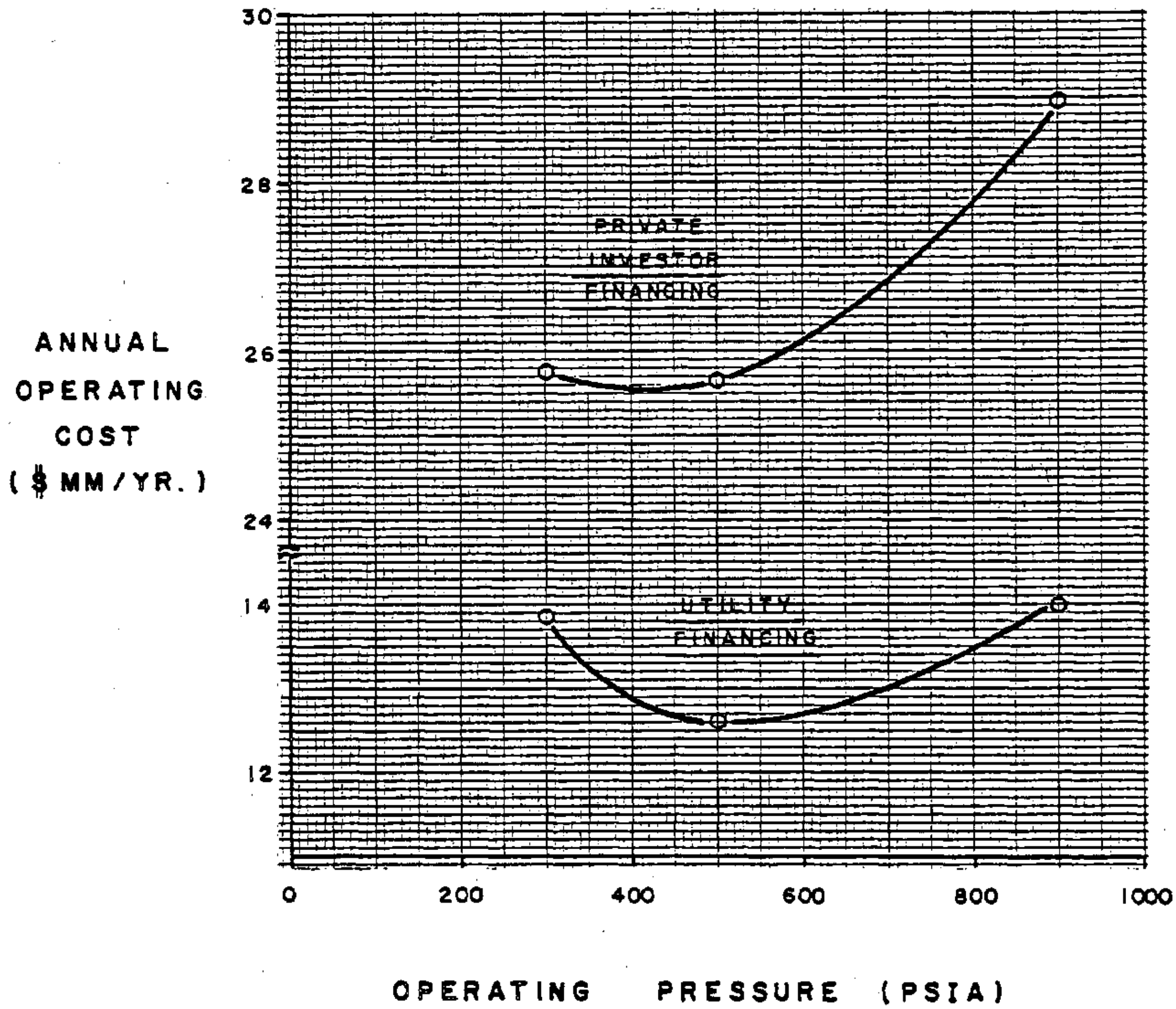
FIGURE IV-F-4-1

PRESSURE OPTIMIZATION STUDY
ANNUAL OPERATING COST
250 MMM BTU / DAY SNG

$2\ H_2/CO$ FEED GAS AT 100 PSIA
PRODUCT SNG DELIVERED AT 1000 PSIA



be a much more distinct optimum, since the effect of capital related charges is diminished. A pressure of 500 psia is the optimum point, but the differences are rather small on both sides of the optimum.

