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A SASOL TYPE PROCESS FOR GASOLINE, METHANOL, SNG, AND LOW-BTU GAS FROM COAL

by

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TASK NO. 13 FINAL REPORT

Submitted To
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I. Introduction

The work reported herein is an enlargement of Task 13 Preliminary Report (1) and was performed for the Environmental Protection Agency, Office of Research and Development under Task 13, Change No. 2, Contract No. 68-02-1308.

Task 13 Preliminary Report describes the SASOL type process for the production of gasoline-from-coal and presents the approximate capital investment. The evaluation was based on published data on the SASOL plant (2) and costs include the power supply to the nearby village which houses the operating and supervisory personnel. To provide a basis for direct comparision with other coal conversion processes, it was decided to extend the study by modifying the basis such that power requirements for the village were deleted. The modified gasoline-from-coal plant is described in the appropriate sections in this report.

In addition to the modification cited above, the Environmental Protection Agency (EPA) requested that Task 13 be enlarged to include the following:

- Block flow diagrams and cost estimates for the production of methanol, substitute natural gas and low Btu gas via the SASOL-type process.

- Annualized Operating and production costs of the above processes

II. Basis of Evaluation

Basis of the estimates is a mine-mouth plant, Western U.S. coal, with capital investment expressed in 1975 dollars. Flow sheets and cost estimates were to be derived from existing information and no detailed estimates were to be prepared.

The only fuel used in the plant is coal. Part of this coal is used for steam and power generation, both of which are needed to operate the plant and off-site facilities, and the remainder is used for the production of synthesis gas and ultimately from it the desired products.

The Western U.S. coal selected for this study is from a New Mexico source and has properties used by El Paso Natural Gas Company in their feasibility study of Burnham Coal Gasification Complex for the manufacture of substitute natural gas (3). It is fully expected that this coal will gasify well in either air- or oxygen-blown Lurgi gasifiers.

Another assumption made at the outset and incorporated in the block flow diagram was that only limited sulfur and hydrocarbon emissions to the atmosphere would be permitted. Also clean-up of liquid effluents would be required.

The quantity of water required by the plant was not estimated since appropriate site information needed were not available. Items such as wet and dry bulb temperatures, available water quantity and quality, annual rain fall, percent run-off, soil conditions, etc., must be known before estimates of water requirements can be made.

III. Summary

A. Gasoline-From-Coal

Deletion of the power supply to the nearby village reduces the coal feed rate to the fuel gas manufacture (Section 1000) by about 21%. Total coal feed to the plant is revised to 34,249 TPD. The block flow diagram for the gasoline-from-coal plant (MWK Dwg. P3925-D) has been revised accordingly. No excess power production is included and the plant is self-contained in that the only input to the plant is coal, air and water. The revised capital investment is estimated to be $505 million, 1975 dollars. Table 1 presents a complete breakdown of the capital investment by individual sections. These figures, represent the capital investment for a plant which produces 44,500 BPSD (Nominal) gasoline and other hydrocarbon liquids from coal, have been derived from published information (2, 3) and updated to 1975 by means of escalation rates for both labor and material. It should be noted that overnight construction of the plant is assumed. That is, it is assumed that the complete plant can be designed and constructed in such a short period of time that no additional increase in labor or material will be encountered. In short, no forward escalation has been included in the estimated cost of $505 million. In actual practice significant increases in costs probably would be encountered since several years would be required to build the plant.

Production rates of the various products are given in Table 2. Using a modified Panhandle Eastern accounting procedure which is recommended for coal conversion facilities (4), the cost of gasoline for a 20-year average price (excluding escalation) is estimated to be $3.05/MMBtu or $15/barrel assuming all appropriate byproducts are marketable. The price of coal which corresponds to these gasoline costs is $3.60/ton (7). The effect of other coal prices on the gasoline cost is given in Fig. 5. Using a different accounting procedure with a fixed capital charge of 18.22% (6), the gasoline cost for a 15-year average price is estimated to be \$4.05/MMBtu or \$20/barrel (with by-product credits).

Approximately 80% of the tar, oil and naphtha produced can be further processed into gasoline product. Assuming addition of such a conversion facility does not alter the total plant investment significantly, gasoline cost can be reduced to \$2.76/MMBtu or \$13.70/barrel using the Panhandle Eastern accounting method. Such a process modification increases the gasoline production rate by about 40%.

It should be pointed out that should the power generation plant for the nearby village (housing the operation and supervisory personnel) be included as an integrated part of the gasoline-from-coal facility, the total coal consumption will be increased to 37,665 TPD (1). Approximately 40% of the coal is used for steam and power generation. The revised total plant investment for the gasoline-from-coal facility including the power plant and other offsites is estimated to be \$533 million, 1975 dollars.

B. Methanol-From-Coal

A block flow diagram has been developed and approximate overall material and energy balances calculated for the methanol-from-coal plant. Using the same coal feed rate to the gasification section as used for the gasoline-from-coal plant, the methanol production rate is 11,338 TPD (Table 4). A maximum size train is estimated to produce 2,800 TPD of methanol and the present plant, therefore, consists of four such units in parallel.

Some seventeen sections are required for the methanol-from-coal plant. Most of these sections are similar in design as well as operation to those used in the gasoline-from-coal plant and the costs for these sections are derived from figures given in the gasoline plant. Sections which are different from those in the gasoline plant are the shift conversion, methane reforming, synthesis gas compression and methanol synthesis and recovery. Costs of these sections are derived from in-house information for a smaller size methanol plant. Capital investment for the seventeen sections is estimated to be $472 million (Table 3), 1975 dollars assuming overnight construction of the plant with no forward escalation. Based on this estimate and using the modified Panhandle Eastern accounting procedure, the cost of producing methanol from coal is estimated to be $1.80/MMBtu ($4.90/barrel) assuming all appropriate byproducts are marketable and a coal cost of $3.60/ton. For a 15-year plant life with a fixed capital charge of 18.22%, the methanol cost is estimated to be $2.34/MMBtu or $6.40/barrel. Appendix B presents the operating and annualized production cost for the methanol-from-coal plant.

C. Substitute Natural Gas-From-Coal

A block flow diagram for the substitute natural gas (SNG) from-coal plant have been developed and overall material and energy balances calculated. The SNG-from-coal plant has the same coal feed (21,274 TPD) to the gasification unit (Section 200) as the two previous coal conversion processes. In order for the plant to be self-contained, an additional coal feed rate of 5,435 TPD is required as input to the fuel gas manufacture section for steam and power generation. The corresponding gas production rate is 258 MMSCFD (972 Btu/SCF).

Eighteen sections are required for the SNG-from-coal plants. Most of these sections are similar in design to the two previous coal conversion processes with the exception of the methane synthesis and synthesis gas compression sections. The cost for these two sections are derived from published information (3).

Capital investment for a SNG-from-coal plant producing 258 MMSCFD of gas is estimated to be $365 million, 1975 dollars assuming overnight construction of the plant with no forward escalation. Table 5 presents a detail breakdown of the capital investment for individual sections. Based on this estimate and assuming all by-products are marketable, the gas prices is estimated to be $1.13/MMBtu using the modified Panhandle Eastern accounting procedure and a coal price of $3.60/ton. The SNG cost resulting from an alternative accounting procedure with a 15-year plant life will be $1.50/MMBtu (with by-product credits). Production rates of the various products are listed in Table 6. Appendix C outlines the annualized production and operating costs of the SNG-from-coal plant.

D. Low Btu Gas-From-Coal

A block flow diagram for a comparable low Btu gas-from-coal plant has been developed and overall material and energy balances calculated. The low Btu gas facility has a coal feed rate of 21,274 TPD to the gas manufacture section which is the same rate used in the other three coal conversion processes. In order for the plant to be self-contained, part of the fuel gas manufactured is used in the steam and power generation plant for process consumption. The remainder of the fuel gas (960 MMSCFD @ 230 Btu/SCF) is transmitted as product.

The capital investment for the low Btu gas-from-coal facility is estimated to be $218 million, 1975 dollars and is derived from the fuel gas manufacturing and other corresponding sections of the gasoline-from-coal plant. Table 7 presents a complete breakdown of the capital investment for individual sections of the low Btu-gas-from-coal facility. Based on this estimate and assuming all by-products are marketable, the low Btu gas cost is $0.86/MMBtu using the modified Panhandle Eastern accounting procedure and a coal price of $3.60/ton.

The low Btu gas resulting from the alternative accounting procedure with a 15-year plant life is estimated to be $1.10/ MMBtu (with by-product credits). These costs of low Btu gas are derived using an on-stream factor of 0.9. The production costs for 0.7 on-stream factor have also been investigated as typical for hook-up with power plants. The low Btu gas costs for the lower on-stream factor are $1.10/MMBtu using the modified Panhandle accounting procedure and $1.44/MMBtu using the alternative (15-year plant life) accounting procedure and with by-product credits. Figure 5 illustrates graphically the sensitivity of on-stream factors on the low Btu gas cost. Production rates of the various products from the low Btu gas-from-coal plant are listed in Table 8. Annualized production and operating costs of the low Btu gas-from-coal plant are given in Appendix D for two cases: 0.9 and 0.7 on-stream factors.

Table 1

Gasoline-From-Coal Investment Summary

	Section	1975, M$
100	Coal Preparation	38,000
200	Coal Gasification	48,000
300	Gas Purification	41,000
400	Methane Splitting	10,000
500	Synthesis	53,000
600	Product Recovery	28,000
700	Chemical Recovery	8,000
800	Hydrogen & Catalyst Manufacture	8,000
900	Oxygen Production	73,000
1000	Fuel Gas Production	62,000
1100	Steam & Power	42,000
1200	Gas Liquor Treating	13,000
1300	Ash Disposal	8,000
1400	Effluent Water Treating	4,000
1500	Sulfur Recovery	9,000
1600	Raw Water Treating	6,000
1700	Cooling Water	15,000
1800	Offsite & General	39,000
	TOTAL	$505,000

Table 2

Gasoline-From-Coal
Production Rate

Product	#/hr	BPD	HHV, 10^9 Btu/hr
Gasoline**	262,353	25,495	5.2654
Diesel Oil	15,121	1,233	0.2913
Waxy Oil	11,847	925	0.2175
Propane LPG	16,056	2,000	0.3412
Acetone	2,650	230	0.0348
Methanol	343	30	0.0033
Propanol	4,832	412	0.0698
i-Butanol	546	46	0.0085
n-Butanol	1,606	136	0.0249
M.E.K.	670	56	0.0097
n-Pentanol	374	32	0.0056
Tar, Oil, Naphtha***	162,074	13,230	3.8804
Phenol	13,665	716	0.0218
Ammonia	28,742	--	0.2763
Sulfur	13,757	--	0.0548
		TOTAL	10.5053

$$\frac{\text{HHV of Products}}{\text{HHV of Coal}} = \frac{10.5053}{25.3209} \times 100\%$$

$$= 41.5\%$$

** Gasoline with research octane number equals 86

***80% of tar, oil, naphtha can be further processed to gasoline product

Table 3

Methanol-From-Coal Investment Summary

	Section	1975, M$
100	Coal Preparation	36,000
200	Coal Gasification	48,000
300	Gas Purification	41,000
400	Shift Conversion	9,000
500	Methane Splitting	32,000
600	Synthesis Gas Compression	19,000
700	Methanol Synthesis & Recovery	65,000
800	Sulfur Recovery	8,000
900	Oxygen Production	43,000
1000	Fuel Gas Manufacture	51,000
1100	Steam & Power Generation	36,000
1200	Gas Liquor Treating	12,000
1300	Ash Disposal	8,000
1400	Effluent Water Treatment	4,000
1500	Raw Water Treatment	6,000
1600	Cooling Water	15,000
1700	Offsites	39,000
TOTAL		472,000

