Table IV-F-2c

COAL GASIFIER PRODUCT SURVEY

WESTERN COALS

	Lurgi	CO_2 Acceptor	Koppers-Totzek	Winkler
Clean Gas Analysis				
H_2 (Vol.%)	41.3	53.8	33.0	41.9
CO	15.2	17.0	58.9	33.4
CO_2	30.6	6.7	7.1	20.5
CH_4	11.3	20.9	-	3.1
C_2+	0.5	0.4	-	-
Inerts	1.1	1.2	1.0	1.0
Coal Basis	Rosebud	Lignite	Sub - Bituminous	Wyoming
Clean Gas Properties				
H_2/CO Ratio	2.72	3.16	0.56	1.25
H_2+CO (Vol.%)	56.5	70.8	91.9	75.3
HHV (BTU/SCF)	305	445	295	275
Pressure (psia)	360	300	15	30
Separate Shift & Methanation				
Steam Required (Lbs/MM BTU SNG)	5.3	-	156.6	67.7
Combined Methanation/Shift				
Steam Required (Lbs/MM BTU SNG)	-	-	88.4	20.4
Acid Gas Removal				
CO_2 Removed (Lbs/MM BTU SNG)	139.8	19.6	217.2	185.8

It was possible to categorize synthesis gases on the basis of the amount of water gas shift required prior to or in conjunction with methanation. With regard to the LPM/S process, these gases could be grouped into four classes on the basis of H_2/CO mole ratios as follows:

- $H_2/CO > 3$; Hygas steam-iron and CO_2 Acceptor.
- $2 < H_2/CO < 3$; Lurgi.
- $1.4 < H_2/CO < 2$; Synthane.
- $H_2/CO < 1.4$; Hygas electrothermal and steam-oxygen, Bi-Gas, Koppers-Totzek and Winkler.

The first group represents those gases which require no shift and, thus, the LPM process can be used by itself. In the second group are those gases which require shift, but, as regards the LPM/S process, no steam injection is necessary. The quantity of catalyst in the LPM reactors must be increased with decreasing H_2/CO ratio, but no other process modifications are necessary. The third group encompasses those gases which require supplemental steam injection into the LPM/S reactors. In the last group are synthesis gases which would require relatively high steam injection rates and have low reaction rates because of the low H_2/CO ratios. When handling these gases, it is probably more economical to shift a portion of the gas to achieve an overall 1.4 H_2/CO ratio and then feed it to the LPM/S reactors.

It should be pointed out that the grouping of synthesis gases as listed above was somewhat arbitrary at the time since optimal steam injection rates over the whole range of H_2/CO ratios were still undetermined. Experimental data had been obtained in both bench scale and process development units for 1.4 H_2/CO to 1.5 H_2/CO synthesis gas feeds. Steam rates from zero to 0.3 mole H_2O/mole of $H_2 + CO$ were investigated.

It has been assumed in this analysis that the steam injection rate falls off linearly up to a ratio of $2H_2/CO$ where no steam is required. However, it is possible that the grouping as listed above is conservative. The $2H_2/CO$ break point could be as low as $1.5H_2/CO$ and the $1.4H_2/CO$ point could be depressed as low as $1H_2/CO$. This would increase the economically useful range of the LPM/S process. However, precise determination of these points would require extensive economic optimization and was beyond the scope of the present study since an experimental basis for such optimization was unavailable.

Steam requirements for separate shift to 3/1 H_2/CO prior to methanation have been determined and are listed in the three tables for all of the synthesis gases studied. The quantity of steam was based on two moles of steam per mole of CO fed to the shift reaction and a CO conversion of 95 percent. Steam for the combined methanation/shift reaction (also shown in the tables) was determined by first shifting the gas to a 1.4 H_2/CO ratio and adding supplemental steam to raise the H_2O/CO ratio to 0.24 if necessary. Steam addition rates for gases not requiring preliminary shift were apportioned linearly as described above. It can be seen from Tables IV-F-2a, b, and c that a substantial savings in steam can be realized in virtually all cases by using the combined Liquid Phase Methanation/Shift process as opposed to conventional shift and methanation processes. The steam savings ranges from 40-70 percent for H_2/CO feed ratios less than 1.4 to approximately 80 percent for feed ratios between 1.4 and 2.0. For ratios above 2.0, all the steam is saved. In terms of a 250 MMM BTU/Day SNG plant, this savings can be as high as 700,000 Lbs/Hr of steam or, possibly, $15,000,000 per year in operating costs.