5. LPM/S Process Flowsheet With $1H_2/CO$ Feed Gas

a. General Description

Figure IV-F-5a-1 is a flowsheet for the Liquid Phase Methanation/Shift process when the starting material is a $1H_2/CO$ synthesis gas. The feed composition, shown below, most closely resembles that produced by the Koppers-Totzek process after acid gas removal although an

Component	Volume Percent
CO	49.8
H_2	49.8
CO_2	0.1
CH_4	0.3
	100.0

actual K-T synthesis gas has a H_2/CO ratio considerably less than 1/1. Other coal gasifiers which provide synthesis gases with H_2/CO ratios close to one, such as Bi-Gas and Hygas steam-oxygen, also produce considerable quantities of methane. The case considered in this section involves a feed gas which contains a low initial concentration of methane and, thus, a large investment is required in methanation reactors and heat recovery equipment in order to produce pipeline quality SNG.

It is possible to feed a $1H_2/CO$ gas directly to the LPM/S reactors, along with a substantial quantity of steam, and achieve the required degree of methanation. However, it has been found that the resulting

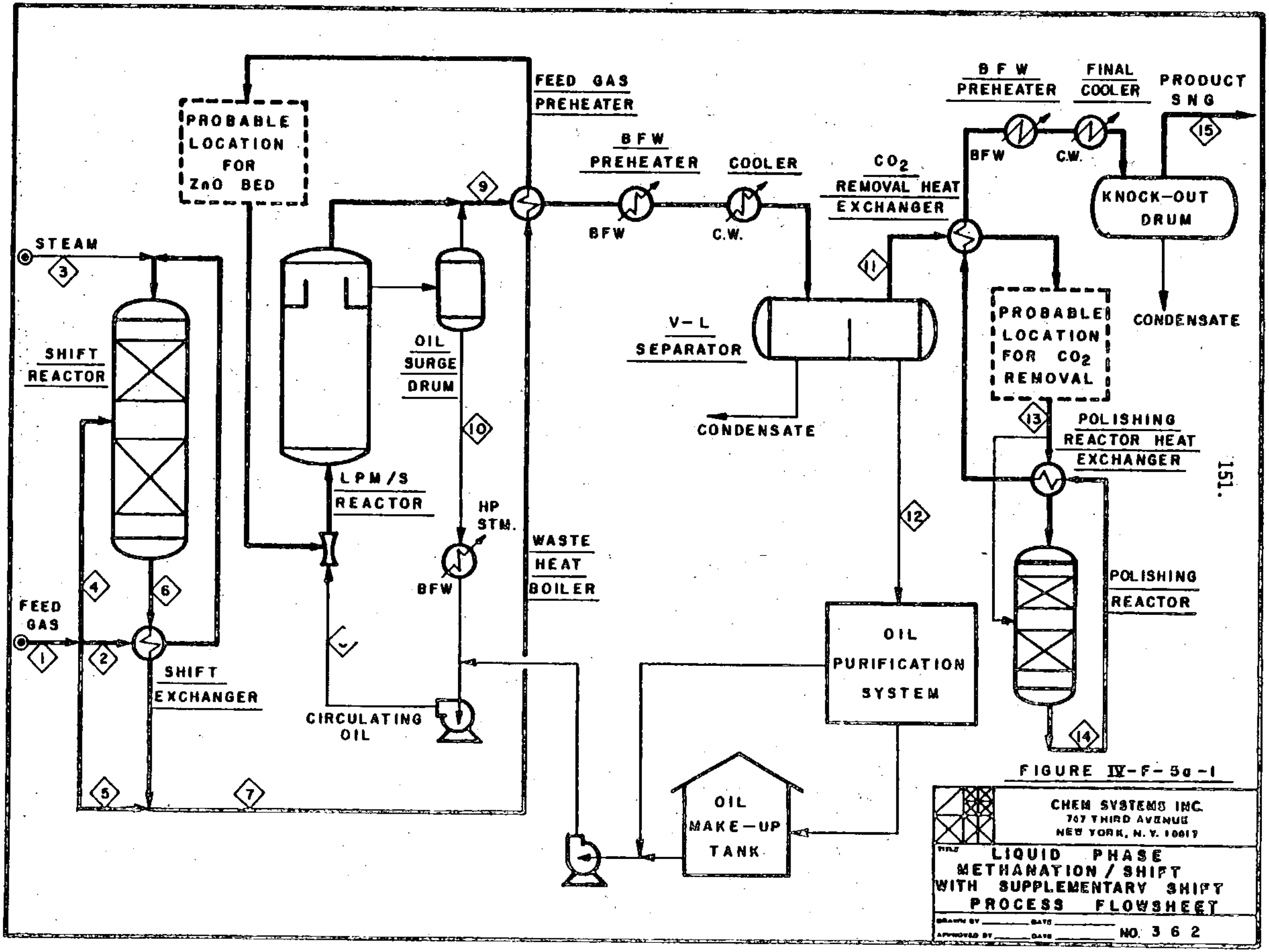


FIGURE IV-F-5a-1

low reaction rate, as well as the large steam consumption, make such a scheme unattractive. Instead, it is preferable to first shift the incoming gas to a $1.5H_2/CO$ ratio and feed this mixture to the LPM/S reactor. Optimal steam injection rate with a $1.5H_2/CO$ gas is 0.2-0.3 moles per mole of carbon monoxide.

Twenty-five percent of the synthesis gas feed, available at 120^oF and 500 psia, is passed through a shift converter along with two moles of steam per mole of CO. The steam is part of that generated from the exothermic methanation/shift reactions. The shifted gas, containing five percent CO, is recombined with the remaining by-passed synthesis gas, producing a LPM/S feed with a $1.5H_2/CO/0.375H_2O$ mole ratio at 420^oF. This is slightly more steam than is necessary for the combined methanation/shift reaction, but it will have only a small effect on reactor sizing. If necessary, the shifted gas could be pre-cooled to condense some of the excess steam.