Table 4

Methanol-From-Coal

Production Rate

Product	#/hr	BPD	HHV, 10^9 Btu/hr
Methanol	927,748	81,433	9.2281
Tar, Oil, Naphtha	151,684	12,382	3.6798
Higher Alcohols & Dimethyl Ether	4,452	367	0.0662
Phenols	13,052	684	0.0209
Ammonia	27,453	--	0.2639
Sulfur	11,323	--	0.0451
		TOTAL	13.3040

$$\frac{\text{HHV of Products}}{\text{HHV of Coal}} = \frac{13.3040}{23.4912} \times 100\%$$

$$= 56.6\%$$

Table 5

Substitute Natural Gas-From-Coal Investment Summary

	Section	1975, M$
100	Coal Preparation	32,000
200	Coal Gasification	48,000
300 + 400	Shift Converter & Gas Cooling	15,000
500	Gas Purification	41,000
600	Methane Synthesis	19,000
700	Gas Compression	7,000
800	Gas Liquor Separation	(Included in 300 and 400)
900	Air Separation	43,000
1000	Fuel Gas Manufacture	29,000
1100	Steam & Power Generation	48,000
1200	Phenol Recovery	10,000
1300	Ash Disposal	4,000
1400	Effluent Water Treatment	5,000
1500	Sulfur Plant	8,000
1600	Raw Water Treatment	5,000
1700	Cooling Water	15,000
1800	Offsites	36,000
	TOTAL	365,000

Table 6

Substitute Natural Gas Production Rate

Product	#/hr	Production	HHV, 10^9 Btu/hr
Substitute Natural Gas	474,069	258 MMSCFD	10.4531
Tar, Oil, Naphtha	149,002	12,160 BPD	2.7565
Phenols	10,440	550 BPD	0.0167
Ammonia	17,353	208 STPD	0.1667
Sulfur	12,698	136 LTPD	0.0507
		Total	13.4437

$$\frac{\text{HHV of Products}}{\text{HHV of Coals}} = \frac{13.4437}{19.7462} \times 100\%$$

$$= 68.1\%$$

Table 7

Low Btu Gas-From-Coal Investment Summary

	Section	1975, M$
100	Coal Preparation	25,000
200	Coal Gasification & Gas Purification (Hot Carbonate System)	98,000
300	Ash Disposal	3,000
400	Steam & Power Generation	38,000
500	Gas Liquor Treatment	8,000
600	Effluent Water Treatment	4,000
700	Raw Water Treatment	4,000
800	Sulfur Recovery	7,000
900	Cooling Water	5,000
1000	Offsites	26,000
	TOTAL	218,000

Table 8

Low Btu Gas-From-Coal
Production Rate

Product	#/hr	Production	HHV, 10^9 Btu/hr
Low Btu Gas	2,437,756	960 MMSCFD	9,1936
Tar, Oil, Naphtha	88,018	7,185 BPD	1.7247
Phenols	8,320	436 BPD	0.0133
Ammonia	13,780	165 STPD	0.0403
Sulfur	10,114	108 LTPD	0.1325
		Total	11.1044

$$\frac{\text{HHV of Products}}{\text{HHV of Coal}} = \frac{11.1044}{15.728} \times 100\%$$

$$= 70.6\%$$

IV. Process Description

A. Gasoline-From-Coal (MWK Dwg. P3925-D)

a. DESCRIPTION

A block flow diagram for the gasoline-from-coal plant is given in the cited drawing. The process may be divided into following steps:

- Coal Gasification and Raw Gas Purification
- Synthesis Gas Preparation
- Synthesis of Gasoline and Recovery
- Fuel Gas Manufacture - Steam & Power Generation

1. Coal Gasification and Raw Gas Purification

Coal received from the mine at section 100, Coal Preparation, undergoes crushing, screening, stockpiling, reclaiming and briquetting treatment according to needs of the plant. The product from section 100 is conveyed by belt to both section 200, Gasification, and section 1000, Fuel Gas Manufacture. In both areas coal is gasified essentially completely for manufacture of both synthesis gas and fuel gas.

Ash from both sections 200 and 1000 is conveyed either by belt or by water to section 1300, Ash Disposal. This section includes thickeners and screens for the recovery of water for reuse in ash quenching and/or sluicing.

Gasification in section 200 employs steam and oxygen in a mixture introduced under the grates of multiples of Lurgi Gasifiers. Operating under pressure of up to 500 psig, the Lurgi gasifier receives its coal input through a lock hopper pressured by product gas. This lock hopper is periodically filled from an overhead bunker when level indicators show that the hopper has been emptied into the gasifier. The gasifier proper is a water-jacketted pressure vessel equipped with a distributor for the coal feed at the top and a water cooled grate at the bottom which has the dual function of distributing the gasification medium (steam and oxygen or air and steam) and discharging the ash. Ash is discharged to a lock hopper which is periodically emptied when the volume of ash builds to a pre-set level.

In the gasifier, the falling fixed bed of coal supported on the grate moves from coal to ash through stages of devolatilization and combustion. The gas product is principally the result of reaction between hot gases leaving the combustion zone and the devolatilized coal. Gas leaving the relatively cool top outlet of the gasifier is quenched to knock down condensible carbonization products and unconverted steam. Tar is separated from the gas liquor by decanting at about the boiling point of water; oil and naphtha are separated from gas liquor by decanting at essentially room temperature. The gas liquor is principally water but contains both phenols and dissolved gases, e.g., H_2S, CO_2 and NH_3.

Gas released from the feed hopper, as the hopper is depressured for refilling, is collected from a battery of gasifiers and recompressed to gasifier pressure. Little gas is involved in the cycling of the ash hopper as the atmosphere under the grate is steam and oxygen, and the steam is condensed before the vessel is opened for discharge of the ashes.

Cooled raw gas is separated from its condensates and sent to section 300, Gas Purification where it undergoes scrubbing with a chilled polar solvent according co a version of the Rectisol process first employed industrially at SASOL. Stripping of the solvent is achieved both by flashing and also by chilled nitrogen from section 900, Air Separation. Off-gases comprise a hydrogen sulfide concentrate sent to section 1500, Sulfur Recovery, and a mixture of mostly nitrogen and carbon dioxide, sent to the stacks of section 1100, Steam and Power.

2. Synthesis Gas Preparation

Pure gas from section 300 is the net feed to the synthesis section but by virtue of having been generated in a pressure Lurgi gasifier, contains methane which is an unwanted component in the synthesis feed. In section 400, Methane Splitting, methane and other light hydrocarbons produced by the synthesis are split into hydrogen and carbon monoxide which, of course, are the ingredients wanted in the synthesis feed. The splitting is performed catalytically at high temperature with proper additions of steam and oxygen to produce the desired synthesis feed. A small fraction of the split gas is diverted to section 800, Hydrogen Plant, where catalytic treatments of shift and methanation produce a stream of 99% hydrogen for synthesis catalyst reduction and other uses in the plant.

3. Synthesis of Gasoline & Recovery

Synthesis gas from the methane splitting are fed to the synthesis section to make gasoline via the Fischer-Tropsch process. This process involves the catalytic reaction of carbon monoxide with hydrogen according to the following reactions:

$$2nH_2 + n\ CO = (CH_2)_n + n\ H_2O$$

$$nH_2 + 2n\ CO = (CH_2)_n + n\ CO_2$$

Catalyst is made from either millscale or pure magnetite in section 800, Catalyst Plant.

Gas and oil products of the synthesis are washed with water to remove water soluble chemicals. This water plus the reaction water and its dissolved chemicals comprise the feed to section 700, Chemical Recovery. The washed gas and oil become the feed to section 600, Product Recovery. Hydrocarbon products are recovered in section 600 through absorption/stripping operations. Light gases not recovered in the lean oil absorption are:

- vented to the fuel gas system of section 1100 for steam and power generation
- returned to the synthesis as aeration gas
- partly returned to section 400, Methane Splitting, for the conversion of light hydrocarbons to synthesis gas (external recycle stream)

4. Fuel Gas Manufacture - Steam & Power Generation

Fuel gas is manufactured by the pressure gasification of coal in steam-and-air-blown Lurgi gasifiers of section 1000, Fuel Gas Manufacture. Fuel gas manufactured is free of fly ash and is purified in much the same way that the synthesis gas was prepared and purified. Condensation products from synthesis gas and fuel gas manufacture are combined and treated. Section 1000 includes a hot carbonate scrubbing system which removes most of the carbon dioxide and hydrogen sulfide from the fuel gas. Foul gas scrubbed from the fuel gas contains hydrogen sulfide and is treated in section 1500, Sulfur Recovery, along with the hydrogen sulfide concentrate from synthesis gas purification. The sulfides are converted to elemental sulfur and recovered as a solid product. Some of the purified fuel gas is fed to gas turbines for the generation of power; the remainder is burned in process furnaces and in boilers raising steam for process and driver uses. Gas turbines exhaust into the fire box of the boiler and these gases are disposed of along with flue gas via power plant stacks. Section 1100 includes the treatment of water for the make-up to the boilers, aeration and blow-down. Cooling water inventory and circulation to the coolers requiring water cooling are provided by section 1700, which includes chlorination and blowdown. Makeup to this section is from section 1600, Water Treatment, where water is collected from various parts of the plant. For example, one stream to section 1600 is from section 1400, Effluent Water Treatment, where foul process condensate is purified by activated sludge waste water treating.

b. PROCESS CONSIDERATIONS

It should be noted that a SASOL-type plant as presented in this report is a plant representative of a twenty-year old technology. Improvements in this technology have been made by SASOL principally to suite their needs for fuels and chemicals. Over the years these needs have changed as new products, (e.g., pipeline gas, ammonia, ethylene, synthetic rubber, etc.) were added to the list manufactured from the raw material feeds, viz., coal, water and air. SASOL's continued study of processes in their complex has doubtless given much fundamental information that would be a real asset to a designer tailoring a coal-to-gasoline facility twenty or more years after the original design was committed. For circumstances plausible for such a plant, changes could be called for in the synthesis catalyst manufacturing, leading to a modified product spectrum to suit economic or other circumstances for the projected plant. Optimization of any such design would require information that is unlikely to be found from any source other than SASOL. Lacking access to SASOL's proprietary information for possible attempts at optimization of the facilities described in this report, the present coal-to-gasoline plant can be presented only on the basis of the twenty-year old technology with some minor updating modifications made public by SASOL in the intervening years.

