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

7.5 Fuel Cell, Steam Cycle, and Plant Performance

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

7.6 Economics

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

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

Stream Data	Stream 12	Stream 13	Stream 14	Stream 15	Stream 16	Stream 17
Description of stream	Gases to membrane process	Gases from membrane process	Gases from heat exchanger 1	Air to compressor	Air from compressor	Gases from burner
Gases (lb·mol/h)						
CO	1,812.26	1,539.74	1,539.74	0.00	0.00	0.00
CO_2	8,724.66	4,180.59	4,180.59	0.00	0.00	5,720.33
H_2	1,693.50	1,692.69	1,692.69	0.00	0.00	0.00
H_2O	135.79	24.61	24.61	0.00	0.00	1,717.29
N_2	36.44	35.82	35.82	44,039.07	44,039.07	44,074.89
Ar	72.73	62.12	62.12	541.05	541.05	603.17
CH_4	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	11,779.07	11,779.07	10,162.86
COS	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	12,475.38	7,535.57	7,535.57	56,359.19	56,359.19	62,278.54
Liquids (lb·mol/h)						
H_2O	0.00	0.00	0.00	0.00	0.00	0.00
Temperature (°F)	150.00	154.88	600.00	81.00	713.05	1,355.19
Pressure (psia)	150.00	140.00	140.00	14.70	150.00	150.00
Enthalpy of stream (Btu/h) (reference, 32°F)	12,463,567	7,542,911	37,780,990	18,972,575	274,226,598	676,482,506

TABLE 7.2 (Cont.)

Stream Data	Stream 18	Stream 19	Stream 20	Stream 21	Stream 22	Stream 23
Description of stream	Gases from heat exchanger 3	Fuel cell cathode exhaust	Gases from expansion turbine	Gases from splitter to heat exchanger 1	Gases from heat exchanger 1	Gases from splitter to heat exchanger 4
Gases (lb·mol/h)						
CO	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	5,720.33	461.70	461.70	75.84	75.84	385.85
H_2	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	1,717.29	1,717.29	1,717.29	282.10	282.10	1,435.19
N_2	44,074.89	44,074.89	44,074.89	7,240.25	7,240.25	36,834.64
Ar	603.17	603.17	603.17	99.08	99.08	504.08
CH_4	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	10,162.86	7,533.54	7,533.54	1,237.55	1,237.55	6,295.99
COS	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	62,278.54	54,390.59	54,390.59	8,934.82	8,934.82	45,455.75
Liquids (lb·mol/h)						
H_2O	0.00	0.00	0.00	0.00	0.00	0.00
Temperature (°F)	980.33	1,300.00	667.24	667.24	200.00	667.24
Pressure (psia)	150.00	150.00	14.70	14.70	14.70	14.70
Enthalpy of stream (Btu/h) (reference, 32°F)	482,489,027	546,065,398	281,008,417	46,161,714	15,923,634	234,846,704

TABLE 7.2 (Cont.)

Stream Data	Stream 24	Stream 25	Stream 26	Stream 27	Stream 28	Stream 29
Description of stream	Gases from heat exchanger 4	Water from condenser	Water from pump	Steam from heat exchanger 5	Steam from heat exchanger 4	Steam from heat exchanger 3
Gases (lb·mol/h)						
CO	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	385.85	0.00	0.00	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	1,435.19	0.00	0.00	10,270.88	15,985.92	28,434.17
N_2	36,834.64	0.00	0.00	0.00	0.00	0.00
Ar	504.08	0.00	0.00	0.00	0.00	0.00
CH_4	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	6,295.99	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	45,455.75	0.00	0.00	10,270.88	15,985.92	28,434.17
Liquids (lb·mol/h)						
H_2O	0.00	22,923.59	34,923.59	24,652.71	18,937.68	6,489.42
Temperature (°F)	400.00	121.36	121.36	356.77	356.79	356.77
Pressure (psia)	14.70	1.76	146.96	146.96	146.96	146.96
Enthalpy of stream (Btu/h) (reference, 32°F)	145,783,678	36,915,818	56,173,167	366,812,595	455,875,619	649,869,098

TABLE 7.2 (Cont.)