There are other factors which differentiate the various synthesis gases and affect the size of the required methanation plant. Most important among these is the amount of methanation required. From the tables, it can be seen that the volume percent of H_2+CO can range from 45 to

92. The quantity of CO_2 to be removed from the product gas can also range from 20 to 220 Lbs/MM BTU SNG. Methanation reaction and heat recovery equipment sizing will be affected by these parameters. Finally, the available synthesis gas pressure will have a bearing on methanation economics mainly because of gas compression costs, but also because of methanation equipment sizing and cost as a function of pressure.

It is impossible to study Liquid Phase Methanation/Shift economics as it is applied to each synthesis gas. However, it is possible to bracket LPM/S economics with two typical case studies. The first of these involves a synthesis gas which requires little or no steam addition as exemplified by a $2H_2/CO$ ratio. Such gases usually contain substantial quantities of methane and CO_2; i.e., Lurgi and Synthane. The bulk of the CO_2 could be removed prior to entering the methanation section and the methane content will contribute to smaller reactors and heat recovery equipment. The second case study involves a synthesis gas which requires steam addition and substantial shifting such as a $1H_2/CO$ ratio. Several gasifiers in this category produce low methane content gases and, therefore, this case was formulated to also illustrate this type of synthesis gas.

The two economic studies outlined above are presented in the following four sections. A feed gas pressure of 500 psia has been chosen for both cases, but a pressure optimization study is carried out for the $2H_2/CO$ ratio case.

3. LPM/S Process Flowsheet With $2H_2/CO$ Feed Gas

a. General Description

Figure IV-F-3a-1 is a process flowsheet for the Liquid Phase Methanation/ Shift process as applied to a $2H_2/CO$ feed gas. The synthesis gas feed composition was assumed to be similar to that produced by the Lurgi