It is significant that the major process steps of the SASOL plant have been proven to be compatible members of the complex and that the complex is amply proven successful for the manufacture of gasoline from coal. It follows that each major process unit is amply proven successful for its intended service.

Sections 200 and 1000 employ the Lurgi pressure gasification process developed by Lurgi Gesellschaft fuer Waermetechnik of Frankfurt, Germany. This process was selected for SASOL in the belief, now confirmed by both pilot plant and later commercial application, that this process and the South African coal were suited for each other. For the present study there is every reason to believe that the same suitability exists for the process and the Western coal chosen. Of course there are other gasification processes that might be used if properties of the coal contemplated were not suited for large-scale pressure gasification by the Lurgi process. Advantages or disadvantages of such processes could only be assessed from a complete knowledge of the coal properties plus possibly some actual test gasification work. Conceivably a coal in question would be handled most economically if pretreated to destroy possibly caking tendencies, or coked to recover byproducts of attractive sales value and then gasified by some process best suited for the product coke.

Gasification by the Lurgi process leads to the production of coal tars, oils, naphthas and phenols, that may or may not have attractive values, and also leads to the formation of methane in the gaseous product. The methane content of the fuel gas is an asset as it gives the fuel gas a higher heating value; however, for the synthesis gas methane is a liability as it requires catalytic splitting with oxygen and steam to supply the reactants needed in the Fischer-Tropsch synthesis. Conceivably a proper choice of gasifiers would involve a low pressure process for synthesis gas production and a pressure gasifier (e.g., Lurgi) for fuel gas production. On the other hand, the Lurgi gasifier with its excellent record at SASOL is not easily replaced at this time by less experienced processes unless it is not at all suited to the coal contemplated.

Gas purification is an important process step in the SASOL-type gasoline-from-coal plant. The Rectisol process first used commercially at SASOL has been an outstanding success and its performance has been phenonenal. In this single continuously-operating process, the wide-cut mixture of gases and vapors is freed of resin formers and objectionable sulfur compounds to mere trace proportions (less than 0.1 ppm total sulfur) while being separated from most of the carbon dioxide contained at about 30% concentration in the raw gas from the gasifiers. Extreme high purity is a requirement of the gas feeding to the iron catalyst of the synthesis, the same as it is for the iron catalyst of ammonia synthesis. This high purity is achieved in an absorber/stripper process in which the lean oil is a refrigerated polar solvent, e.g., methanol. For the present plant design a departure from the SASOL Rectisol design was made to achieve a more concentrated H_2S stream for sulfur recovery and a nearly sulfur-free stream for discharge to atmosphere. This departure involved the use of nitrogen from the oxygen plant for stripping of foul methanol. This is according to a variation of the Rectisol process proposed by Gesellschaft fuer Linde's Eismaschinen, Hoellriegelskreuth, Germany, developers of the Rectisol process.

Conversion of light hydrocarbons into Fischer-Tropsch reactants (H_2 + CO) is another important operation in the SASOL-type plant. The procedure whereby unwanted light hydrocarbons and methane are converted into reactants is a Kellogg development employing high temperature reforming of the hydrocarbons in a steam atmosphere, oxygen being used to supply the endothermic reaction heat. This is a regenerative process in which the reaction feed streams are heated by the reaction product stream. Unconverted steam is condensed as the product gas is cooled to cooling water temperature. This condensate is sent for treatment preparing it for reuse in steam raising. Methane splitting in this plant, as at SASOL, is accomplished in a fixed-bed catalytic reactor. The composition of the synthesis feed gas is adjusted in this reforming operation by proper choice of oxygen and steam rates corresponding to the inlet gas composition and rate.

Although the H_2/CO ratio of the synthesis feed leaving the methane splitter is essentially the ratio at which these reactants are consumed in the synthesis reactions, the ratio prevailing in the reactor is much higher. The reactor H_2/CO ratio is made higher by the internal recycle of the synthesis cycle, this gas being a light tail gas from the synthesis. The mixture of feed and internal recycle is fed to the high velocity fluid catalytic reactor especially developed by Kellogg for the highly exothermic Fischer-Tropsch reaction. The mixed stream, only slightly preheated above cooling water temperature, contacts hot fluid iron catalyst descending a standpipe leg of a catalyst loop circuit and becomes heated to a kindling temperature at which the reactions begin. This mixture of catalyst and reacting gases rises vertically through the reactor leg of the catalyst loop circuit, transferring reaction heat to an external cooling medium through surfaces built into the reactor leg. Although there is considerable shrinkage of the gas volume through reaction, and a corresponding reduction in velocity, the mixture travels completely through the vertical reactor and over an inverted U-bend to the separator vessel in which the catalyst disengages from the gas stream to complete the loop via the bottom-connected standpipe leg. The product distribution obtained in this reaction is dependent on the catalyst type and activity as well as on the outlet temperature and the reactor temperature profile. The high velocity fluid reactor puts the important variables in easy control of the operator.

Heat removed from the reacting stream traversing the reactor is used to raise steam at about 175 psig which in term is used in turbine drives for compressors, e.g., internal recycle compressors. Hot gas separated from the catalyst flows to an oil scrubber in which cooling and condensation is effected through direct contact with circulating oil, dumping heat to boiler feed water for boilers in the synthesis section. The oil circulation circuit is followed by a water circulation circuit that, through direct contact with the reactor gases and vapors, effects further cooling and condensation of reactor products. Tail gas following these condensation stages is split four ways: 1) a large stream of internal recycle; 2) a net stream that is to be subsequently processed into light gas for external recycle; 3) a purge stream which is used as fuel in steam and power generation section; and 4) a heavy hydrocarbon stream feeding product separators.

Bottoms of the oil scrubber contain catalyst fines washed down in the condensation process; these bottoms are returned to the catalyst circulation. Just above the catalyst settling zone of the scrubber, a heavy oil is decanted and sent from the unit as a separate stream for further processing. Light oil separated by gravity from the water circulation circuit is first water washed to remove water soluble chemicals and then sent from the unit for further processing. Tail gas is similarly water washed and then sent on to the product recovery area. Aqueous streams condensed from the reactor product join the wash water streams and become the feed to the Chemical Recovery section 700.

Heavy oil decanted above the catalyst settling zone of the synthesis unit oil scrubber enters section 600, Product Recovery, where it is flashed to remove its very heavy components. The lighter fractions of this oil join the lighter oil of the synthesis production and together the mixture undergoes vapor phase catalytic clay treatment for the removal of oil soluble chemicals and mild catalytic cracking of the synthetic crude molecules. The catalyst is regenerated in the usual way by burning off carbon with a mixture of air and nitrogen.

Synthesis tail gas from section 500, Synthesis, is passed directly to the absorber of the section 600, Product Recovery. Lean oil stripped in the lean oil distillation tower preferentially absorbs the heavier fractions of the gas and only a small amount of the lighter unwanted fractions. The unwanted fractions are partly purged from the system to remove the inerts (e.g., N_2 and Ar brought in with coal and oxygen) and mostly returned to the Methane Splitter, Section 400, for conversion to CO and H_2 for the synthesis. Absorbed components become feed for the catalytic polymerization unit which, under high pressure and in the presence of a catalyst, unites unsaturated molecules to form high octane gasoline molecules. Some excess of stripped molecules over those consumed by polymerization are liquefied and transfered to LPG storage for eventual marketing. Part of the cat poly feed is generated in the clay treating of the synthetic oil.

A fractionator separates the liquids recovered from clay treating and cat poly into gasoline, diesel oil and furnace, or waxy, oil.

The aqueous stream from section 500, Synthesis, is treated in section 700, Chemical Recovery, for the recovery of alcohols and ketones. The first separation is made to dispose of the acids and the bulk of the water. This mixture is sent to the activated sludge treatment unit in section 1400. Overhead product of the first separation is rich in alcohols and ketones and this stream is separated into two main streams in the following distillation tower. Both of these streams are processed further in a system of eight distillation towers and two hydrogenation reactors to yield ethyl alcohol, propyl alcohol, a stream of heavier alcohols and a mixture of ketones as intermediate products. The heavier alcohols are simply distilled to yield butyl alcohol, pentyl alcohol and a small residue of heavier alcohols, used as fuel. The mixture of ketones is first caustic treated then distilled to recover acetone and MEK (methylethyl ketone).

Ethyl alcohol is blended with the gasoline product for the octane benefits it supplies. All other products of section 700 are pumped to storage in the offsites. Section 600 products are pumped to storage also with the gasoline getting the treatment and/or additives it may require for marketing, e.g., color addition, inhibitor addition.

Sulfurous gases from sections 300, Gas Purification, and 1000, Fuel Gas Manufacture, become feed to the section 1500, Sulfur Recovery Unit, which, employing a Stretford solution, recovers the sulfur of H_2S as solid sulfur. The waste gas is vented through stacks. Air for the oxidation is supplied by a compressor incorporated in section 1500.