The recombined synthesis gas is preheated to $575-580^oF$ by exchange with reactor product gas. The remainder of the process is similar to that described in Section IV-F-3 with the exception that the polishing reactor feed is split. Half is fed at 550^oF and half is used as a quench at 250^oF. Some additional heat is recovered from the final product stream in a boiler feed water exchanger.

b. Material Balance

The material balance for the $1H_2/CO$ case is presented in Table IV-F-5b-1 for a commercial plant producing 250 MMM BTU/Day of SNG. A CO conversion of 95 percent has been assumed in the LPM/S reactors and with simultaneous water gas shift, the product has a $2.5H_2/CO$ mole ratio. This gas, after removing 99 percent of the CO_2 formed by the shift reaction, is sent to the polishing reactor where all the remaining CO and part of the H_2 and CO_2 are converted to methane. The product gas contains approximately 2 percent H_2 and 2 percent CO_2.

Table IV-F-5b-1

MATERIAL BALANCE - LPM/S

1 H_2/CO FEED GAS

	Feed Gas	Gas to Shift	Steam to Shift	Quench Gas to Shift	Shift Bypass	Shift Product	Feed to LPM/S	Circulating Oil
	1	2	3	4	5	6	7	8
Moles/HR								
CO	54,480	9,080	-	4,540	40,860	2,724	43,584	-
H_2	54,480	9,080	-	4,540	40,860	24,516	65,376	-
CO_2	109	18	-	9	82	10,923	11,005	-
CH_4	329	55	-	27	247	82	329	-
H_2O	-	-	27,240	-	-	16,344	16,344	-
Oil	-	-	-	-	-	-	-	150,000
Total M/Hr	109,398	18,233	27,240	9,116	82,049	54,589	136,638	150,000
MM SCFD	995.1	165.8	247.8	82.9	746.3	496.5	1,242.9	
Flow, GPM								139,500
T (^oF)	120	120	650	120	120	969	420	579
P (psia)	500	500	500	500	500	480	475	475

Table IV-F-5b-1 (con't.)