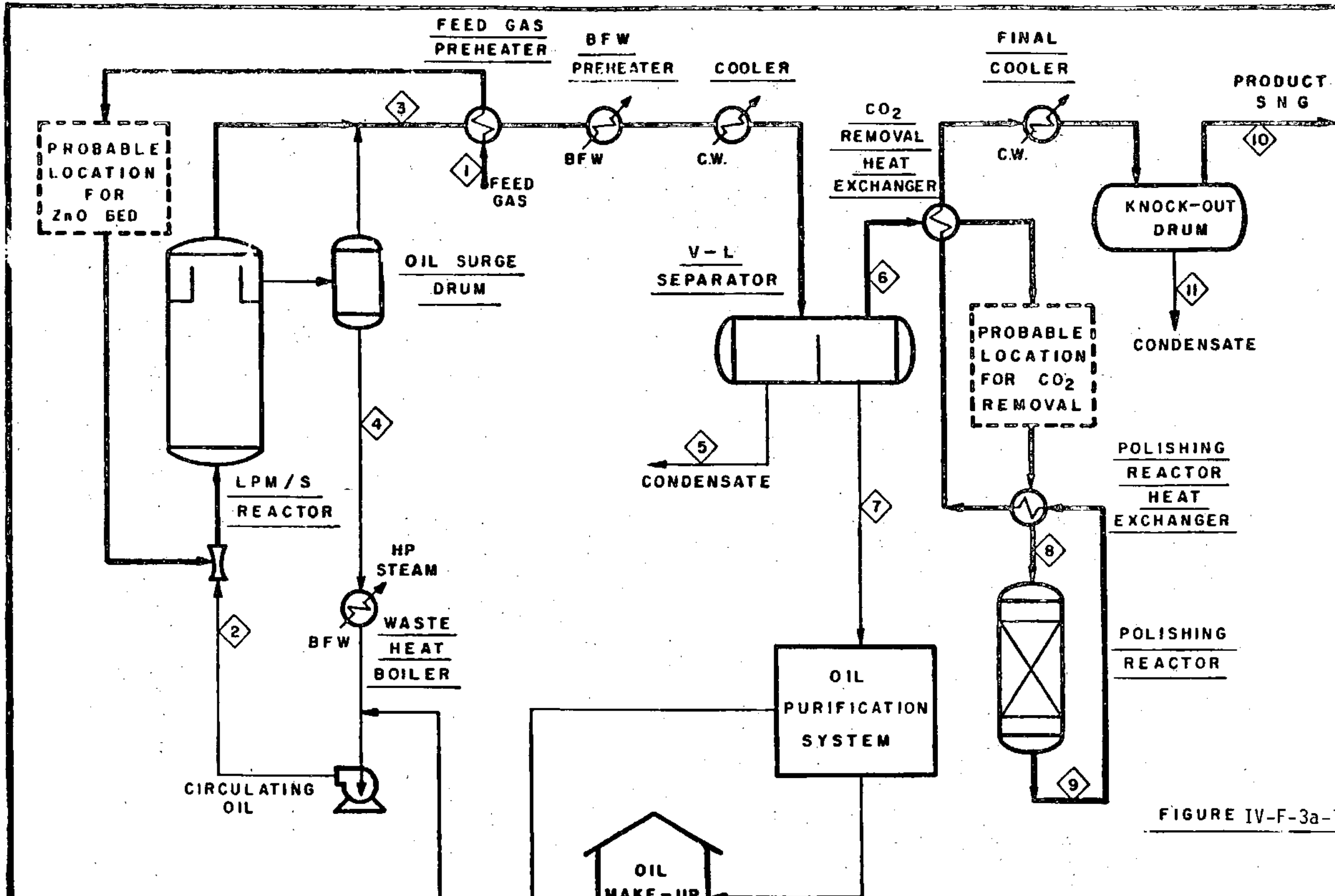

FIGURE IV-F-3a-1

TITLE LIQUID PHASE METHANATION / SHIFT PROCESS FLOWSHEET or Synthane processes after acid gas removal. The simplified composition is shown below:

Component	Volume Percent
CO	27.0
H_2	54.0
CH_4	19.0
	100.0

The synthesis gas feed, available at 120°F and 500 psia, is preheated by exchange with reactor product gas and passes through a desulfurization vessel. The gas enters the bottom of the LPM/S reactor and flows upward through a liquid fluidized catalyst bed. Simultaneous methanation and shift reactions occur without the need for steam addition and the heat of reaction is absorbed by sensible heat rise of the liquid. The catalyst particles remain in the reactor. A vapor-liquid disengaging zone is provided at the top of the reactor and the effluent gases are separated from the liquid. An oil surge drum is provided for the dual purpose of handling fluctuations in liquid level and supplying additional surface area for vapor-liquid disengagement. The liquid is sent through a heat exchanger where the heat of reaction is removed by generating steam and then recirculated to the bottom of the reactor by a liquid circulating pump. The liquid circulation rate controls the degree of catalyst fluidization in the reactor.

The LPM/S reactors are designed for 95-98 percent conversion of the carbon monoxide. The effluent gas is cooled in a series of heat exchangers first by heating feed gas, then by preheating boiler feedwater and finally with cooling water or air. Water of reaction and any vaporized liquid are condensed at 120°F and flow together with the cooled gas to a vapor-liquid separator. The water is phase separated from the hydrocarbon liquid and enters the condensate return system. The hydrocarbon liquid is pumped back into the reactor circulating oil system. An oil purification system, consisting of a small vacuum lower to remove both light and heavy ends, is provided. Both these products may be formed from slight thermal degradation of the hydrocarbon liquid. Either liquid from the vapor-liquid separator or a slip stream from the reactor circulating oil can be sent to the oil purification system.