Stream Data	Stream 30	Stream 31	Stream 32	Stream 33	Stream 34A	Stream 34B	Stream 34C
Description of stream	Steam from heat exchanger 2	Steam for heating feed to fuel cell	Steam to steam turbine	Steam turbine exhaust	Makeup water to pump	Makeup water to pump	Makeup water to pump
Gases (lb·mol/h)							
CO	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	34,923.59	12,000.00	22,923.59	21,596.52	0.00	0.00	0.00
N_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ar	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	34,923.59	12,000.00	22,923.59	21,596.52	0.00	0.00	0.00
Liquids (lb·mol/h)							
H_2O	0.00	0.00	0.00	1,327.07	13,443.34	12,000.00	1,443.34
Temperature (°F)	775.00	775.00	775.00	121.36	150.00	150.00	150.00
Pressure (psia)	146.96	146.96	146.96	1.76	146.96	146.96	146.96
Enthalpy of stream (Btu/h) (reference, 32°F)	889,843,009	305,756,514	584,086,495	435,269,723	28,553,650	25,488,000	3,065,650

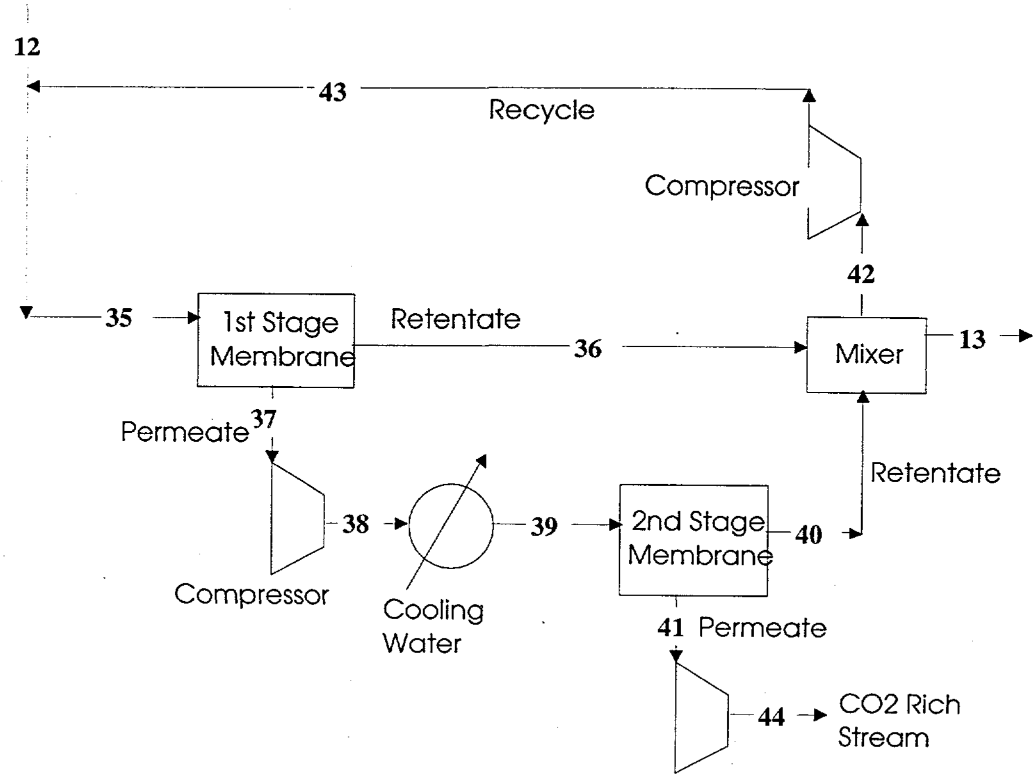

FIGURE 7.2 Flow Diagram of Membrane Process for CO_2 Removal in Case 4

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

Stream Data	Stream 12	Stream 13	Stream 35	Stream 36	Stream 37	Stream 38
Description of stream	Feed gas from fuel cell system	H_2-rich gas to fuel cell	To 1st-stage membrane	1st-stage retentate	1st-stage permeate	Gases from compressor
Gases (lb·mol/h)						
CO	1,812.26	1,539.74	2,042.34	1,296.30	746.04	746.04
CO_2	8,724.66	4,180.59	9,349.34	2,831.36	6,517.98	6,517.98
H_2	1,693.50	1,692.69	1,946.43	1,906.70	39.72	39.72
H_2O	135.79	24.61	139.47	14.94	124.53	124.53
N_2	36.44	35.82	41.79	36.72	5.07	5.07
Ar	72.73	62.12	82.01	52.51	29.50	29.50
CH_4	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
Total gas flow	12,475.38	7,535.57	13,601.38	6,138.53	7,462.84	7,462.84
Liquids (lb·mol/h)	0.00	0.00	0.00	0.00	0.00	0.00
Temperature (°F)	150.00	154.88	128.70	128.70	128.70	472.66
Pressure (psia)	150.00	140.00	150.00	140.00	25.00	150.00
Enthalpy of stream (Btu/h) (reference, 32°F)	12,462,913	7,542,574	11,053,736	4,694,179	6,359,557	31,178,763

TABLE 7.3 (Cont.)

