IV. APPENDIX A - CONCEPTUAL PROCESS FLOW SHEET

The Kellogg Coal Gasification Process, being developed with the support of the Office of Coal Research, uses molten sodium carbonate to gasify powdered coal with steam. This section of the report describes this process for a bituminous coal feed, though other coals can also be gasified. The description of the process includes a process flow sheet, a process description by section, projected economics for pipeline gas, and a review of the design bases.

A. PROCESS DESCRIPTION (COMMERCIAL PLANT)

The process flow sheet for a plant capable of producing 250,000,000 standard cubic feet per day (SCFD) of pipeline gas from bituminous coal is presented as Drawing No. 6026-3-5. Flow rates and compositions of the various numbered streams on this flow sheet are shown in Table A. A brief description of the flow sheet follows.

1. Section 100 - Coal Storage and Preparation

It has been assumed that during eight hours each day, coal is received by truck or conveyor belt from an adjacent coal mine at the rate of 1,650 tons per hour. The raw coal travels by belt conveyor to a coal distributing center, where about 550 tons per hour is dispatched for immediate use and the remainder is conveyed to the storage area.

Coal is distributed to several storage piles by a shuttle conveyor. These piles contain enough coal to permit the plant to operate for 30 days at normal capacity in the event the coal supply is cut off.

During the 16 hours each day that the mine is assumed not to be operating, coal is reclaimed from a ready storage pile for use in the process.

Coal from storage flows by belt conveyor to a Bradford Breaker, where it is reduced in size to about minus 2 inch. This coal is then fed to four parallel hammermills designed to reduce the coal particle size to minus 12 mesh. The crushed coal travels by conveyor belt to three parallel coal storage bunkers, from which it is carried by conveyor belt to Section 200.

2. Sections 200 and 600 - Gasification and Ash Removal

Bituminous coal from storage, ground to approximately -12 mesh in Section 100, is fed to a set of lock hoppers, F-201, a & b, whose purpose is to receive the coal at atmospheric pressure and to deliver it to the process at the operating pressure of about 415 psia. Each of the lock hoppers operates on a 30-minute cycle comprising the following steps:

a. Filling with coal at atmospheric pressure.

b. Pressurizing to about 450 psia with synthesis gas or product gas.

c. Discharging the coal into the steam line.

d. De-pressurizing

The lock hopper system is designed so that while one hopper is being filled and pressurized, the other is discharging coal into the process stream. This provides a continuous flow of coal to the gasifier.

Coal from F-201, a & b, flows by gravity at the rate of 1,100,000 pounds per hour into a steam line where it is mixed with 420 psia steam superheated to 1000°F in preheater C-201. The amount of steam used as a carrier gas is such that the coal will not be heated above about 500°F, thus preventing the coal from becoming plastic and sticky in the lines. Another stream of this 1000°F steam is used to pick up a recycle stream of $NaHCO_3$- Na_2CO_3 from lock hoppers, F-202, a & b. The two solids-containing steam streams then flow to the gasification section of gasifier-combustor D-201 together with the remainder of the 1000°F steam. The total steam fed to the gasifier is 1,000,000 pounds per hour. Of this, 920,600 pounds per hour is generated in waste heat boilers C-204, C-206, C-207, and 79,400 pounds per hour is generated in waste heat boilers in other sections of the plant.

In the gasifier, the steam and coal are heated to 1830°F by intimate contact with melt (a mixture of molten sodium carbonate, coal ash, and coal) and react according to the following reactions to produce synthesis gas.

$$C + H_2O + \text{heat} \rightarrow H_2 + CO \qquad (1)$$

$$C + 2H_2O \rightarrow CO_2 + 2H_2 \qquad (2)$$

$$CO + H_2O \rightleftarrows CO_2 + H_2 \qquad (3)$$

$$CO + 3H_2 \rightleftarrows CH_4 + H_2O \qquad (4)$$

$$C + 2H_2 \rightleftarrows CH_4 \qquad (5)$$

$$C_mH_n + H_2 \rightarrow CH_4 + C \qquad (6)$$

The raw synthesis gas leaves the melt at about 1830°F and 405 psia and flows through separator G-201 designed to remove entrained coal particles and melt droplets. The gas is then cooled to 700°F in exchangers C-201, C-203, and C-204 and leaves the gasification section at the rate of 118,105 moles per hour.