Gas from the vapor-liquid separator is reheated to 250°F and sent to a CO_2 removal system where the CO_2 formed by the shift reaction is removed. The gas, saturated with water vapor at 250°F from the wash step of the CO_2 removal unit, is then heated to 550°F and sent to a polishing reactor for final conversion of the CO. This unit contains a conventional methanation catalyst in a fixed bed reactor. The polishing reactor product is cooled by exchange against polishing reactor feed and then cooled to 120°F using cooling water. Additional water that was produced is separated in a knockout drum and the product SNG sent to a final drying section.

b. Material Balance

A material balance is presented in Table IV-F-3b-1 for a commercial plant producing 250 MMM BTU/Day of SNG. A CO conversion of 97 percent has been assumed in the LPM/S reactors. The water gas shift reaction takes place simultaneously, thus providing a product containing $7H_2/CO$.

This gas, after removing approximately 85 percent of the CO_2 formed in the LPM/S reactors, is passed through the polishing reactor where it has been assumed that essentially all the remaining CO and H_2 is converted to CH_4. Part of the remaining CO_2 is also converted. Side reactions to ethane and propane are assumed negligible. Solubilities of gaseous components in the hydrocarbon liquid as well as hydrocarbon liquid decomposition products themselves have been ignored in the material balance.

Table IV-F-3b-1

MATERIAL BALANCE - LPM/S

$2H_2/CO$ FEED GAS

	Gas Feed	Circulating Oil Feed	LPM/S Reactor Product	Oil From Reactor	Water Condensate	Gas to CO_2 Removal
	1	2	3	4	5	6
Moles/HR						
CO	19,020		570.6		-	570.6
H_2	38,040		3,994.2		-	3,994.2
CO_2	-		5,325.6		-	5,325.6
CH_4	13,384		26,507.8		-	26,507.8
H_2O	-		7,798.2		7,674.8	123.4
Oil	-	94,000	2,131.1	91,868.9	-	-
Total M/Hr	70,444	94,000	46,327.5	91,868.9	7,674.8	36,521.6
MM SCFD	640.8		421.4			323.2
FLOW, GPM		87,188		85,212		
T (^oF)	120	593	650	650	120	120
P (psia)	500	500	475	475	470	470

Table IV-F-3b-1 (con't.)

	Condensed Oil Return	Gas to Polishing Reactor	Polishing Reactor Product	Cooled Product	Water Condensate
	7	8	9	10	11
Moles/HR					
CO		569.2	-	-	-
H_2		3,992.6	-	-	-
CO_2		841.2	270.0	270.0	-
CH_4		26,163.2	27,303.7	27,303.7	-
H_2O		2,238.6	3,950.3	93.5	3,856.8
Oil	2,131.1	-	-	-	-
Total M/Hr	2,131.1	33,804.8	31,524.0	27,667.2	3,856.8
MM SCFD		307.5	286.7	251.7	
FLOW, GPM	1,976				
T (^oF)	120	550	810	120	120
P (psia)	470	450	440	435	435

c. Equipment Sizing and Cost Estimation

All battery limits equipment, as shown in Figure IV-F-3a-1 were sized and priced out for inclusion in the plant investment except for the ZnO guard system and the acid gas removal system (shown in dotted lines in Figure IV-F-3a-1). The oil purification system and oil make-up system are included under miscellaneous items (taken as 10 percent of BLCC) which includes other minor equipment not shown in Figure IV-F-3a-1. Sizing was based upon two parallel lines of 125 MMM BTU SNG/Day with reactors further subdivided to at least two per line. Single spares for all the pumps were included.