	LPM/S Reactor Product	Oil From Reactor	Gas to CO_2 Removal	Condensed Oil Return	Gas to Polishing Reactor	Polishing Reactor Product	Cooled Product
	9	10	11	12	13	14	15
Mo. /HR							
CO	2,179.2	-	2,179.2	-	2,172.7	-	-
H_2	5,448.0	-	5,448.0	-	5,434.4	819.9	819.9
CO_2	27,076.6	-	27,076.6	-	270.0	745.9	745.9
CH_4	25,662.2	-	25,662.2	-	25,328.6	27,025.4	27,025.4
H_2O	25,605.6	-	290.0	-	2,371.8	3,592.7	117.0
Oil	5,585.0	144,415.0	3.4	5,581.6	-	-	-
Total M/HR	91,556.6	144,415.0	60,658.8	5,581.6	35,577.5	31,183.9	28,708.2
MM SCFD	832.8		551.8		323.6	283.6	261.1
Flow, GPM		134,305		5,191			
T (^oF)	650	650	120	120	250	839	120
P (psia)	450	450	445	445	425	415	410

In the material balance, Table IV-F-5b-1, solubilities of gaseous components in the hydrocarbon liquid and hydrocarbon liquid decomposition by-products themselves have been ignored. Heat recovery from the exothermic methanation/shift reaction produces 1415 ST/Hr of 900 psia, saturated steam. Of this, 245 ST/Hr are superheated to 800^{o}F, let down through turbines to drive the process pumps and sent to the shift converter at 500 psia. The remaining 1170 ST/Hr of steam are exported; i.e., available for upstream use in gasification or for compressor drives.

c. Equipment Sizing and Cost Estimation

The equipment depicted in Figure IV-F-5a-1 was sized and costed for inclusion in the plant investment with the exception of the ZnO guard system and the CO_2 removal system. A $2MM allowance under "Miscellaneous" covers the cost of oil purification, oil makeup and storage and other minor equipment not shown in Figure IV-F-5a-1. A detailed list of major equipment and costs is shown in Table IV-F-5c-1. Installed costs were based upon first quarter 1976 prices. An installed/purchased equipment ratio of 3.3 was used on all items except the circulating oil coolers where a ratio of 3.0 was taken since a large number (26) of identical heat exchangers was required.

The LPM/S reactors were sized on the basis of the rate expression given in Section IV-F-3. A gas space velocity, including steam, of 1280 Hr^{-1} results from the rate expression when 95 percent CO conversion is required starting with a 1.5 H_2/CO feed gas containing essentially no initial methane. Thus, at 475 psia, with an exit temperature of 650^{o}F, 33,800 Ft^3 of catalyst are required. Allowing a 94 percent bed expansion using Witco 40 mineral oil as the liquid medium, six reactors each 18.5' ID by 47' T-T, are required which includes disengagement area. The oil surge drums are sized to contain the total reactor liquid inventory if necessary.

Table IV-F-5c-1

MAJOR EQUIPMENT AND CAPITAL COSTS
250 MMM BTU SNG AT 500 PSIA
1 H_2/CO FEED GAS

Items	Unit Description	Number of Units	Installed Cost
Shift Reactors	10' ID x 17' T-T Refractory Lined Shell Thickness = 2.0" Material = 1%Cr - $\frac{1}{2}$% Mo	4	$ 1,510,000
LPM/S Reactors	18.5' ID x 47' T-T Shell Thickness = 3.6" Material = 1%Cr - $\frac{1}{2}$% Mo.	6	$11,690,000
Oil Surge Drums	12' ID x 32' T-T Thickness = 2.2" Material = 1%Cr - $\frac{1}{2}$% Mo.	4	$ 2,100,000
V-L Separator	10' ID x 32' T-T Thickness = 1.8" Material = CS	2	$ 630,000
Polishing Reactor	8' ID x 17' T-T Thickness = 1.8" Material = 1%Cr - $\frac{1}{2}$% Mo.	4	$ 630,000
Knockout Drum	6' ID x 12' T-T Thickness = 1.2" Material = CS	2	$ 100,000

Table IV-F-5c-1(con't.)