The primary gasification reaction (Reaction 1) is highly endothermic; the required heat of reaction is supplied as sensible heat of circulated melt. In the version shown in this flow sheet, the gasifier-combustor is divided into two sections by a vertical wall which is perforated below the liquid level. By proper choice of gas velocity, a difference in degree of aeration of the melt in the two sections is induced. This causes melt to circulate from the synthesis gas section, where it provides the reaction heat, to the heating or combustion section. In this latter section, heat is added to the melt by direct contact with hot combustion gases.

The flue gases which supply heat to the combustion section of the gasifier-combustor are generated by combustion of coal with air. Air for this purpose is compressed at the rate of 4,285,200 pounds per hour to 420 psia in compressor J-201 and is preheated to 1500°F in exchangers C-202 and C-203. Combustion occurs in direct contact with the melt, transferring heat to the reacting system efficiently.

The hot flue gases leave the bed at about 2200°F and 405 psia and pass through separator G-202 to remove entrained coal and melt. The gas then flows through exchangers C-206 and C-202 where it is cooled to 1500°F. This stream is expanded to substantially atmospheric pressure in J-202, which provides all the power required for air compression plus an additional 76,800 kw of electricity in generator J-204. Expanded flue gas at 615°F and about 18 psia is cooled to 180°F in exchangers C-207, C-208, and C-209 and is vented to the atmosphere at the rate of 4,655,300 pounds per hour.

The ash left in the melt by the combustion and gasification of the coal is allowed to build up to a level of 8 weight percent. A slipstream of the ash - carbon - Na_2CO_3 mixture is continuously withdrawn from D-201 and flows to E-601, where it is quenched to 444°F with a portion of a recycle solution saturated with $NaHCO_3$ at 100°F. Solid melt particles in the resulting slurry are ground in L-603 to facilitate dissolution of the melt stream. The remainder of the recycle $NaHCO_3$ stream is then mixed with this slurry, cooling the mixture to 228°F. This stream is then flashed to 16.1 psia in F-601, where sufficient holding time is provided to dissolve the Na_2CO_3. The bottoms slurry from F-601 is filtered in L-601 to separate the ash and carbon (and some undissolved Na_2CO_3) from the solution. This residue is sent to disposal.

The filtrate from L-601 is pumped to 30 psia in J-602 and is fed to carbonation tower E-602. In this tower the Na_2CO_3 is reacted with CO_2 from the gas purification system according to Reaction 7.

$$Na_2CO_3 + CO_2 + H_2O \rightleftarrows 2NaHCO_3 \qquad (7)$$

Overhead gas from the tower at 200°F is cooled to 95°F in C-601 to condense water, a portion of it is purged to remove the impurities brought in with the CO_2 stream, and the remainder is recycled through J-604 to the tower. Fresh CO_2 is added to the tower at the rate of 10,396 moles per hour.

The operating temperature at the bottom of E-602 is 100°F. At this temperature the $NaHCO_3$ (along with some ash) is precipitated. The resulting slurry is filtered in L-602, the $NaHCO_3$ solution being pumped up to 400 psia and recycled to E-601. The filter residue ($NaHCO_3$) is dried by hot gases flowing through the drying hood and is returned to lock hoppers F-202, a & b to be recycled to the gasifier-combustor.

Because of the large volumes of gases processed, Sections 200 and 600 consist of nine parallel trains of operating equipment.