To size the LPM/S reactors, a kinetic rate constant of

$$\frac{K}{K_H\,(M/\rho_L)} = 2 \times 10^{-6} \quad \frac{\text{gm. mole}}{\text{Sec-Atm-gm. cat.}}$$

was chosen based upon available data. Then, using the first order rate expression [(a)]

$$\mathrm{Ln}\left(\frac{1}{1-X_T}\right) = \frac{K}{(K_H\,(M/\rho_L)}\ \frac{W}{F}\ \frac{(P_T - P^*)}{(1-(\frac{R+1}{2})\,Y^o_{CO})}\ \frac{R}{3}$$

a gas space velocity of 1590 Hr^{-1} was calculated for the conditions shown in the material balance. At 500 psia, 13,930 Ft^3 of catalyst are required when the exit temperature is 650^oF. Allowing 100 percent bed expansion using Witco 40 mineral oil as the liquid medium, four reactors, each 17.4' diameter by 35' T-T, are required. The reactor sizing includes 6' for disengagement. The oil surge drums are sized to contain the total reactor liquid inventory if necessary. Vessel thickness are based upon an allowable stress of 15,000 psi for 1 percent Cr- $\frac{1}{2}$ percent Mo steel at $700^{\circ}F$ and a design pressure of 10 percent above operating pressure.

The polishing reactors are sized by assuming a gas space velocity of 10,000 Hr^{-1} at 1000 psi and scaling down to 5/8 of that value at 500 psi (Sv = 6250 Hr^{-1}). Catalyst requirements are thus 1900 Ft^3 and four reactors, 6' ID by 17'T-T, can be used for this purpose. The reaction vessels themselves are 1 percent Cr - $\frac{1}{2}$ percent Mo. The design pressure used was 10 percent above operating pressure and the design temperature was $900^{\circ}F$ giving an allowable stress of 13,100 psi.

Installed costs were based on first quarter 1976 prices and an installed/purchased equipment ratio of 3.3 was used on all items. The total BLCC for this plant was $20,410,000. A detailed list of major equipment and costs is given in Table IV-F-3c-1. The initial capital cost also includes catalyst at $200/Ft³ and mineral oil at $1/Gal for a total of $3,270,000.

d. Annual Costs

Annual catalyst replacement costs are based upon a one-year catalyst life in the LPM/S reactors and a five-year life in the polishing reactors at a net catalyst value of $200/ft³. Oil replacement is also required on an annual basis.

Annual capital related charges are usually proportional to the total fixed capital investment. A simple approach to capital charges is to take a lumped percentage of the total investment. Using the "Coal Gasification Commercial Concepts Gas Cost Guidelines"(1) as a basis, the annual capital related charges are equivalent to 20-25 percent of the fixed capital investment (including initial catalysts and chemicals) when the utility financing method is applied to methanation plants. For private investor financing at 12 percent DCF, the equivalent lumped

(1) Derivation & definition of terms in earlier reports.

Table IV-F-3c-1

MAJOR EQUIPMENT AND CAPITAL COSTS

250 MMM BTU/DAY SNG AT 500 PSIA

Item	Unit Description	Number of Units	Installed Cost
LPM/S Reactors	17.4' ID x 35' T-T Thickness = 3.6" Material = 1%Cr - $\frac{1}{2}$% Mo	4	$5,640,000
Oil Surge Drums	12' ID x 29' T-T Thickness = 2.5" Material = 1%Cr - $\frac{1}{2}$% Mo	2	$ 980,000
V-L Separator	8' ID x 24' T-T Thickness = 1.5" Material = CS	2	$ 330,000
Polishing Reactor	6' ID x 17' T-T Thickness - 1.4" Material = 1%Cr - $\frac{1}{2}$% Mo	4	$ 350,000
Knockout Drum	6' ID x 12' T-T Thickness = 1.2" Material = CS	2	$ 100,000
Circulating Oil Coolers	9,300 Ft^2 Q = 1049 MM BTU/Hr (Tot.) Material = 1%Cr - $\frac{1}{2}$% Mo	10	$3,620,000

Table IV-F-3c-1(con't.)