Item	Unit Description	Number of Units	Installed Cost
Shift Exchanger	900 Ft^2 Q = 68.2 MM BTU/Hr Material = 304 SS Tubes	2	\$ 530,000
Circulating Oil Coolers	9,400 Ft^2 Q = 2,176 MM BTU/Hr (Tot.) Material = 1%Cr - $\frac{1}{2}$% Mo	26	\$ 8,730,000
Feed Gas Preheaters	8,500 Ft^2 Q = 164 MM BTU/Hr (Tot.) Material - 1%Cr - $\frac{1}{2}$% Mo	2	540,000
BFW Preheaters	9,700 Ft^2 Q = 950 MM BTU/Hr (Tot.) Material = 1%Cr - $\frac{1}{2}$% Mo	6	\$ 2,280,000
Coolers	10,000 Ft^2 Q = 156 MM BTU/Hr (Tot.) Material = 1%Cr - $\frac{1}{2}$% Mo.	6	\$ 1,860,000
CO_2 Removal Heat Exchanger	1,100 Ft^2 Q = 72 MM BTU/Hr Material = 1%Cr - $\frac{1}{2}$% Mo.	2	\$ 190,000
Polishing Reactor Heat Exchanger	900 Ft^2 Q = 53 MM BTU/Hr Material - 1%Cr - $\frac{1}{2}$ % Mo	2	\$ 170,000
BFW Preheater	4,500 Ft^2 Q = 92 MM BTU/Hr Material = 1% Cr - $\frac{1}{2}$% Mo	2	\$ 460,000

Table IV-F-5c-1(con't.)

Item	Unit Description	Number of Units	Installed Cost
Final Cooler	6,700 Ft^2 Q = 91 MM BTU/Hr Material = 1%Cr - $\frac{1}{2}$% Mo	2	$ 580,000
Circulating Oil Pumps	17,400 GPM x 30 psi 435 Horsepower Material = 1%Cr - $\frac{1}{2}$% Mo	8 + Spare	$ 3,170,000
Condensed Oil Return Pumps	2,600 GPM x 100 psi 220 Horsepower Material = 1%Cr - $\frac{1}{2}$% Mo	2 + Spare	$ 500,000
BFW Pumps	710 GPM x 950 psi 560 Horsepower Material = CS	8 + Spare	$ 2,280,000
Miscellaneous	Oil Makeup, Purification, Etc.		$ 2,000,000
	Total BLCC		$39,950,000

The shift reactors contain 5070 Ft^3 of catalyst split into upper and lower sections with intermediate cold gas injection. Because of the high exit temperature (970^oF), this vessel is refractory lined.

The polishing reactors are sized by assuming a gas hourly space velocity of 5000 HR^{-1} at 420 psia thus yielding a catalyst requirement of 2700 Ft^3. Four reactors, 8' ID by 17' T-T, were chosen for this purpose.

The total BLCC for this plant was $39,950,000 which is nearly twice the cost of the $2H_2/CO$ case outlined in Section IV-F-3. However, it should be pointed out that the $1H_2/CO$ case requires twice as much methanation since the initial methane concentration is negligible. Also, more than twice as much net exportable steam is produced. Thus, this plant is actually double sized in terms of H_2-CO conversion capacity. The initial capital cost includes shift catalyst at $50/$Ft^3$, methanation catalyst at $200/$Ft^3$ and mineral oil at $1/Gal for a total of $7,650,000.

d. <u>Annual Costs</u>

Utility requirements are summarized in Table IV-F-5d-1 where all internal power is supplied by passing 850 psig superheated steam through turbines and exhausting at 500 psig. This steam is then fed to the shift reactor.

Annual operating costs are calculated on the same basis as described in Section IV-F-3. Shift catalyst life was assumed to be two years. The resultant operating costs for a 250 MMM BTU/Day SNG plant, starting with a $1H_2/CO$ synthesis gas at 500 psia, are summarized in Table IV-F-5d-2. Assuming no feed or product compression and ignoring labor costs, the LPM/S plant operates at a net credit of $2.9MM/Yr based upon a private investor financing method or a net credit of $14.8 MM/Yr based upon a utility financing method. This is better than the operating

Table IV-F-5d-1

UTILITY REQUIREMENTS
250 MMM BTU/D SNG
$1H_2/CO$ FEED GAS AT 500 PSI

			$/Yr
Fuel			
Superheating saturated steam to 800°F for internal steam drives:			
96.6 MM BTU/Hr @ $2.25/MM BTU			1,739,000
Boiler Feedwater	339,090 Gal/Hr	@ $.55/M Gal	1,492,000
Cooling Water	1,533,000 Gal/Hr	@ $.03/M Gal	368,000
Steam			
Generated: 900 psia, sat'd:	2,830,000 Lbs/Hr		
Steam to internal drives and, then, to shift:	478,000 Lbs/Hr[a]		
Additional steam to shift:	12,300 Lbs/Hr		
Net	2,339,700 Lbs/Hr		
	@ $2.00/M Lbs		(37,435,000)
Total Utility Cost (Credit)			(33,836,000)