3. Section 300 - Shift Conversion

Synthesis gas leaves Section 200 at the rate of 118,105 moles per hour and flows to Section 300, where about 68 percent of it is fed to shift converter D-301. Boiler feed water at 200°F is fed between beds of catalyst at the rate of 146,980 pounds per hour to absorb the heat of reaction. The remainder of the synthesis gas is bypassed and is combined with the shift effluent. This stream, at 735°F and having a H_2/CO ratio of about 3.15/1 (in preparation for methane synthesis) is cooled to 265°F in exchangers C-301, C-302 and C-303. Condensed water is separated from the gas in F-302, and the gas is further cooled to 100°F in C-304. The gas then flows to scrubber E-301, where the gas is countercurrently scubbed with clean water to remove trace amounts of ammonia which might be present.

Section 300 is designed as five parallel operating units.

4. Section 400 - Gas Purification

The gas purification section is designed to reduce the CO_2 concentration in the shifted synthesis gas to 1.0 mole percent and to reduce the total sulfur concentration to about 0.004 grains/100 SCF of gas. The purification sequence consists of the "Fluor Solvent CO_2 Removal Process" for CO_2 and bulk sulfur removal, followed by sponge iron (iron oxide) and activated carbon for residual H_2S and organic sulfur removal, respectively. All organic sulfur is assumed to be carbonyl sulfide (COS).

The shifted synthesis gas from Section 300 is cooled to 35°F in exchangers C-401, C-402, and C-403. Hydrate and ice formation is prevented by injection of lean ethylene glycol into the gas. The glycol is subsequently separated from the gas in F-402. The chilled gas is then fed to absorber E-401, where it is countercurrently contacted with lean Fluor solvent. Rich solvent leaving the bottom of E-401 is passed through a power recovery turbine on J-402 to a 20 psia flash drum, F-401, for partial desorption of gases. This is followed by a vacuum flash at 3 psia for complete regeneration of the solvent. Lean solvent from the vacuum flash step is recirculated to E-401 through J-401. Intermediate flash gas is exchanged against feed gas in C-402 and is combined with the final flash gas after this gas is compressed from 3 psia to 20 psia in J-402. About 60 percent of the flashed gases is purged from the system; the remainder of the gas is compressed to 30 psia in J-403 and flows to Section 600.

Clean synthesis gas leaves the absorber at 25°F and is heated to 80°F against feed gas in C-401. At this point the gas contains 1.0 percent CO_2, 6 grains H_2S/100 SCF, and 0.2 grains COS/100 SCF, and is next treated for removal of residual H_2S. The gas is contacted with finely divided iron oxide supported on wood chips, commonly called "sponge iron", which removes H_2S practically quantitatively by the classic iron oxide reaction, as follows:

$$Fe_2O_3 \cdot 6H_2O + 3H_2S \rightarrow Fe_2S_3 \cdot 6H_2O + 3H_2O \qquad (8)$$

Periodically, the sponge iron is revivified with air to carry out the following reaction:

$$Fe_2S_3 \cdot 6H_2O + 1\ 1/2O_2 \rightarrow Fe_2O_3 \cdot 6H_2O + 3S \qquad (9)$$

The sulfur is deposited in the amorphous form around the sponge iron particles and eventually encloses the sponge iron in an impervious mat, whereupon it must be discarded. Each of the five trains of equipment contains four parallel iron oxide drums, D-401 a-d, followed by an iron oxide guard chamber D-402.

Synthesis gas leaving the iron oxide drums is finally treated for COS removal by adsorption on fixed beds of activated carbon, D-403 a-c. The activated carbon drums are arranged in ten parallel trains, each train consisting of three drums which are manifolded for cyclic operation. Typically, for each train, gas flows through two vessels in series for twelve hours while the third is being regenerated with steam. During this period any COS leakage from the first drum is retained in the second. At the end of the period the first drum in line is taken off stream for regeneration, the second drum is moved into first position, and the freshly regenerated drum is placed into second position. This cycle is repeated every twelve hours.

The purified gas, containing about 1.0 percent CO_2 and an average of about 0.004 grains total sulfur per 100 SCF, proceeds to Section 500.

5. Section 500 - Methane Synthesis

In this section the purified synthesis gas is catalytically converted to methane as follows:

$$CO + 3H_2 \rightleftarrows CH_4 + H_2O + \text{heat} \qquad (4)$$

The methanation scheme selected for use here is the hot-gas-recycle system developed by the Bureau of Mines. The basis of this selection was an economic comparison with the other possible methanation systems, the results of which are summarized in the body of the report.