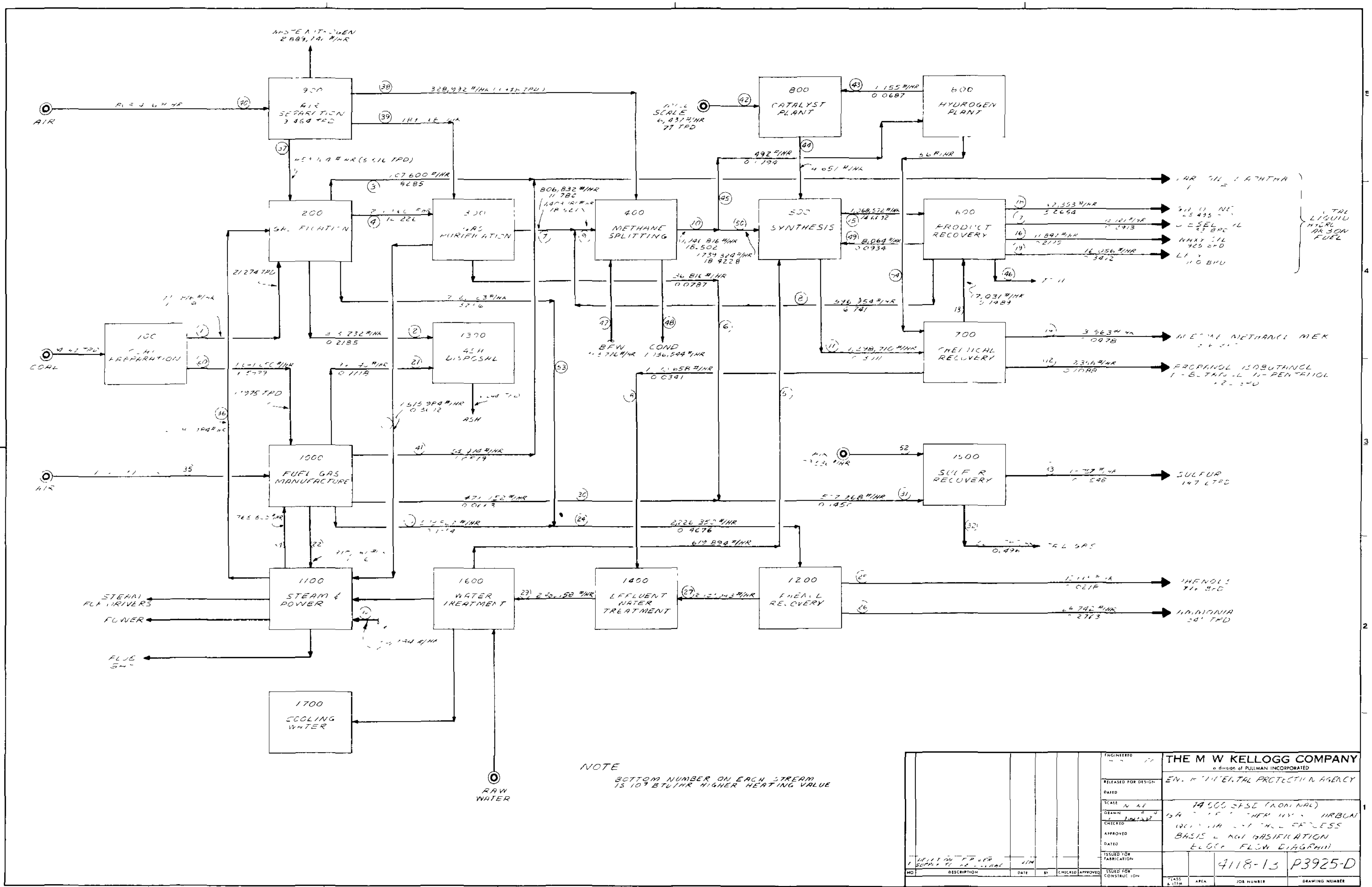
The steam and power section, 1100, is visualized in a location close to air separation, section 900. Large compressors of the air separation plant could thus be driven by gas turbines exhausting into the firebox of boilers and steam superheaters of section 1100. Steam systems have not been sketched for the plant and for the section 1100 in particular. It is very likely that steam generation, unlike steam generation at SASOL, will be at high pressure, e.g., 1500 psig. Superheated steam at this pressure likely will be sent to topping turbines for the generation of electrical power for the plant. Exhaust of these turbines will supply the process steam for gasification in sections 200 and 1000. Low level steam generated in section 500 will doubtless find use for turbine drives within the section, probably supplemented with the process steam level established by the gasifiers. Still lower level steam generated in waste heat boilers of the gasification plant probably will supply boiler feed water deaeration needs, reboiler duties, space heating requirements, etc. Condensates will be collected in separate tanks according to whether they are expected to be always clean or whether possibly contaminated by gases or liquids reaching the condensates, possibly through equipment leaks.

It is possible that this plant could be redesigned for a more interesting array of products from coal at a much reduced plant cost. The prospects would be dependent upon the product array sought and newer technology available than were at hand for the SASOL plant design. This newer technology may possibly exist with SASOL at the present or could result from the aims of research and development that may need to be undertaken. It is clear that the Methane Splitting could, for example, be eliminated if the desired product array included a reasonably high Btu gas. The external recycle would go to fuel gas. Conceivably the Lurgi gas issuing from the gasifiers could be reformed catalytically with oxygen addition at temperatures high enough to convert the tars, phenols, oils and methane to the reactants CO and H_2. Considerable savings would be made if this step could be taken. Of course, the incentive would depend on possible sales values of phenols and coal tar products.

Depending on product array desired, the catalyst may be profitably altered with possible changes in amount or kind of catalyst modifiers. Presumably there is knowledge and experience to guide steps that may be taken to this end development.

Considerable savings could doubtless be made if coal fines could be fired directly to the power plant and expensive stack gas cleaning were not a requirement for the power plant.

Tie-ins with other industries may have some interesting prospects for a gasoline-from-coal plant. Gases from a refinery handling natural crude possibly could be processed with gases from a synthesis plant to a mutual advantage. Gases or liquids from a steel mill might be exchanged with products of the synthesis to a mutual advantage. It should be remembered that the products of the synthesis are remarkably free of many troublesome contaminants, e.g., metals, sulfur. The probable octane of the gasoline from the subject plant is about 86 research. If octane levels were important, the scope of operations would probably have to be increased to include isomerization and alkylation.



900 AIR SEPARATION 3 464 TPD
WASTE NITROGEN 2 889,141 #/HR
200 GASIFICATION
300 GAS PURIFICATION
400 METHANE SPLITTING
500 SYNTHESIS
800 CATALYST PLANT
600 HYDROGEN PLANT
600 PRODUCT RECOVERY
700 CHEMICAL RECOVERY
100 COAL PREPARATION
1300 ASH DISPOSAL
ASH
1000 FUEL GAS MANUFACTURE
1500 SULFUR RECOVERY
1100 STEAM & POWER
1600 WATER TREATMENT
1400 EFFLUENT WATER TREATMENT
1200 PHENOL RECOVERY
1700 COOLING WATER
AIR
COAL
RAW WATER
BFW
COND 1 136,544 #/HR
POWER
SULFUR 147 LTPD
PHENOLS
AMMONIA
LIQUID HYDROCARBON FUEL
NOTE
BOTTOM NUMBER ON EACH STREAM IS 10^9 BTU/HR HIGHER HEATING VALUE
THE M W KELLOGG COMPANY
a division of PULLMAN INCORPORATED
14 500 STSD (NOMINAL)
BASIS LURGI GASIFICATION
4118-13
P3925-D

THE M. W. KELLGG COMPANY
A DIVISION OF PULLN INCORPORATED

TABLE 9 MATERIAL & ENERGY BALANCE

DATE
BY HBG

JOB NO. 4118-13

DESCRIPTION COAL-BASED SYNTHOL PLANT

CUSTOMER EPA

	COAL FEED	GASIF. ASH	GASIF. TON	RAW SYN. GAS	RECT. FUEL GAS	RECT. H_2S	PURIF. SYN GAS	RECYCLE GAS TO SYNTHOL	FEED TO METHANOL SPLIT	METHANE SPLITTER PROD.	SYNTHOL SOLVENTS	HVY ALCOHOLS	ETHANOL
	1	2	3	4	5	6	7	8	9	10	11	12	13
	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR
Moisture	292,304												
DAF Coal	1,173,856	6,576											
Ash	306,656	306,656											
#/HR	1,772,816	323,232	107,600	2,169,600	1,515,98	36,816	806,832	596,954	1,403,786	1,746,816	1,298,710	7,358	17,031
HHV Btu/HR x 10^9	15.47	0.2185	2.8285	12.2260	0.369	0.0787	11.780	6.741	18.521	18.502	0.3911	0.1008	0.1489
Tar			69,632										
Tar-Oil Naphtha			37,968										
Water											1,265,203		
Phenols													
Acids											5,455		
NH_3													
Sulfur													
LPG													
Gasoline													
Diesel Oil													
Waxy Oil													
Acetone											2,650		
Methanol											343		
Propanol											4,832	4,832	
Isobutanol											546	546	
n-Butanol											1,606	1,606	
MEK											670		
nPentanol											374	374	
Ethanol											17,031		17,031
	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH
CO_2				31,731.0	29,742.[illegible]	587.4	1401.6	5,315.2	6716.8	10,320.0			
H_2S				320.0		320.0	-	-	-	-			
COS						1.0	-	-	-	-			
C_2H_4				372.8	176.[illegible]		196.8	766.4	963.2	-			
CO				16,030.4	56.[illegible]		15,974.4	752.0	16,726.4	30,832.0			
H_2				39,988.4	65.[illegible]		39,923.2	17,428.8	57,352.0	93,744.0			
CH_4				12,176.0	201.[illegible]		11,974.4	9401.6	21,376.0	8,256.0			
C_2H_6				542.4	248.[illegible]		294.4	382.4	676.8	-			
N_2				417.6	6786.[illegible]	1.0	414.4	3963.2	4377.6	2,512.0			
Ar										592.0			
C_3H_6								192.0	192.0	-			
C_3H_8										-			
C_4H_8										-			
C_4H_{10}										-			
H_2Ov								67.2	67.2	880.0			
Light Oil										-			
Heavy Oil													
MPH Total				101,579.2	37,276.[illegible]	909.3	70,179.2	38,268.8	108,448.0	147,136.0			

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THE M W KELLOGG COMPANY
A DIVISION OF PULLMAN INCORPORATED

TABLE 9 CONT.

CUSTOMER

DATE

BY HBG

JOB NO. 4118-13

DESCRIPTION COAL-BASED SYNTHOL PLANT

	LIGHT SOLVENTS	SYNTHOL PROD.	WAXY OIL	DIESEL OIL	GASOLINE	LPG	COAL TO FUEL GAS GEN.	ASH FROM F.G. GEN.	LOW BTU F.G.	GAS LIQUOR	FEED TO PHENOSO.	PHENOLS	AMMONIA
	14	15	16	17	18	19	20	21	22	23	24	25	26
	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR
Moisture							178,278						
DAF Coal							715,941	10,110					
Ash							187,031	187,030					
#/HR	3,663	1,068,572	11,847	15,121	262,353	16,056	1,081,250	197,140	1,910,181	523,827	2,226,350	13,665	28,742
HHV Btu/HR x 10^9	0.0478	14.6692	0.2175	0.2913	5.2654	0.3412	9.5929	0.1118	7.2906	0.1354	0.4676	0.0218	0.2763
Tar													
Tar-Oil Naphtha													
Water										523,827**	2,183,942		
Phenols											13,665	13,665	
Acids													
NH_3											28,743		28,743
Sulfur													
LPG						16,056							
Gasoline					245,322								
Diesel Oil				15,121									
Waxy Oil			11,847										
Acetone	2,650												
Methanol	343												
n-Butanol													
MEK	670												
nPentanol													
Ethanol					17,031								
	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH
CO_2		6720.0							4,394.8				
H_2S									18.8				
COS													
C_2H_4		1024.0							245.2				
CO		976.0							16,500.0				
H_2		22,080.0							21,682.3				
CH_4		11,936.0							5,627.2				
C_2H_6		608.0							369.2				
N_2		4304.0							39,038.5				
Ar		512.0											
C_3H_6		1280.0											
C_3H_8		208.0											
C_4H_8		672.0											
C_4H_{10}		96.0											
C5±		336.0											
C6.5±		160.0											
H_2Ov		160.0											
Light Oil		1404.8											
Heavy Oil		51.2											
MPH Total		52,688.0							87876.9	** Gas Liquor			

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TABLE 9 CONT.