Feed Gas Preheaters	5,700 Ft^2 Q = 255 MM BTU/Hr (Tot.) Material = 1%Cr - $\frac{1}{2}$% Mo	4	$1,080,000
BFW Preheaters	9,000 Ft^2 Q = 239 MM BTU/Hr (Tot.) Material = 1%Cr - $\frac{1}{2}$% Mo	2	$ 730,000
Coolers	8,000 Ft^2 Q = 120 MM BTU/Hr (Tot.) Material = 1%Cr - $\frac{1}{2}$% Mo	2	$ 640,000
CO_2 Removal Heat Exchanger	900 Ft^2 Q = 43 MM BTU/Hr Material = 1%Cr - $\frac{1}{2}$% Mo	2	$ 170,000
Polishing Reactor Heat Exchanger	2,200 Ft^2 Q = 97 MM BTU/Hr Material = 1%Cr - $\frac{1}{2}$% Mo	2	$ 310,000
Final Cooler	4,900 Ft^2 Q = 115 MM BTU/Hr Material - 1%Cr - $\frac{1}{2}$% Mo	2	$ 480,000
Circulating Oil Pumps	21,900 GPM x 30 psi 590 Horsepower Material = 1%Cr - $\frac{1}{2}$% Mo	4 + Spare	$2,550,000
Condensed Oil Return Pumps	1,070 GPM x 100 psi 90 Horsepower Material = 1%Cr $-\frac{1}{2}$% Mo	2 + Spare	$ 390,000

Table IV-F-3c-1(con't.)

BFW Pumps	570 GPM x 950 psi 450 Horsepower Material = CS	4 + Spare	$1,240,000
Miscellaneous	Oil Makeup, Purification, Etc.		$ 1,800,000
	Total BLCC		$20,410,000

annual capital related charge is 40-45 percent of the fixed capital investment. Therefore, the economics of methanation units were herein based upon capital related charges of 20 percent and 40 percent of the fixed investment which would be representative of the range encountered when using the more detailed procedures of Ref. (1).

Utilities were based on the following rates:

Power	\$0.012/KWH
Fuel Gas	\$2.25/MM BTU
900 spia, 533°F sat'd. steam	\$2.00/M Lbs
Boiler feed water	\$0.55/M Gal
Cooling Water	\$0.03/M Gal

All pumps and compressors were assumed to be steam turbine drive units. Steam generated from the methanation/shift exothermic heat of reaction was 566 ST/Hr at 900 psia (sat'd.), part of which is required to drive the internal power users. Supplementary gas fired steam superheaters are provided to improve the power generation efficiency of internally used steam. A summary of utility requirements and costs is given in Table IV-F-3d-1.

The resultant annual operating costs for the 250 MMM BTU/D SNG plant utilizing the LPM/S process at 500 psi are summarized in Table IV-F-3d-2. Because of the large quantities of surplus steam generated, the methanation plant operates at a net credit of \$1.7MM/Yr for the case of private investor financing and a net credit of \$7.6 MM/Yr for the case of utility financing when labor costs are ignored and no product SNG compression is included. Raw material costs have also been excluded. The overall economics of this case could be further improved by decreasing the conversion in the LPM/S reactors from 97 percent to approximately 95 percent. This would increase the work performed

Table IV-F-3d-1

UTILITY REQUIREMENTS
250 MMM BTU/D SNG
$2H_2/CO$ FEED GAS AT 500 PSI

		$/Yr
Fuel		
Superheating saturated steam to 900°F for internal steam drives: 9.1 MM BTU/Hr @$2.25/MM BTU		164,000
Boiler Feedwater	135,650 Gal/Hr @ $.55/M Gal	597,000
Cooling Water	940,000 Gal/Hr @ $.03/M Gal	226,000
Steam		
Generated: 900 psia, sat'd:	1,132,700 Lbs/Hr	
Steam to internal drives:	35,600 Lbs/Hr[a]	
Net	1,097,100 Lbs/Hr	
	@ $2.00/M Lbs	(17,552,000)
Total Utility Cost (Credit)		(16,565,000)

(a) Total Horsepower Required: 4,960 HP
Power: In terms of internal steam drives using 850 psig, 900°F steam to 4" Hg abs. (condensing): 35,600 Lbs/Hr