(a) Total Horsepower Required: 8,310 HP
Power: In terms of internal steam drives using
850 psig, 800°F steam to 500 psig: 478,000 Lbs/Hr

Table IV-F-5d-2

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

	$	
BLCC	39,950,000	
Associated Offsites @30% of BLCC	11,990,000	
Initial Catalyst and Oil	7,650,000	
Total Fixed Capital Investment	59,590,000	
Private Financing Equivalent	**\$/Yr**	**¢/MM BTU**
Capital Related Costs (@40%)	23,840,000	
Catalyst and Oil Replacement	7,100,000	
Utilities	(33,840,000)	
Annual Operating Cost	(2,900,000)	(3.5)
Utility Financing Equivalent		
Capital Related Costs (@20%)	11,920,000	
Catalyst and Oil Replacement	7,100,000	
Utilities	(33,840,000)	
Annual Operating Cost	(14,820,000)	(18.0)

costs calculated for the $2H_2/CO$ case in Section Iv-F-3 and illustrates that the process has a high efficiency despite the large investment in reactors and heat recovery equipment. Of course, the annual operating costs are kept low only if the credit for by-product steam can be assumed reasonable such as the $2/M Lbs taken in this study.

6. Comparison Of LPM/S With Conventional Routes

In Sections IV-F-3 and 5, the flowsheets and economics of the combined Liquid Phase Methanation/Shift process have been developed for two different synthesis gas feedstocks. It remains, then, to compare these systems against combinations of conventional methanation and shift processing in order to determine the relative advantages of LPM/S over the alternatives. Consequently, each of the feedstocks has been converted via conventional water-gas shift to a $3H_2/CO$ ratio and then processed in a standard LPM system. In addition, each feed gas was studied via conventional shift followed by the Lurgi Hot Gas Recycle route as an example of conventional fixed bed methanation processing. The discussion and economic comparison of these alternatives follow.

Three cases are developed for the $2H_2/CO$ synthesis gas containing 19 percent methane:

Case 1: Liquid Phase Methanation/Shift

Case 2: Conventional Shift to $3H_2/CO$ followed by CO_2 removal and then Liquid Phase Methanation.

Case 3: Conventional Shift to $3H_2/CO$ followed by CO_2 removal and then Lurgi Hot Gas Recycle Methanation.

In each case the feed gas is available at 500 psia and the product is compressed to 1000 psia. All power requirements are handled internally by steam turbines and the use of superheated steam from methanation.

This steam is also used as a feed to the shift converter where required. The flowsheet and costs for Case 1 are exactly as presented in Section IV-F-3, except for the addition of a product compressor and the associated changes in utilities. In Case 2, 35 percent of the feed gas passes through a shift converter containing 2400 Ft^3 of catalyst. Steam is added at a rate of 235,000 Lbs/Hr and the shift product has a H_2/CO molar ratio of 10/1. The shifted gas is cooled to 250^oF by generating 90 psia steam and preheating incoming feed, and, then, passes through a CO_2 removal unit (not included in cost estimate). This gas is then recombined with by-passed feed and proceeds to the methanation section which is similar to that described in Section IV-F-3, except that the LPM reactors contain only 9700 Ft^3 of catalyst. The LPM product goes to a polishing reactor and is then compressed to 1000 psia.

In Case 3, after passing through a shift section identical to that of Case 2, the feed gas enters the Lurgi Hot Gas Recycle system which has been described in detail previously.(2) In brief, the feed gases are preheated to 550^oF and sent to eight parallel fixed bed reactors containing 27,000 Ft^3 of catalyst. The product is cooled by generating 600 psia steam. Part of the effluent gas is recycled while still over 500^oF and the remainder proceeds to finishing reactors, liquid separation facilities, and compression to 1000 psia.