THE M. W. KELLOGG COMPANY
A DIVISION OF PULLMAN INCORPORATED

DATE

BY HBG

PAGE NO. 3 OF 5

JOB NO. 4118-13

DESCRIPTION COAL-BASED SYNTHOL PLANT

CUSTOMER EPA

	WATER FROM PHENOSOL-VAN	ACID WATER FROM 700	TREATED WATER	HOT CARB. H_2S	H_2S TO SRU	SRU TAIL GAS	SULFUR	STEAM TO F.G. GEN.	AIR TO F.G. GEN.	STEAM TO GASIFIER	O_2 TO GASIF.	O_2 TO METH. SPLIT	N_2 TO RECT.
	27	28	29	30	31	32	33	34	35	36	37	38	39
	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR
Moisture													
DAF Coal													
Ash													
#/HR	2,224,983	1,270,658	2,485,581	471,152	507,968	524,628	13,757	766,605	1,442,051	2,089,984	459,704	328,992	189,968
HHV Btu/HR x 10^9	-	0.0341	-	0.0663	0.1450	0.02496	0.05481	-	-	-	-	-	-
Tar													
Tar-Oil Naphtha													
Water	2,224,983	1,265,205	2,485,581					766,605	9,038	2,089,984			
Phenols													
Acids		5,453											
NH_3													
Sulfur							13,757						
LPG													
Gasoline													
Diesel Oil													
Waxy Oil													
Acetone													
Methanol													
n-Butanol													
MEK													
nPentanol													
Ethanol													
	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH
CO_2				10,468.8	10,860.4	11,101.5							
H_2S				192.2	452.3	22.6							
COS					0.8	0.8							
C_2H_4				2.1		2.1							
CO				34.8		34.8							
H_2				54.4		54.4							
CH_4				13.8		13.8							
C_2H_6				2.1		2.1							
N_2				91.4	93.6	902.1			38666.2		792.4	185.6	6784.5
Ar												20.8	
C_3H_6													
C_3H_8													
C_4H_8													
C_4H_{10}													
C5±													
C6.5±													
H_2Ov						429.7		42,589	502.1	116,110.2			
Light Oil													
Heavy Oil													
O_2									10344.5		13,660.0	10,092.8	
MPH Total				10,856.0	11,407.1	12563.5		42,589	49512.1	116,110.2	14,452.4	10,299.2	6784.5

TABLE 9 CONT.

DATE
BY HBG

PAGE NO. 4 OF 5
JOB NO. 4118-13

DESCRIPTION COAL-BASED SYNTHOL PLANT

CUSTOMER EPA

	AIR TO O_2 PLANT	F.G. GEN. TON	MILL-SCALE	H_2 TO CAT. PLANT	CAT. TO SYNTHOL	SYN. GAS TO H_2 PLANT	PURGE GAS	BFW TO METH. SPLIT	METH. SPLIT COND.	GAS RECYCLE TO SYN.	SYNTHOL FEED	WATER TO SYNTHOL	AIR TO SRU
	40	41	42	43	44	45	46	47	48	49	50	51	52
	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR
Moisture													
DAF Coal													
Ash													
#/HR	3,868,406	54,474	6,437	1155.2	4,651	7,492	150,784	953,776	1,036,544	8,064	1,739,324	619,894	29,536
HHV Btu/HR x 10^9	-	1.0519	-	0.0687	-	0.0794	1.675	-	-	0.0934	18.428	-	-
Tar		54,474											
Tar-Oil Naphtha													
Water								953,776	1,036,544			619,894	
Phenols													
Acids													
NH_3													
Sulfur													
LPG													
Gasoline													
Diesel Oil													
Waxy Oil													
Acetone													
Methanol													
n-Butanol													
Air													29,536
Millscale			6,437										
Catalyst					4,651								
	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH
CO_2						44.3	1332.8			72.0	10,275.7		
H_2S						-				-			
COS										-			
C_2H_4							193.6			4.8			
CO						132.2	212.8			11.2	30,699.8		
H_2				534.4		402.1	4412.8			238.4	93,341.9		
CH_4				5.4		35.4	2404.8			129.6	8,220.6		
C_2H_6							97.6			3.2			
N_2	101,669.0					10.8	852.3			46.4	2,501.2		
Ar	119.7					2.5	107.2			6.4	589.5		
C_3H_6							48.0			1.6			
C_3H_8													
C_4H_8													
C_4H_{10}													
C5±													
C6.5±													
H_2Ov	7742.0						19.2			11.2	876.2		
Light Oil													
Heavy Oil													
O_2	26,976.3					3.8							
MPH Total	136,507.0			539.8		631.1	9681.6			528.0	147,504.9		

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TABLE 9 CONT.

DESCRIPTION COAL-BASED SYNTHOL PLANT

CUSTOMER EPA

DATE

BY HBG

JOB NO. 4118-13

	GAS LIQUOR FROM GASIF.	H_2 TO SECT. 700
	53	54
	#/HR	#/HR
Moisture		
DAF Coal		
Ash		
#/HR	1,702,523	56
HHV Btu/HR x 10^9	0.3256	
Tar		
Tar-Oil Naphtha		
Water	1,675,006	
Phenols	8867	
Acids		
NH_3	18,650	
Sulfur		
LPG		
Gasoline		
Diesel Oil		
Waxy Oil		
Acetone		
Methanol		
n-Butanol		
MEK		
Millscale		
Catalyst		
	MPH	MPH
CO_2		
H_2S		
COS		
C_2H_4		
CO		
H_2		26
CH_4		0.3
C_2H_6		
N_2		
Ar		
C_3H_6		
C_3H_8		
C_4H_8		
C_4H_{10}		
C5 ±		
C6.5±		
H_2Ov		
Light Oil		
Heavy Oil		
O_2		
MPH Total		26.3

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B. Methanol-From-Coal (MWK Dwg. P3355-D)

A block flow diagram for the methanol-from-coal process is given in the cited drawing. The process may be divided into three steps:

- Coal Gasification and Raw Gas Purification
- Synthesis Gas Preparation
- Methanol Synthesis and Purification

1. Coal Gasification and Raw Gas Purification

This step is identical to the one in the gasoline-from-coal plant.

2. Synthesis Gas Preparation

Feed gas introduced into the methanol synthesis loop must contain the correct proportions of carbon monoxide, carbon dioxide and hydrogen according to the following overall reactions:

$$CO + 2H_2 = CH_3OH$$

$$CO_2 + 3H_2 = CH_3OH + H_2O$$

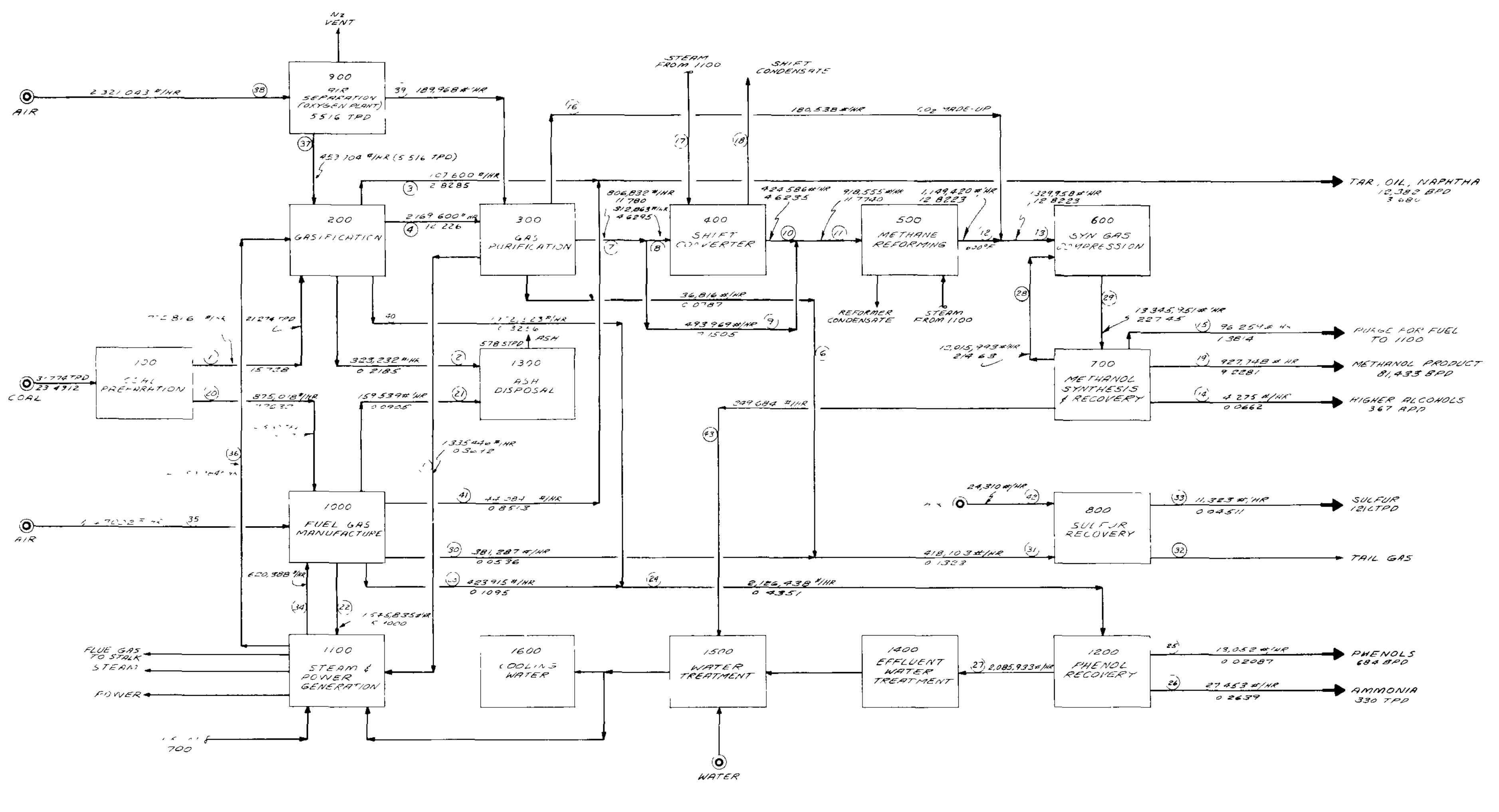
Synthesis gas composition will be adjusted by shift reaction and steam reforming. Raw synthesis gas from the purification section contains methane which is an unwanted component in the synthesis feed. The methane and other light hydrocarbon produced are split into hydrogen and carbon oxides by steam reforming. However, due to the relatively high concentration of carbon monoxide in the raw synthesis gas, carbon may be formed by disproportionation and deposit on the reforming catalyst surface and render the catalyst inactive. Part of the synthesis gas is sent to shift conversion and the remainder is by-passed. In the shift reaction carbon monoxide reacts with steam to form equivalent amounts of carbon dioxide and hydrogen. The shift reaction is exothermic and the shift effluent is cooled against boiler feed water. Water condensed in the cooling of the gas is separated and the cooled gas is then sent to the methane reforming unit. Steam reforming of methane is highly endothermic and heat is required as input to the reformer. Reformer effluent exits at about 1600°F, is cooled against the incoming feed and also used to boil feed water. The final synthesis gas is adjusted to the composition desired by a carbon dioxide make-up stream from the Rectisol unit.