Gas from Section 400 at 80°F and 340 psia is preheated to 400°F in C-501. This stream is then mixed with recycle gas at 583°F and flows downward through iron catalyst reactor D-501. Each reactor contains four beds of iron catalyst in the form of lathe turnings, with internal cooling coils located between beds. As the gas flows through the reactors, heat of reaction is removed by generating steam in the coils of waste heat boilers C-502, C-503, and C-504.

Reactor effluent at 626°F is split into three streams; a hot recycle stream, a cold recycle stream, and a net reactor effluent. The cold recycle stream is cooled in exchangers C-505, C-512, and C-506 to 150°F and is separated from entrained condensate in F-503. The cold recycle is compressed by J-502, preheated to 410°F in C-505, and mixed with the hot recycle stream. Total recycle gas at a mix temperature of 573°F is compressed by J-501 and returned to the reactor inlets.

Net effluent from D-501, containing an appreciable quantity of hydrocarbons heavier than methane, flows to nickel catalyst reactor D-502, where the heavy hydrocarbons are cracked and the methanation reaction is completed. This reactor is physically separated into three sections, with each section containing an equal volume of Raney nickel catalyst. External coolers C-507, C-508, and C-509 are provided between sections to remove the heat of reaction which is used to generate steam. D-502 is also designed for recycle operation with the total recycle gas being fed to the top bed. Feed to the reactors is split into three streams which are fed to each of the three reactor sections.

Net product from D-502 is cooled to 193°F by heat exchange with fresh feed in C-501, further cooled to 100°F in C-510, and is separated from condensate in F-504. The gas is then compressed to 1015 psia in J-506, cooled to 100°F in C-511, and delivered to F-505, where entrained condensate is separated and purged from the system. Final product gas with a heating value of 914 BTU/SCF flows from F-505 to the gas mains at 1000 psia and 100°F at the rate of 250,000,000 SCFD.

Section 500 employs five parallel operating trains of equipment, as indicated on the flow sheet.

6. Section 1100 - Offsite Facilities

Section 1100 (not shown on Drawing No. 6026-3-5) includes facilities for:

a. generating steam and electric power

b. supplying cooling water, process water, and boiler feed water

c. providing miscellaneous services necessary to make this a completely self-sufficient plant

Steam generation facilities consist solely of a start-up boiler capable of producing 375,000 pounds per hour of 420 psia steam at 1000°F and 125,000 pounds per hour of 420 psia, saturated steam. Once the plant is in full operation, enough steam is generated by waste heat in the process to provide all the high and low pressure steam needed.

Electric power is produced at 13,800 volts by turbogenerators using condensing steam turbine drives. An electric substation is provided to reduce the voltage to 4160, 440, and 110 volts.

A complete utilities summary is included as Table B.