The economics of these three cases are compared in Table IV-F-6-1. From the viewpoint of capital costs, Case 1 is the least expensive ($35.9 MM) while Case 3 is, by far, the most expensive ($56.7 MM). In comparing annual operating costs based upon private investor financing, capital related items were taken as 40 percent of the total fixed capital investment. Catalyst and oil replacement were calculated as described in Sections IV-F-3 and 5, except that the Lurgi HGR catalyst was assumed to have a five-year life. Utilities were also calculated on the same basis as previously described with the addition of 600 psia,

(2) Liquid Phase Methanation, Interim Report No. 2, ERDA, June 30, 1974, Section VI-E, P. 102

Table IV-F-6-1

COMPARISON OF ALTERNATIVES
FOR $2H_2/CO$ FEED GAS WITH
PRODUCT DELIVERED AT 1000 PSIA

	Case 1	Case 2	Case 3
	LPM/S Only	Shift to $3H_2/CO$ Plus LPM	Shift to $3H_2/CO$ Plus Lurgi HGR
Capital Cost, $M			
BLCC	25,070	27,550	39,300
Associated Offsites @ 30% of BLCC	7,520	8,270	11,790
Init. Catalyst & Oil	3,270	2,540	5,590
Total FCI	35,860	38,360	56,680
Power & Steam			
Horsepower Requirements	20,500	21,000	35,700
Net 900 Psia Steam, Lbs/Hr	978,000	797,000	-
Net 600 Psia Steam, Lbs/Hr	-	-	750,000
Net 90 Psia Steam, Lbs/Hr	-	94,000	94,000
Annual Costs (Private Financing), $M/Yr			
Capital Related (@40%)	14,340	15,340	22,670
Cat. & Oil Replacement	2,970	2,170	1,160
Utilities	(14,240)	(11,320)	(9,090)
Annual Operating Cost	3,070	6,190	14,740
Cost in ¢/MM BTU	3.7	7.4	17.7
Annual Costs (Utility Financing), $M/Yr			
Capital Related (@20%)	7,170	7,670	11,340
Cat. & Oil Replacement	2,970	2,170	1,160
Utilities	(14,240)	(11,320)	(9,090)
Annual Operating Cost	(4,100)	(1,480)	3,410
Cost in ¢/MM BTU	(4.9)	(1.8)	4.1

saturated steam at \$1.80/M Lbs and 90 psia, saturated steam at \$1.20/M Lbs. Overall, the annual operating cost is \$3.1 for Case 1, \$6.2 MM for Case 2 and \$14.7 MM for Case 3. Using Case 3 as a basis, it can be seen that substitution of LPM for Lurgi HGR results in a savings of 10.3 ¢/MM BTU SNG. Substitution of LPM/S for the shift and LPM results in an additional savings of 3.7 ¢/MM BTU or an overall savings of 14.0 ¢/MM BTU as compared to the Lurgi Hot Gas Recycle case. A comparison of annual operating costs, based upon utility financing where capital related items were taken as 20 percent of the total fixed capital investment, shows a net credit of \$4.1 MM/Yr for Case 1 and \$1.5 MM/Yr for Case 2 while Case 3 has a cost of \$3.4 MM/Yr. Thus, for utility financing, LPM/S results in a savings of 3.1 ¢/MM BTU over shift plus LPM or an overall savings of 9.0 ¢/MM BTU over the Lurgi HGR case.

Three cases were developed for the $1H_2/CO$ synthesis gas without methane:

Case 4: Conventional Shift to $1.5H_2/CO$ followed by Liquid Phase Methanation/Shift.

Case 5: Conventional Shift to $3H_2/CO$ followed by CO_2 removal and then Liquid Phase Methanation.

Case 6: Conventional Shift to $3H_2/CO$ followed by CO_2 removal and then Lurgi Hot Gas Recycle Methanation.

The basic assumptions are identical with those described in Section IV-F-5 and the only modification in Case 4 from the previous flowsheet and costs is the addition of product compression to 1000 psia. In Case 5, 63 percent of the feed gas passes through a shift converter containing 12,700 Ft^3 of catalyst. Steam is added at the rate of 1,230,000 Lbs/Hr and the shifted gas is cooled to 250^oF by preheating incoming feed and generating 90 psia steam. The cooled gas passes through a CO_2 removal unit, which is not included in the cost estimate, and is then recombined with by-passed gas. After preheating, this gas proceeds to the LPM reactors which contain 10,300 Ft^3 of catalyst. In most other respects, the methanation section is similar to that described in Section IV-F-5. The LPM product goes to a polishing reactor, is cooled and is then compressed to 1000 psia.