3. Methanol Synthesis and Purification

Synthesis gas, containing hydrogen, carbon monoxide, carbon dioxide and small amounts of nitrogen, argon and methane, flows to the suction of the two-stage centrifugal feed compressor. The compressor is driven by a steam turbine using high pressure superheated steam. Fresh synthesis gas is compressed to about 1370 psig in two stages with intercooling between stages. Following cooling and separation of condensate, the compressed synthesis feed gas joins synthesis loop recycle gas containing about 0.3 percent methanol vapor. The combined stream is then compressed by a single-stage, turbine-driven centrifugal compressor to about 1485 psig and delivered to the methanol converter. Prior to entering the converter, a major portion of the feed flows to the interchanger which preheats the gas by exchange with the hot methanol converter effluent. The two overall reactions occuring in the converter are those associated with the combination of hydrogen and carbon monoxide to form methanol and the reaction of hydrogen and carbon dioxide to form carbon monoxide and water. Other side reactions involve the formation of dimethyl ether, ketones and higher alcohols. The hot effluent is cooled by the interchanger and again by water-cooled exchanger to 100°F thus condensing out the crude methanol product. The vapor/liquid stream then flows to the catchpot for separation of vapor from liquid. Disengaged vapor, containing unreacted hydrogen, carbon monoxide, carbon dioxide, methanol vapor, water and dimethylether flows to recycle compressor suction where it is combined with fresh feed make-up gas. Prior to compression, a proportion of the recycle gas is vented continuously to the fuel system to control the concentrations of methane, nitrogen and argon in the synthesis loop. These components would otherwise build up in the system and reduce the effective synthesis pressure, which would be reflected by lower methanol conversion per pass and reduced production capacity. Purge gas is delivered to the fuel gas system for power and steam generation.

Crude methanol from the catchpot, containing methanol, water and various impurities, flows to flash drums where the stream is flashed at 50 psig for removing the bulk of gas dissolved in the stream. Flash gas from this pressure reduction step also flows to the fuel system. Liquid from flash drums flow to the crude methanol storage tank.

Purification of methanol is accomplished in a two-tower distillation facility. The fractionation system consists of a topping column whose primary purpose is to remove light-end impurities such as dimethyl ether, ketones and aldehydes and a refining column to remove the heavy ends including ethanol and other higher alcohols from the methanol product. It should be noted that a two tower purification system may not be necessary for "fuel grade" methanol in which case the investment and operating cost would be reduced slightly.



NOTE:
BOTTOM NUMBER ON EACH STREAM IS 10^9 BTU/HR HIGHER HEATING VALUE

ENGINEERED: FKC	THE M. W. KELLOGG COMPANY, a division of PULLMAN INCORPORATED
RELEASED FOR DESIGN	ENVIRONMENTAL PROTECTION AGENCY
SCALE: NONE	81,000 BPSD (NOMINAL)
DRAWN: 3-18-74 JONES	METHANOL FROM COAL
CHECKED:	BASIS LURGI GASIFICATION
APPROVED / DATED	BLOCK FLOW DIAGRAM
JOB NUMBER	4118-13
DRAWING NUMBER	P3955-D

THE M. W. KELLOGG COMPANY
A DIVISION OF PULLMAN INCORPORATED

TABLE 10

DESCRIPTION 11338 TPD METHANOL-FROM-COAL MATERIAL BALANCE CUSTOMER EPA DATE 3/5/74 BY FC EST NO. 10F4 JOB NO. 4118-13

DESCRIPTION	COAL FEED	GASIF. ASH	TAR, OIL NAPHTHA	RAW SYN. GAS	RECT. FUEL GAS	RECT. H_2S GAS	PURIF. SYN. GAS	SHIFT CONV. FEED	SHIFT CONV. BY-PASS	SHIFT CONV. EXIT	FEED TO REFORMER	EXIT FROM REFORMER	SYN. G. TO COM-PRESSOR
STREAM	1	2	3	4	5	6	7	8	9	10	11	12	13
	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH
CO_2				31,731.0	25,640.2	587.4	1,401.6	550.8	850.8	6,514.1	7,364.9	10,401.7	14,503.7
H_2S				320.0	-	320.0	-	-	-	-	-	-	-
COS					-	1.0		-	-	-	-	-	-
C_2H_4				372.8	176.0		196.8	77.3	119.5	77.1	196.9	-	-
CO				16,030.4	56.0		15,974.4	6,277.9	9,696.5	312.5	10,009.0	16,971.9	16,971.9
H_2				39,988.4	65.6		39,923.2	15,689.8	24,233.4	21,654.4	45,887.8	78,229.3	78,229.3
CH_4				12,176.0	201.6		11,974.4	4,705.9	7,268.5	4,705.0	11,973.5	2,957.0	2,957.6
C_2H_6				542.4	248.0		294.4	115.7	178.7	115.7	294.4	-	-
N_2				417.6	6,787.7	1.0	414.4	162.9	251.5	162.9	414.4	414.4	414.4
AR													
C_3H_6													
H_2O (v)													
O_2													
CH_3OH													
BuOH, DME, ETC.													
TOTAL MPH				101,579.2	33,174.2	909.3	70,179.2	27,580.4	42,598.8	33,526.0	76,140.6	108,974.6	113,076.9
	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR
TAR, OIL, NAPHTHA			107,600										
WATER													
PHENOLS													
AMMONIA													
SULFUR													
ASH	306,656												
D.A.F. COAL	1,173,856	6,576											
MOISTURE	292,304	306,656											
TOTAL #/HR	1,772,816	323,232	107,600	2,169,600	1,335,446	36,816	806,832	312,863	493,969	424,586	918,555	1,149,420	1,329,958
HHV 10^9 BTU/HR	15.7280	0.2185	2.8285	12.226	0.3692	0.0787	11.7800	4.6295	7.1505	4.6235	11.7740	12.8223	12.8223

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TABLE 10 CONT.

THE M. W. KELLOGG COMPANY
A DIVISION OF PULLMAN INCORPORATED

DATE 3/5/74
BY FC

JOB NO. 4113-13

DESCRIPTION 11338 TPD METHANOL FROM COAL
CUSTOMER EPA

DESCRIPTION	HIGHER ALCOHOLS	PURGE FOR FUEL GAS	PURE CO_2 MAKEUP	STEAM TO SHIFT	STEAM CONDEN-SATE	METHANOL PRODUCT	COAL TO FUEL GAS GEN.	ASH FROM FUEL GAS GEN	FUEL GAS TO STEAM & POWER GEN.	GASGAS LIQUOR	FEED TO PHENOSOL-VAN	PHENOLS BYPRODUCT	AMMONIA BYPRODUCT
STREAM	14	15	16	17	18	19	20	21	22	23	24	25	26
	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH
CO_2		521.4	4,102.0						3,574.8				
H_2S		–							15.3				
COS									–				
C_2H_4		–							199.4				
CO		212.0							13,421.2				
H_2		3,173.8							17,637.5				
CH_4		2,903.3							4,577.2				
C_2H_6		–							300.3				
N_2		413.9							31,754.4				
AR		–											
C_3H_6		–											
H_2O (V)		9.5		7,533.0	1,567.8								
O_2		–											
CH_3OH		73.1				28,956.0							
BuOH, DME, ETC.	67.5	5.8											
TOTAL MPH	67.5	7,313.0	4,102.0	7,533.0	1,567.8	28,956.0			71,430.1				
	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR
TAR, OIL, NAPHTHA													
WATER										410,927	2,085,933		
PHENOLS										4185.0	13,052	13052.0	
AMMONIA										8803.0	27,453		27,453
SULFUR													
ASH							151357.6	151357.6					
DAF COAL							579386.0	8181.7					
MOISTURE							144,273.8						
TOTAL #/HR	4,275	96,258.8	180,537.8	135,745.	28,251.8	927,748.3	875,018	159,539	1,545,835	423,915	2,126,438	13,052	27,453
HHV 10^9 BTU/HR	0.0662	1.3814	–	–	–	9.2281	7.7632	0.0905	5.9000	0.0536	0.4351	0.02087	0.2639

TABLE 10 CONT.

THE M. W. KELLOGG COMPANY
A DIVISION OF PULLMAN INCORPORATED

DATE 3/5/74
BY FC

JOB NO. 4118-13

DESCRIPTION 11338 TPD METHANOL FROM COAL

CUSTOMER EPA

DESCRIPTION	WATER FROM PHENOSOLVAN	SYN. GAS RECYCLE	INLET TO METHANOL SYNTHESIS	HOT H_2S	H_2S TO SULFUR PLANT	TAIL GAS TO STACK	SULFUR BYPRODUCT	STEAM TO FUEL GAS MFG.	COMP. AIR TO FUEL GAS GEN.	STEAM TO GASIF.	O_2 TO GASIFIER	COMP. AIR TO OXYGEN PLANT	N2 TO RETISOL PLANT
STREAM	27	28	29	30	31	32	33	34	35	36	37	38	39
	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH
		-											
CO_2		45,937.0	60,440.7	8,472.1	9,059.5	9,137.5							
H_2S		-	-	155.5	475.5	18.6							
COS		-	-	-	1.0	1.0							
C_2H_4		-	-	1.7	1.7	1.8							
CO		31,836.8	48,808.7	28.2	28.2	29.1							
H_2		486,126.8	564,356.1	44.0	44.0	45.4							
CH_4		392,276.7	395,234.3	11.1	11.1	11.5							
C_2H_6		-	-	1.7	1.7	1.8							
N_2		61,577.9	61,992.3	74.1	75.1	742.5			31291.2		792.4	61001.4	6785.4
AR												71.8	
C_3H_6		-	-	-									
H_2O (V)		340.0	340.0	-				34,427.7	406.3	116,110.2		4645.2	
O_2		-	-	-					8371.5		13,660.0	16185.8	
CH_3OH		2,913.0	2,913.0	-									
BuOH, DME, ETC.		216.5	216.5	-									
TOTAL MPH		1,021,224	1134301.6	8786.7	9696.0	9989.2		34,427.7	40,069.2	116,110.2	14452.4	81,904.2	6785.4
	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR
TAR, OIL, NAPHTHA													
WATER	2,125,133												
PHENOLS													
AMMONIA							11,323						
SULFUR													
ASH													
DAF COAL													
MOISTURE													
TOTAL #/HR	2,125,133	12,015,993	13,345,951	381,287	418,103	431,816	11,323	620,388	1,167,002	2,089,984	459,704	2,321,043	189,968
HHV 10^9 BTU/HR	-	214.6311	227.4534	0.0536	0.1323	0.02055	0.04511	-	-	-	-	-	-

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THE M. W. KELLOGG COMPANY
A DIVISION OF PULLMAN INCORPORATED

TABLE 10 CONT.