TABLE A

PROCESS STREAM BALANCE

250,000,000 SCFD PIPELINE GAS FROM BITUMINOUS COAL

FLOW RATE	Stream No. 1 Temperature - 1830°F Pressure - 405 psia			Stream No. 2 Temperature - 80°F Pressure - 14.7 psia			Stream No. 3 Temperature - 180°F Pressure - 15 psia			Stream No. 4 Temperature - 100°F Pressure - 370 psia			Stream No. 5 Temperature - 70°F Pressure - 20 psia		
	LB/HR	MOLS/HR	MOL. % (DRY)	LB/HR	MOLS/HR	MOL. % (DRY)	LB/HR	MOLS/HR	MOL. % (DRY)	LB/HR	MOLS/HR	MOL. % (DRY)	LB/HR	MOLS/HR	MOL. %
CO	859,000	30,630	33.5	---	---	--	---	---	--	486,000	17,330	16.5	4,380	156	1.0
CO_2	535,000	12,160	13.3	---	---	--	1,108,140	25,200	17.6	1,120,530	25,460	24.3	647,330	14,714	96.7
CH_4	111,500	6,970	7.6	---	---	--	---	---	--	111,500	6,970	6.7	1,590	99	0.7
H_2	82,044	41,022	44.9	---	---	--	---	---	--	108,644	54,322	51.9	166	83	0.5
N_2	9,950	355	0.4	3,250,000	116,000	79.0	3,255,460	116,195	81.3	9,950	355	0.3	17	0.6	—
H_2S	9,350	275	0.3	---	---	--	---	---	--	9,550	281	0.3	5,550	163	1.1
COS	780	13	0.01	---	---	--	---	---	--	430	7	0.007	253	4	0.03
O_2	---	---	---	989,000	30,900	21.0	46,600	1,457	1.0	---	---	---	---	---	—
SO_2	---	---	---	---	---	--	10,100	158	0.1	---	---	---	---	---	--
C_2^+	---	---	---	---	---	--	---	---	--	---	---	---	---	---	--
Total Dry Gas	1,607,624	91,425	100.0	4,239,000	146,900	100.0	4,420,300	143,010	100.0	1,846,604	104,725	100.0	659,286	15,219.6	100.0
H_2O	482,000	26,680		46,200	2,570		235,000	13,040		5,000	278		2,882	159	
Total Wet Gas	2,089,624	118,105		4,285,200	149,470		4,655,300	156,050		1,851,604	105,003		662,168	15,378.6	

FLOW RATE	Stream No. 6 Temperature - 80°F Pressure - 340 psia			Stream No. 7 Temperature - 626°F Pressure - 325 psia			Stream No. 8 Temperature - 210°F Pressure - 16.1 psia			Stream No. 9 Temperature - 95°F Pressure - 27 psia			Stream No. 10 Temperature - 100°F Pressure - 1010 psia		
	LB/HR	MOLS/HR	MOL. % (DRY)	LB/HR	MOLS/HR	MOL. % (DRY)	LB/HR	MOLS/HR	MOL. % (DRY)	LB/HR	MOLS/HR	MOL. % (DRY)	LB/HR	MOLS/HR	MOL. % (DRY)
CO	478,650	17,068	21.6	7,160	256	0.9	---	---	--	2,970	106	3.6	700	25	0.1
CO_2	35,200	800	1.0	43,700	993	3.6	6,460	147	100.0	113,560	2,581	88.3	28,600	650	2.4
CH_4	108,840	6,804	8.6	246,000	15,344	56.1	---	---	--	1,070	67	2.3	384,000	23,997	87.3
H_2	108,364	54,182	68.4	16,360	8,180	29.9	---	---	--	114	57	1.9	4,906	2,453	8.9
N_2	9,922	354	0.4	9,922	354	1.3	---	---	--	11	0.4	--	9,922	354	1.3
H_2S	---	---	--	---	---	--	---	---	--	3,740	110	3.8	---	---	—
COS	---	---	--	---	---	--	----	---	--	168	3	0.1	---	---	—
O_2	---	---	--	---	---	--	---	---	--	---	---	--	---	---	—
SO_2	---	---	--	---	---	--	---	---	--	---	---	--	---	---	—
C_2^+	---	---	--	121,586	2,240	8.2	---	---	--	---	---	--	---	---	—
Total Dry Gas	740,976	79,200	100.0	444,728	27,367	100.0	6,460	147	100.0	121,573	2,924.4	100.0	428,128	27,479	100.0
H_2O	198	11		81,000	4,500		19,400	1,080		1,660	92		464	26	
Total Wet Gas	741,174	79,219		525,728	31,867		25,860	1,227		123,233	3,016.4		428,592	27,505	

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TABLE B

UTILITIES SUMMARY

250,000,000 SCFD PIPELINE GAS FROM BITUMINOUS COAL

STEAM

420 psia, 449°F

GENERATION

Section		Item	Normal Generation Lb./Hr.
200	F-203	Steam Drum	920,600
300	F-301	Steam Drum	233,800
500	F-501	Steam Drum	1,304,700
	F-502	Steam Drum	258,000
			2,717,100