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

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

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

TABLE 7.4 (Cont.)

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

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

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

[a] Includes H_2S recovery system energy use.

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

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

TABLE 7.6 (Cont.)

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

TABLE 7.6 (Cont.)

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

TABLE 7.6 (Cont.)

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

TABLE 7.6 (Cont.)

11.	**Pump**		
	Horsepower	106	
	Size exponent	1	
	Purchased cost in 1987	$12,000	
	(includes motor, coupling, base:cast iron, horizontal)		
	Module factor	1.5	
	CE index for process equipment in 1987	320	
	CE index for process equipment in 1995	373.9	
	Installed cost of pump in 1995		**$21,035**
12.	**Fuel Cell Stack**		
	Fuel cell power output (kW)	77,989	
	Unit cost per kilowatt	$180	
	Total cost		**$14,038,020**
13.	**Fuel Cell Invertor**		
	Unit cost per kilowatt	$100	
	Total cost		**$7,798,900**
14.	**Fuel Cell Controls**		
	Unit cost per kilowatt	$140	
	Total cost		**$10,918,460**
15.	**Fuel Cell and Component Assembly**		
	Unit cost per kilowatt	$110	
	Total cost		**$8,578,790**
	Total Direct Cost		**$93,973,387**
	Total Direct Cost for Three Trains		**$281,920,162**

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

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

TABLE 7.7 (Cont.)

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

8 CO_2 Pipeline Transport and Sequestering

8.1 Pipeline Transport of CO_2

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

8.2 CO_2 Sequestering

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

9 Conclusions — Energy Cycle/Economic Comparisons

9.1 Energy Consumption and CO_2 Emissions

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

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

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

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

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

9.3 Costs of Electricity

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

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

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

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

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

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

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

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

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

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

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

CO2 emission rate/net cycle	**0.368 kg CO2/kWh**
CO2 Sequestering power use	**75.03 MW**
Power use/CO2 in reservoir	**0.294 kWh/kg CO2**

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

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

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

	Electricity MW	CO2 release kg/h
Mining and Transport		
Raw Coal in Mine	-2.36	2,356
Coal Rail Transport	-0.05	523
Limestone Mining	-0.25	250
Limestone Rail Tansport	-0.02	156
Subtotal	-2.67	3,286
IGCC Power Plant		
Coal/Limestone Preparation	-3.49	11,374
Gasifier Island	-21.11	137
Power Island	-11.10	315,029
CO2 Recovery	-17.21	-285,499
CO2 Compression (to 2100psi)	-32.21	
Subtotal	-85.11	41,041
Power - Gas Turbine	274.39	
Power - Steam Turbine	186.50	
GROSS Power	**460.88**	
NET Power	**375.77**	
Pipeline/Sequester		
Pipeline CO2		285,499
Pipeline booster stations	-1.80	1,798
Geological reservoir (2% loss)	0.00	-279,789
Subtotal	-1.80	7,508
Energy Cycle Power Use	-89.58	
NET Energy Cycle	**371.30**	**51,834**
Derating from O2-blown Base Case	39.79	
Make-up Power	39.79	39,794
TOTAL	**411.09**	**91,628**
CO2 emission rate/net cycle	**0.223**	**kg CO2/kWh**
CO2 Sequestering power use	**39.79**	**MW**
Power use/CO2 in reservoir	**0.142**	**kWh/kg CO2**

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

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

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

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

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

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

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

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

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

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

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

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

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

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

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

COSTS OF ELECTRICITY

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

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

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

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

COSTS OF ELECTRICITY

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

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

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

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

COSTS OF ELECTRICITY

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

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

Net Power (MW) =	**330.00**
Capacity factor=	**65%**
Annual Net Power Production (MW)=	**1,879,020**
Net Energy-cycle Power (MW)=	295.02

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

COSTS OF ELECTRICITY

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

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

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

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

COSTS OF ELECTRICITY

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

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

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

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

COSTS OF ELECTRICITY

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

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

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

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

COSTS OF ELECTRICITY

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

TABLE 9.21 Summary of Comparative Costs of IGCC Systems

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

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