DATE 3/5/73
BY FC

PAGE ~~NO.~~ 4 OF 4
JOB NO. 4118-13

DESCRIPTION 11338 TPD METHANOL FROM COAL

CUSTOMER EPA

DESCRIPTION		GAS LIQUOR FROM GASIFICATION	TAR, OIL NAPHTHA FROM FUEL GAS	AIR TO SULFUR PLANT	PROCESS WATER RECOVERED									
STREAM		40	41	42	43									
		MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH
CO_2														
H_2S														
COS														
C_2H_4														
CO														
H_2														
CH_4														
C_2H_6														
N_2				660										
AR														
C_3H_6														
H_2O (V)														
O_2				175.4										
CH_3OH														
BuOH, DME, ETC.														
TOTAL	MPH			835.4										
TAR, OIL, NAPHTHA			44,084											
WATER		1,675,006			249,684									
PHENOLS		8867.												
AMMONIA		18,650												
SULFUR														
ASH														
DAF COAL														
MOISTURE														
TOTAL	#/HR	1,702,523	44,084	24,310	249,684									
HHV	10^9 BTU/HR	0.3256	0.8513	-	-									

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C. Substitute Natural Gas-From-Coal (MWK Dwg. P3356-D)

A block flow diagram for the substitute natural gas (SNG)-from coal process is given in the cited drawing. The process may be divided into three steps:

- Coal Gasification and Raw Gas Purification
- Methane Synthesis
- Synthesis Gas Compression

1. Coal Gasification and Raw Gas Purification

This step is identical to the one in the gasoline-from-coal process which is described previously. The only difference from the gasoline-from-coal plant design is the addition of a shift conversion area which is designed to produce hydrogen by the "water gas shift" reaction:

$$CO + H_2O \quad = \quad CO_2 + H_2$$

Approximately one-half of the total crude gas from the gasifiers is subjected to shift conversion; the remainder is bypassed directly to the gas cooling area. The ratio of the two gas streams will be adjusted to achieve the desired H_2:CO ratio for proper feed to the methanation unit. Crude gas feed to the shift conversion area is quenched and washed first. The washed gas is heated in a series of heat exchangers before entering the first shift reactor where the bulk of carbon monoxide is catalytically converted to equivalent amount of hydrogen and carbon-dioxide. The first stage hot effluent is cooled in counter-current exchange with the feed gas before entering the second shift reactor where further conversion of carbon monoxide will take place. Effluent gas from the second shift reactor is cooled by indirect exchange with the feed gas before leaving the shift conversion unit.

2. Methane Synthesis

The methanation step converts low Btu synthesis gas to methane rich high Btu gas by the following overall chemical reactions:

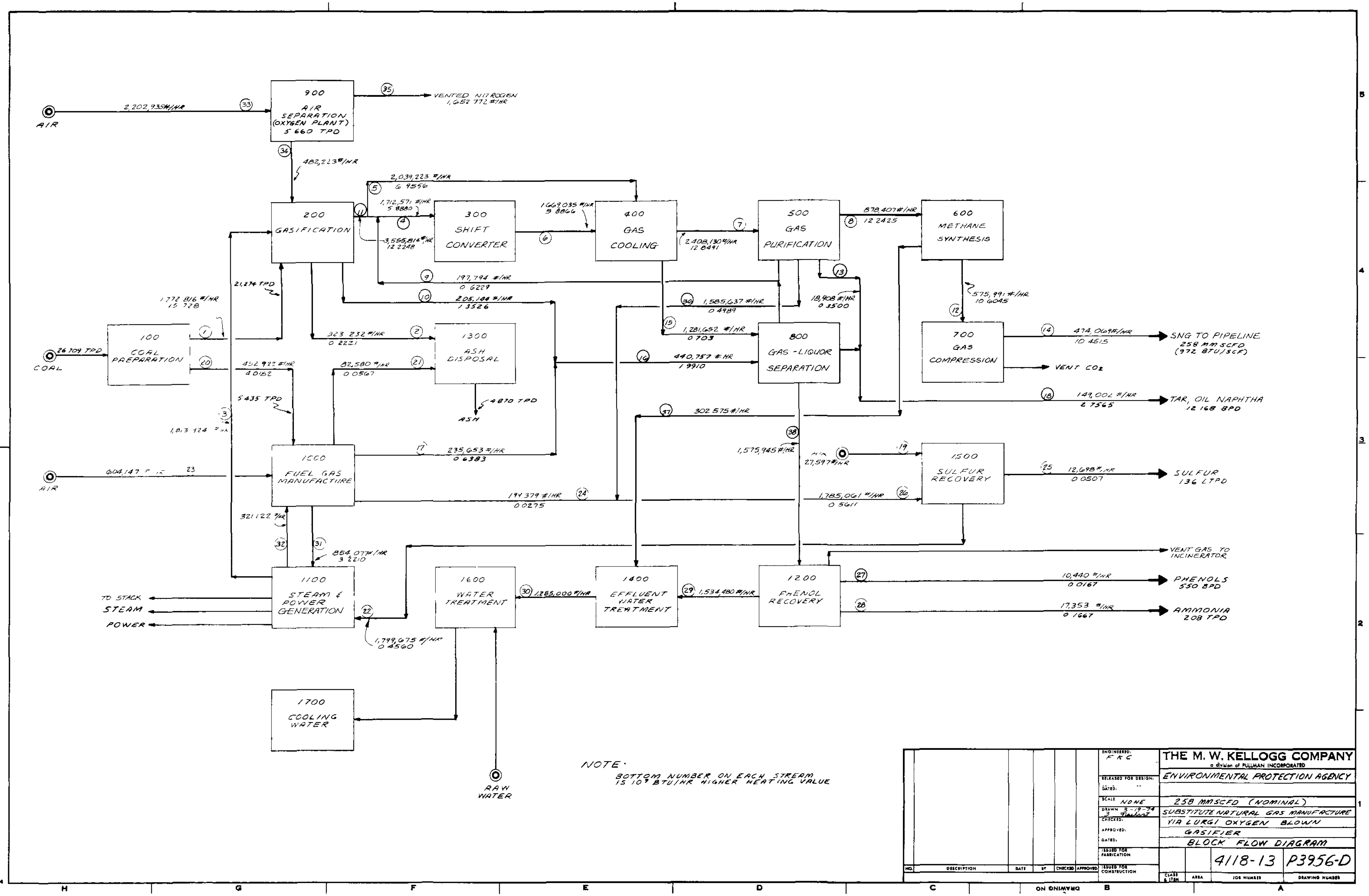
$$CO + 3H_2 \quad = \quad CH_4 + H_2O$$

$$CO_2 + 4H_2 \quad = \quad CH_4 + 2H_2O$$

Both of these reactions are highly exothermic and the heat released is used to heat the incoming feed gas as well as for steam generation in waste heat boilers. Hot feed gas, after indirect exchange with the product gas, is passed through a sulfur guard reactor to remove last traces of impurities before entering the synthesis loop. The synthesis loop consists of a methanator, waste heat boilers and a recycle compressor. Feed gas composition to the methanator will be set by combining the fresh feed gas stream with the gas stream circulated by the recycle compressor. Reaction heat from the methanator is removed in the high and low pressure waste heat boilers. Product gas from the synthesis loop is cooled in a feed/recycle product heat exchanger and further cooled in a final product cooler to ambient temperature. Condensed water is removed in a product-condensate separator.

3. Synthesis Gas Compression

Synthesis gas from the methanation area is compressed by a steam driven centrifugal compressor from 225 psia to 600 psia. The compressed gas is cooled to 90°F and sent to gas purification for final acid gas removal and dehydration. Gas from the gas purification area is returned to the second stage centrifugal compressor where it is boosted to pipeline pressure.


AIR
COAL
AIR
900 AIR SEPARATION (OXYGEN PLANT) 5 660 TPD
VENTED NITROGEN 1,652 772 #/HR
100 COAL PREPARATION
200 GASIFICATION
300 SHIFT CONVERTER
400 GAS COOLING
500 GAS PURIFICATION
600 METHANE SYNTHESIS
700 GAS COMPRESSION
800 GAS-LIQUOR SEPARATION
1000 FUEL GAS MANUFACTURE
1100 STEAM & POWER GENERATION
1200 PHENOL RECOVERY
1300 ASH DISPOSAL
1400 EFFLUENT WATER TREATMENT
1500 SULFUR RECOVERY
1600 WATER TREATMENT
1700 COOLING WATER
SNG TO PIPELINE 258 MM SCFD (972 BTU/SCF)
VENT CO2
TAR, OIL NAPHTHA 12 168 BPD
SULFUR 136 LTPD
VENT GAS TO INCINERATOR
PHENOLS 550 BPD
AMMONIA 208 TPD
ASH
TO STACK
STEAM
POWER
RAW WATER
NOTE: BOTTOM NUMBER ON EACH STREAM IS 10⁹ BTU/HR HIGHER HEATING VALUE
THE M. W. KELLOGG COMPANY
a division of PULLMAN INCORPORATED
ENVIRONMENTAL PROTECTION AGENCY
258 MMSCFD (NOMINAL)
SUBSTITUTE NATURAL GAS MANUFACTURE
VIA LURGI OXYGEN BLOWN
GASIFIER
BLOCK FLOW DIAGRAM
4118-13
P3956-D

THE M. W. KELLOGG COMPANY
A DIVISION OF PULLMAN INCORPORATED

TABLE 11

DESCRIPTION MATERIAL & ENERGY FLOW (SNG)

CUSTOMER EPA

DATE 3/14/74

BY FC

PAGE NO. 1 OF 3

JOB NO. 4118-13

DESCRIPTION	COAL TO GAS	ASH TO POND	STEAM AND BFW TO GAS	RAW GAS TO SHIFT	RAW GAS TO COOL-ING	CONVERTER GAS	RANGE GAS TO RECTI-SOL	METHANE SYN. FEED	RECYCLE FEED TO SHIFT	GAS LIQ-UOR EFF-LUENT	RAW GAS. FROM GAS.	GAS PROD-UCT TO COMPRESS.	NAPHTHA TO STORE
STREAM	1	2	3	4	5	6	7	8	9	10	11	12	13
	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH
CO_2				14,011	17,516	21,330	38,011	4,687	2,528		29,092	2,700	
H_2S				155	199	155	346	-	32		322	-	
C_2H_4				194	255	194	449	208	46		403	-	
CO				9,478	11,056	2,159	13,215	12,935	854		19,680	34.0	
H_2				18,816	21,913	26,135	48,048	47,588	1,661		39,068	212	
CH_4				5,377	6,421	5,377	11,798	11,678	635		11,164	27,604	
C_2H_6				296	371	296	667	311	53		614	-	
N_2 + AR				155	181	155	336	329	14		322	329	
O_2				-	-		-	-	-		-	-	
TOTAL DRY GAS				48,482	57,912	55,802	112,870	77,735	5,824		100,664	30,879	
MOL. WT.				21.27	21.78	20.85	21.14	11.30	26.31		21.27	18.60	
DRY GAS #/HR				1,031,218	1,261,323	1,163,470	2,386,072	878,407	153,233		2,141,121	574,350	
	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR
H_2O	292,315		1,813,924	638,378	731,646	478,169	3,150	-	44,561	114,650	1,325,463	1,641	-
COAL (DAF)	1,173,846	16,577		-	-	-	-	-	-	-	-	-	-
ASH	306,655	306,655		-	-	-	-	-	-	-	-	-	-
TAR, OIL NAPHTHA				40,811	43,925	27,397	18,908	-	-	69,631	84,737	-	18,908
PHENOLS				2,164	2,329	-	-	-	-	7,850	4,493	-	-
SULFUR				-	-	-	-	-	-	13,010	-	-	-
AMMONIA				-	-	-	-	-	-	-	-	-	-
TOTAL #/HR	1,772,816	323,232	1,813,924	1,712,571	2,039,223	1,669,035	2,408,130	878,407	197,794	205,141	3,555,814	575,991	18,908
HHV, 10^9 Btu	15.728	0.2221	-	5.8880	6.9556	5.8866	12.8491	12.2425	0.6229	1.3526	12.2248	10.6045	0.3500
MMSCFD (DRY)	-	-	-	441.6	528.3	508.2	1027.9	708	53.0	-	916.86	281.2	-
HHV (BTU/SCF DRY)	-	-	-	320	316	278	300	415	282	-	320	905	-

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THE M. W. KELLOGG COMPANY
A DIVISION OF PULLMAN INCORPORATED

TABLE 11 CONT'D.