CONSUMPTION

Section		Item	Normal Consumption Lb./Hr.
200	D-201	Gasifiers	1,000,000
400	J-402	Turbine	167,000
	--	Misc. Fluor Requirements	22,600
	J-403	Turbine	21,700
		Regeneration Steam to D-403	57,600
500	J-501	Turbine	133,100
	J-502	Turbine	13,400
	J-505	Turbine	18,900
	J-506	Turbine	211,000
600	J-604	Turbine	20,700
1100	N-1101	Turbogenerator	1,051,100
			2,717,100

TABLE B (continued)

ELECTRIC POWER

GENERATION

Section	Item		Normal Generation kw
200	J-201	Generator	76,800
1100	N-1101	Turbogenerator	85,500
			162,300

CONSUMPTION

Section	Item		Normal Consumption kw
100	L-104	Bradford Breaker	125
	L-108	Hammermills	1,230
	--	Miscellaneous	145
200	J-203	Pump	161
300	J-301	Pump	41
500	J-503	Pump	228
	J-504	Pump	45
600	J-601	Pump	67
	J-602	Pump	336
	J-603	Pump	2,390
		Available	157,532
			162,300

TABLE B (continued)

COOLING WATER

PRODUCTION

Section	Item		Normal Production GPM
1100	L-1101	Cooling Towers	241,970

CONSUMPTION

Section	Item		Normal Consumption GPM
200	C-205-a	Intercooler	18,200
	C-205-b	Intercooler	20,400
		Total Section 200	38,600
300	C-304	Cooler	25,400
400	J-402	Surface Condenser	11,400
	J-403	Surface Condenser	1,480
		Total Section 400	12,880
500	C-506	Cooler	25,800
	C-510	Cooler	3,000
	C-511	Aftercooler	4,530
	J-501	Surface Condenser	9,060
	J-502	Surface Condenser	610
	J-505	Surface Condenser	1,280
	J-506	Surface Condenser	14,400
		Total Section 500	58,680
600	C-601	Condenser	33,200
	J-604	Surface Condenser	1,410
		Total Section 600	34,610
1100	N-1101	Surface Condenser	71,800

TABLE B (continued)

COOLING WATER CONSUMPTION SUMMARY

Section	Title	Normal Consumption GPM
200	Gasification	38,600
300	Shift Conversion	25,400
400	Gas Purification	12,880
500	Methane Synthesis	58,680
600	Ash Removal	34,610
1100	Offsite Facilities	71,800
	Total Consumption	241,970

COOLING WATER BALANCE

	GPM
Recirculated water	241,370
Make-up water from river water pumps	21,600
Total water to cooling towers	262,970
Water losses in cooling towers	21,000
Total water to process	241,970
Warm water returned to river	600
Recirculated water	241,370

TABLE B (continued)

BOILER FEED WATER

200°F, 15 psia

PRODUCTION

Section		Item	Normal Production Lb./Hr.
200	C-209	BFW Heater	1,217,980
300	C-303	BFW Heater	1,730,000
			2,947,980

CONSUMPTION

Section		Item	Normal Consumption Lb./Hr.
300	D-301	Quench	146,980
1100	--	De-aerators	2,801,000
			2,947,980

214°F, 15 psia

PRODUCTION

Section	Item	Normal Production Lb./Hr.
1100	De-aerators	2,801,000

CONSUMPTION

Section		Item	Normal Consumption Lb./Hr.
200	C-208	BFW Heater	1,160,000
300	C-302	BFW Heater	1,641,000
			2,801,000

TABLE B (continued)

400°F, 420 psia

PRODUCTION

Section	Item		Normal Production Lb./Hr.
200	C-208	BFW Heater	1,160,000
300	C-302	BFW Heater	1,641,000
			2,801,000

CONSUMPTION

Section	Item		Normal Consumption Lb./Hr.
200	F-203	Steam Drum	949,000
300	F-301	Steam Drum	241,000
500	F-501	Steam Drum	1,345,000
	F-502	Steam Drum	266,000
			2,801,000