Case 6 consists of a shift section, similar to that described in Case 5, and a Lurgi Hot Gas Recycle methanation unit in place of the LPM system. In order to contain the first stage methanation reactors within the required 550-900°F temperature limits, a recycle to feed ratio of 5.5/1 is required. Thus, recycle compressors totaling 42,000 horsepower must be specified and the first stage reactors must be sized for 42,700 Ft^3 of catalyst, assuming a space velocity of 6000 Hr^{-1}. The product from the first stage reactors proceeds to polishing reactors containing 2000 Ft^3 of catalyst, is then cooled and compressed to 1000 psia.

The economics of these three cases are compared in Table IV-F-6-2. In contrast to the cases with a $2H_2/CO$ synthesis gas feed, shifting to $3H_2/CO$ followed by LPM (Case 5) requires less capital investment, \$56.2 MM, as compared to shifting to 1.5 H_2/CO followed by LPM/S, \$66.1 MM. This is due to the large reactors and catalyst requirements in Case 4. The Lurgi HGR case (Case 6) is considerably more expensive than either of the LPM cases, having a capital investment of \$82.7 MM. In terms of annual costs, based upon private investor financing, the combined shift/methanation case is the lowest with \$2.2 MM/Yr as compared to \$5.6 MM/Yr for separate shift followed by Liquid Phase Methanation and \$22.2 MM/Yr for the Lurgi methanation case. LPM thus results in a savings of 19.9 ¢/MM BTU over Lurgi HGR and LPM/S increases the savings by 4.1 ¢/MM BTU to 24 ¢/MM BTU overall. Based upon utility financing, the annual operating credit for LPM/S is \$11.0 MM/Yr and \$5.7 MM/Yr for LPM while the operating cost of Lurgi HGR is \$5.6 MM/Yr. The resultant savings of LPM over Lurgi is 13.5 ¢/MM BTU and LPM/S adds an additional savings of 6.4 ¢/MM BTU for an overall savings of 19.9 ¢/MM BTU. The savings result from the difference in capital investment

Table IV-F-6-2

COMPARISON OF ALTERNATIVES FOR
$1H_2/CO$ FEED GAS WITH PRODUCT
DELIVERED AT 1000 PSIA

	Case 4	Case 5	Case 6
	Shift to 1.5 H_2/CO Plus LPM/S	Shift to 3 H_2/CO Plus LPM	Shift to 3 H_2/CO Plus Lurgi HGR
Capital Costs, $M			
BLCC	44,930	40,650	56,270
Offsites (30% of BLCC)	13,480	12,200	16,880
Init. Catalyst & Oil	7,650	3,330	9,580
Total FCI	66,060	56,180	82,730
Power & Steam			
Horsepower Requirements	24,900	24,400	64,000
Net 900 Psia Steam, Lbs/Hr	2,210,000	1,140,000	-
Net 600 Psia Steam, Lbs/Hr	-	-	890,000
Net 90 Psia Steam, Lbs/Hr	-	820,000	820,000
Annual Costs (Private Financing), $M/Yr			
Capital Related (@40%)	26,420	22,470	33,090
Cat. & Oil Replacement	7,100	2,580	2,110
Utilities	(31,350)	(19,470)	(13,040)
Annual Operating Cost	2,170	5,580	22,160
Cost in ¢/MM BTU	2.6	6.7	26.6
Annual Costs (Utility Financing), $M/Yr			
Capital Related (@20%)	13,210	11,240	16,550
Cat. & Oil Replacement	7,100	2,580	2,110
Utilities	(31,350)	(19,470)	(13,040)
Annual Operating Cost	(11,040)	(5,650)	5,620
Cost in ¢/MM BTU	(13.2)	(6.8)	6.7

and from the fact that the 600 psia steam generated in the Lurgi case yields no useful power before being utilized in the shift converters, whereas most of the power required in the LPM case is obtained from the 900 psia steam that is eventually fed to the shift converters at 500 psia.

In conclusion, it has been demonstrated that the Liquid Phase Methanation/Shift process is economically attractive for 250 MM BTU/Day SNG coal conversion complexes. The LPM/S process can be tailored to fit a wide variety of synthesis gases and system conditions where the feed H_2/CO ratio is less than 3. In all cases, LPM/S operates at lower annual cost than separate shift and methanation, however, in cases of very low H_2/CO ratios, the capital investment may be higher than conventional shift followed by Liquid Phase Methanation. Further studies are probably required to pinpoint the optimal combination of shift followed by LPM/S for such synthesis gases.

V. APPENDIX

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