Table IV-F-3d-2

ANNUAL OPERATING COSTS
250 MMM BTU/D SNG
2 H_2/CO FEED GAS AT 500 PSI

	$M	
BLCC	20,410,000	
Associated Offsites @30% of BLCC	6,120,000	
Initial Catalyst and Oil	3,270,000	
Total Fixed Capital Investment	29,800,000	
Private Financing Equivalent	**\$/Yr**	**¢/MM BTU**
Capital Related Costs (@40%)	11,920,000	
Catalyst and Oil Replacement	2,970,000	
Utilities	(16,570,000)	
Annual Operating Cost	(1,680,000)	(2.0)
Utility Financing Equivalent		
Capital Related Costs (@20%)	5,960,000	
Catalyst and Oil Replacement	2,970,000	
Utilities	(16,570,000)	
Annual Operating Cost	(7,640,000)	(9.3)

in the polishing reactors and slightly decrease the surplus steam, but, the annual operating cost would also be less.

4. Pressure Optimization With A $2H_2/CO$ Feed Gas

A commercial plant producing 250 MMM BTU/Day of SNG was selected for this study. The feed composition and processing requirements remained as stated in Section IV-F-3. However, for uniformity, the feed gas was assumed to be available at 100 psia and the product had to be delivered at 1000 psia. This required feed gas compression to the processing pressure and product compression to the delivery pressure. Multistage or single stage steam driven centrifugal compressors were specified for these purposes.

Three operating pressure levels were examined, 300, 500 and 900 psia. For each case, capital costs were generated as outlined in Section IV-F-3 and compression costs were added as necessary. The resulting capital costs are summarized in Table IV-F-4-1. Capital costs for this plant configuration are directly proportional to operating pressure with the 300 psia case costing \$60 MM and the 900 psia case costing \$75 MM.

Annual operating costs were investigated using procedures outlined in Section IV-F-3. The results are presented in Table IV-F-4-2 and Figure IV-F-4-1. Credit for surplus steam was much less significant in the overall operating costs since a larger portion of this steam was used to drive the feed and product compressors. Under conditions of 100 psia feed gas and 1000 psia delivered product, the optimum is very flat over a wide pressure range of 250 to 600 psia. This result is essentially identical with previous pressure optimization studies performed at a $3H_2/CO$ feed gas ratio. The conclusion is that the LPM/S system should be operated at the available feed gas pressure providing that this pressure is 300 psi or higher when the economic basis is private investor financing. For utility financing, there appears to

Table IV-F-4-1

CAPITAL COST SUMMARY

250 MMM BTU/D SNG

$2H_2/CO$ FEED GAS

Operating Pressure (psia)	300	500	900
Major Equipment, \$M			
Vessels	1,910	2,240	2,460
Heat Exchangers	1,560	2,130	2,440
Pumps	1,210	1,270	1,120
Miscellaneous	460	550	590
Subtotal	5,140	6,190	6,610
A. Installed (x 3.3)	16,960	20,410	21,810
Compressors, \$M			
Feed From 100 psia	8,270	11,250	16,410
Product to 1000 psia	4,050	2,290	730
Subtotal	12,320	13,540	17,140
B. Installed (x 2.0)	24,640	27,080	34,280
Fixed Capital Cost (A+B)	41,600	47,490	56,090
Associated Offsites @30% of FCI	12,480	14,250	16,830
Initial Catalyst and Oil	5,540	3,270	1,860
Total	59,620	65,010	74,780

Table IV-F-4-2

ANNUAL OPERATING COSTS
2 H_2/CO FEED GAS AT 100 PSIA
PRODUCT SNG DELIVERED AT 1000 PSIA

	$M		
Operating Pressure (psia)	300	500	900
Private Investor Financing Method			
Capital Related Costs (@40%)	23,850	26,000	29,910
Catalyst & Oil Replacement	5,030	2,970	1,660
Utilities	(3,140)	(3,330)	(2,630)
Annual Operating Cost	25,740	25,640	28,940
Utility Financing Method			
Capital Related Costs (@20%)	11,930	13,000	14,960
Catalyst & Oil Replacement	5,030	2,970	1,660
Utilities	(3,140)	(3,330)	(2,630)
Annual Operating Cost	13,815	12,640	13,990