BOILER FEED WATER BALANCE

LOSSES FROM SYSTEM

Section	Item		Lb./Hr.
200	D-201	Gasifier Steam	1,000,000
	F-203	Blow-Down	28,400
300	D-301	Quench	146,980
	F-301	Blow-Down	7,200
400	--	Miscellaneous Fluor Requirements	22,600
	D-403	Regeneration	57,600
500	F-501	Blow-Down	40,300
	F-502	Blow-Down	8,000
		Total	1,311,080

TABLE B (continued)

BOILER FEED WATER BALANCE (continued)

RECIRCULATION

Feed Water Make-up	1,311,080
Condensate from Surface Condensers	1,636,900
Total Boiler Feed Water	2,947,980

PROCESS WATER

GENERATION

Section	Item	Normal Generation Lb./Hr.
1100	Process water pumps	803,690

CONSUMPTION

Section	Item	Normal Consumption Lb./Hr.
300	Water to E-301	360,000
600	Wash water to L-601	443,690
	Total Process Water	803,690

B. ECONOMICS OF PIPELINE GAS

The cost of producing 250,000,000 SCFD of pipeline gas from bituminous coal according to the process sequence just described is calculated in Tables C and D, assuming 90 percent stream efficiency. The procedure used is in accordance with the OCR's tentative standard for cost estimating of pipeline gas plants (7).

Estimated capital investment is summarized in Table C. Shift catalyst and activated carbon are included in fixed investment because they have very long lifetimes. Total capital investment is about $140,000,000. It should be noted here that although the reaction system shown in Drawing No. 6026-3-5 indicates a combined gasifier-combustor, calculations have indicated that there is no economic difference between this type of construction and a separate gasifier and combustor as proposed for the pilot plant.

Estimated operating expenses and gas selling price are shown in Table D. Bituminous coal is charged at $4 per ton. Total operating expense is calculated to be 41.6¢/MSCF, and gas selling price, based on a 20-year average return on equity capital of about 9.4 percent, is 50.3¢/MSCF.

It should be noted here that a credit of 7.6¢/MSCF is taken in the economics for the excess power produced by the waste heat (steam) available from the process. It seems entirely logical to treat the plant as an energy center, supplying an area not only with pipeline gas but with a portion of its electric power needs as well. Credit is taken for the available power at 5 mills/kwh, the cost of producing it in a conventional 160-megawatt power plant burning coal at $4 per ton.

On the other hand, if exported power were not generated, rather substantial savings in capital equipment costs (turbines, heat exchangers, generators, etc.) could be realized. If this is taken into account total capital investment would be reduced from $140,275,000 to about $112,000,000. Using this adjusted investment and eliminating all power credit calculated gas manufacturing cost increases from 41.6¢/MSCF to 45.1¢/MSCF. Gas selling price, however, increases only slightly from 50.3¢/MSCF to 52¢/MSCF.

TABLE C

INVESTMENT SUMMARY

PIPELINE GAS FROM BITUMINOUS COAL

BASES: 250,000,000 SCFD OF PIPELINE GAS
90% STREAM EFFICIENCY

Section	Title	Bare Cost*
100	Coal Storage and Preparation	$ 4,556,300
200	Gasification	56,408,100**
300	Shift Conversion	3,874,400
400	Gas Purification	19,909,200
500	Methane Synthesis	10,373,100
600	Ash Removal	4,662,900
1100	Offsite Facilities	14,491,000
	Total Bare Cost	$114,275,000
	Interest During Construction and Contractor's Overhead and Profit	20,112,000
	Total Fixed Investment . . .	$134,387,000
	Working Capital	
	30 days coal inventory	$ 1,580,000
	30 days carbonate inventory	211,000
	30 days catalyst inventory	73,000
	Catalyst charge	294,000
	Accounts receivable at 11% of total operating expense	3,730,000
	Total Working Capital	$ 5,888,000
	TOTAL CAPITAL INVESTMENT .	$140,275,000

*Bare cost includes materials, freight, construction labor, field administration and supervision, insurance during construction, cost of tools, field office expense, and cost of home office engineering and procurement.

TABLE C (continued)

**The gas expanders required in this process operate at pressures and temperatures higher than those in commercial use today. The costs used in this estimate were based on internally available information on gas turbines and high-temperature expanders as well as on recommendations of machinery manufacturers. The costs include a reasonable allowance for developmental costs distributed over a large number of machines but do not include a heavy burden as would be the case for the first machines built.

TABLE D

ESTIMATED ANNUAL OPERATING COST

AND GAS SELLING PRICE

PIPELINE GAS FROM BITUMINOUS COAL

BASES: 250,000,000 SCFD OF PIPELINE GAS
90% STREAM EFFICIENCY

ITEM	$/YEAR	¢/MSCF
Bituminous coal at $4 per ton	$17,410,000	21.1
Sodium carbonate make-up at 1.55¢/lb.	2,320,000	2.8
Miscellaneous chemicals	173,000	0.2
Sponge iron make-up	58,000	0.07
Methanation catalyst make-up	743,000	0.9
Direct operating labor @ $3.20 per man-hour	1,400,000	1.7
Power credit at 5 mills per kwh	(6,240,000)	(7.6)
Maintenance at 3% of bare cost	3,430,000	4.2
Supplies at 15% of maintenance	515,000	0.6
Supervision at 10% of operating labor	140,000	0.2
Payroll overhead at 10% of operating labor + supervision	154,000	0.2
General overhead at 50% of maintenance + supplies + operating labor + supervision	2,740,000	3.3
Plant Operating Expenses......	$22,843,000	27.7
Depreciation at 5% of fixed investment	6,720,000	8.2
Local taxes and insurance at 3% of fixed investment	4,030,000	4.9
Sub-total....................	$33,593,000	40.8
Contingencies	672,000	0.8
Total Operating Expense . . .	$34,265,000	41.6
20-year average total revenue requirement	$41,500,000	
AVERAGE GAS SELLING PRICE		50.3

The figures given on Table D were obtained before the data concerning the melt's ability to retain sulfur were available. Therefore, as can be seen from the flow sheet, the plant contains equipment for sulfur removal from the synthesis gas. Elimination of this equipment results in a decrease in gas selling price of about 2¢/MSCF. In addition, if the sulfur were recovered via the Claus process and could be sold at $45/ton, an additional savings of 2.4¢/MSCF could be realized. The combination of these reductions would result in a pipeline gas selling price of about 46¢/MSCF.

C. DESIGN BASES

The process flow sheet which has served for development and evaluation of the Kellogg Coal Gasification Process, shown as Drawing No. 6026-3-5, has been set up using a number of design bases. This section of the report presents these bases. A discussion of their impact on the economics of the process is presented in the body of the report. It is this analysis that has served to guide the organization of the experimental program.

The important design bases are listed below:

1. Monofrax A is a satisfactory material of construction to be in contact with the melt;

2. Gasification and combustion rates of about 21 and 12 pounds C/hr/CF melt, respectively, can be obtained at the following conditions:

   a. 1830°F gasification temperature, 1900°F combustion temperature

   b. 400 psig reactor pressure

   c. 8 percent ash and 4 percent carbon in melt

   d. ~15 feet expanded bed height

   e. outlet velocities of 1.2 and 1.75 fps in gasifier and combustor, respectively.

3. Carbon can be burned to CO_2 and the subsequent heat can be transferred to the melt at efficiencies of about 85 percent;

4. Melt can be adequately circulated (with a minimum of process gas) from combustor to gasifier in order to transfer the required heat;

5. Sodium carbonate losses are 2.4 percent per pass of the carbonate fed to the ash removal system;

6. Steam/carbon feed ratio = 2 lb/lb (corresponding to about a 70 percent steam conversion);

7. Entrained and/or volatilized sodium, silica and tar can be removed from the synthesis and flue gases (e.g., by quenching or using separators) to a point where the flue gas expander can operate and where no solids will deposit on downstream surfaces.