DESCRIPTION MATERIAL & ENERGY BALANCE, (SNG)

CUSTOMER EPA

DATE 3/14/74
BY FC

JOB NO. 4118-13

DESCRIPTION	SNG TO PIPELINE	GAS LIQUOR TO SEPARATION	TAR, OIL TO SEPERATOR	GAS LIQUOR FROM FUEL	TAR, OIL NAPHTHA	AIR TO CLAUS PLANT	COAL TO FUEL GAS SECT.	ASH TO POND	CLAUS PLANT TAIL GAS	COMP. AIR TO GAS	ACID GAS TO CLAUS PLANT	SULFUR BYPRODUCT	FEED TO CLAUS
STREAM	14	15	16	17	18	19	20	21	22	23	24	25	26
	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH
CO_2	569	901							38,659		4,385		39,659
H_2S	-	8							**25		80		421
C_2H_4	-	-							209		1		209
CO	34	4							76		15		76
H_2	212	1							84		23		84
CH_4	27,192	-							264		6		263
C_2H_6	-	-							323		1		323
N_2 + AR	329.0	-				750			801	16,434	38		41
O_2						200			-	4,369	-		-
TOTAL DRY GAS	28,336	914				950			41,439	20,803	4,549.0		41,076
MOL. WT.	16.73	43.69				28.86			43.21	28.86	43.40		43.41
DRY GAS #/HR	474,069	39,935				27,424			1,790,587	600,360	197,427		1,783,109
	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR
H_2O	-	1,206,665	320,882	206,232		173	74,681		9,088	3,787	1,952		1,952
COAL (DAF)	-	-	-				299,896	4,235	-	-	-		-
ASH	-	-	-				78,345	78,345	-	-	-		-
TAR, OIL, NAPHTHA	-	37,975	92,119	22,488	149,002		-	-	-	-	-		-
PHENOLS	-	-	10,440	2,590			-	-	-	-	-		-
SULFUR	-	-	17,353	4,344			-	-	-	-	-	12,698	-
AMMONIA	-	-	-				-	-	-	-	-	-	-
TOTAL, #/HR	474,069	1,281,652	440,794	235,653	149,002	27,597	452,922	82,580	1,799,675	604,147	199,379	12,698	1,785,061
HHV 10^9 Btu	10.4531	0.703	1.9910	0.6383	2.7565	-	4.0182	0.0567	0.4560	-	0.0275	0.0507	0.5611
MMSCFD (DRY)	258.1	-	-	-	-	-	-	-	377.4	-	41.48	-	374.1
HHV (BTU/SCF DRY)	972	-	-	-	-	-	-	-	29	-	15.9	-	36
LT/D												136	

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TABLE 11 CONT'D.

THE M. W. KELLOGG COMPANY
A DIVISION OF PULLMAN INCORPORATED

DATE 3/14/74
BY FC

JOB NO. 4118-13

DESCRIPTION MATERIAL & ENERGY BALANCE, (SNG)
CUSTOMER EPA

DESCRIPTION	PHENOLS BYPRODUCT	AMMONIA BYPRODUCT	H_2O EFFLUENT FROM PHENSOLVAN	TO WATER TREATMENT	LOW BTU GAS	STEAM TO AIR GASIF.	AIR TO O_2 PLANT	O_2 TO GASIFIER	N_2 VENT	ACID GAS TO CLAUS	WATER CONDENSATE	FEED TO PHENOSOLVAN	
STREAM	27	28	29	30	31	32	33	34	35	36	37	38	
	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	
CO_2					1,841					35,274			
H_2S					8					341			
C_2H_4					103					208			
CO					6,912					61			
H_2					9,084					61			
CH_4					2,357					257			
C_2H_6					155					322			
N_2 + AR					16,353		58,427	301	58,126	13			
O_2							15,239	14,740	499	-			
TOTAL DRY GAS					36,811		73,666	15,041	58,625	36,527			
MOL. WT.					21.64		28.82	32.06	28	43.41			
DRY GAS #/HR					796,600		2,123,571	482,223	1,641,500	1,585,637			
	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR
H_2O			1,534,480	1,285,000	46,353	321,122	79,364		11,272		302,575	1,534,480	
COAL (DAF)					-	-					-	-	
ASH					-	-					-	-	
TAR, OIL, NAPHTHA					**11,123	-					-		
PHENOLS	10,440				-	-					-	10,440	
SULFUR	-				-	-					-	-	
AMMONIA	-	17,353			-	-					-	17,353	
							2,202,935	482,223	1,652.772	1,585,637	302,575	1,562,273	
TOTAL, #/HR	10,440	17,353	1,534,480	1,285,000	854,077	321,122	-	-	-	0.4948	-	0.1886	
HHV, 10^9 Btu	0.0167	0.1667	-	-	3.2210	-				332.6		-	
MMSCFD (DRY)	-	-	-	-	336.4	-	-	-	-	36.	-	-	
HHV (BTU/SCF DRY)	-	-	-	-	229.8	-							
LT/D													
**IN GAS STREAM													

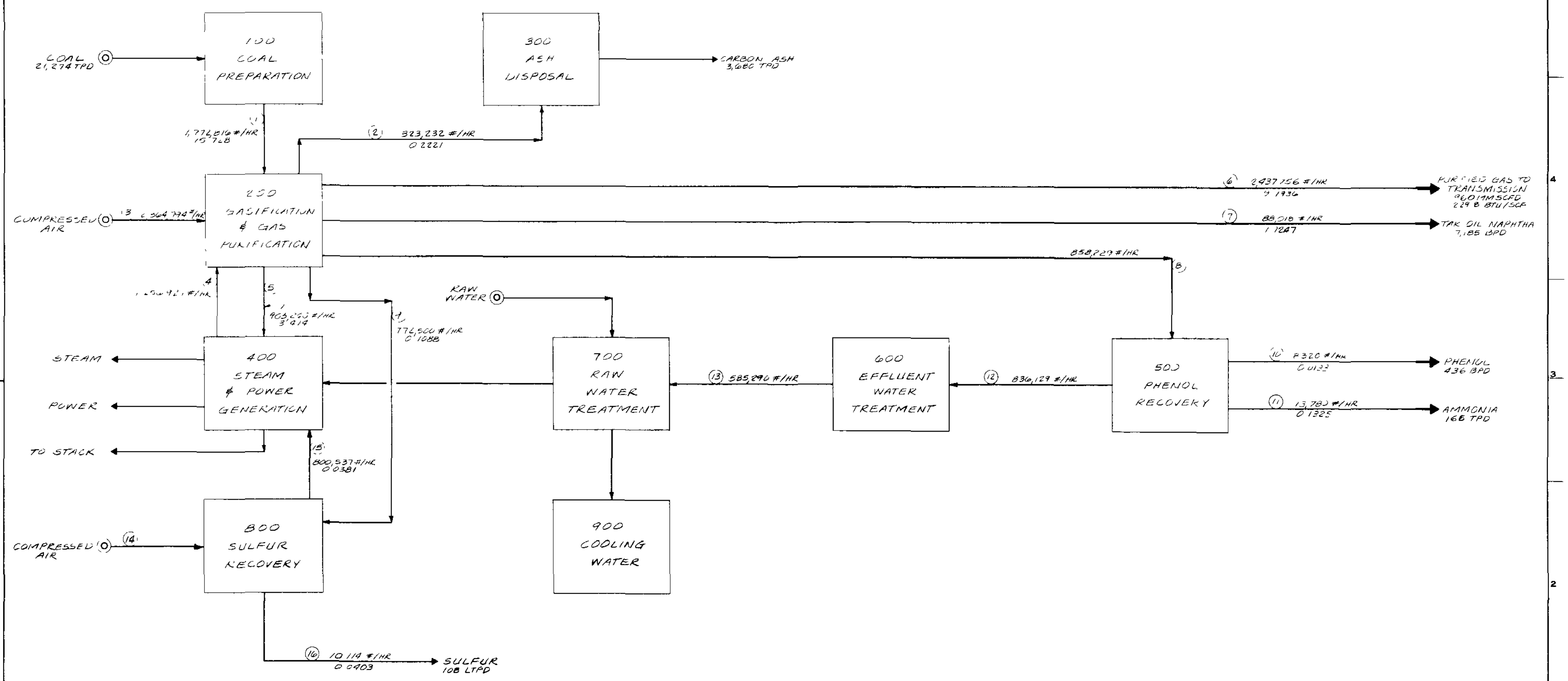
COOA 11-11-64

D. Low Btu Gas-From-Coal (MWK Dwg. P3357-D)

A block flow diagram for the low Btu gas-from-coal process is given in the cited drawing. The air blown gasification process is adopted from Lurgi's design.

Coal is conveyed from the coal preparation area to coal bunkers located above the coal gasifiers. Coal is fed to the gasifiers through coal locks which are pressured by a slip stream of the raw gas. Hot compressed air and process steam is mixed and introduced into the gasifiers. Ash is removed at the bottom of the gasifiers through ash locks and transported to ash disposal. Hot raw gas leaving the gasifiers is cooled by quenching with a gas liquor spray in wash coolers. Raw gas from the wash cooler is further cooled and cleaned by gas liquor in wash scrubbers. A purge stream of gas liquor is sent to the phenol recovery section. After cooling, the gas liquor is flashed to atmospheric pressure in an expansion vessel to remove dissolved gases. Coal tar is separated from the gas liquor by gravity and sent to product storage. A portion of the clarified gas liquor is recycled to the wash scrubber as make-up. The remainder is sent to phenol recovery where dissolved ammonia and phenol will be removed.

Expansion gas and coal lock vent gas are compressed and combined with the raw gas. Desulfurization of raw gas is accomplished by a hot potassium carbonate system in which hydrogen sulfide and the bulk of carbon dioxide are removed. Acid gas from the regenerator is cooled and sent to the sulfur recovery section. Part of the fuel gas produced is sent to steam and power generation section for process requirement. The remainder of the fuel gas is transmitted to pipeline as primary product.



NOTE
BOTTOM NUMBER ON EACH STREAM IS 10^9 BTU/HR HIGHER HEATING VALUE

ENGINEERED	F K C
SCALE	NONE
	THE M W KELLOGG COMPANY a division of PULLMAN INCORPORATED
	ENVIRONMENTAL PROTECTION AGENCY
	LOW BTU GAS MANUFACTURE VIA LURGI AIR BLOWN GASIFIER 960 MMSCFD (NOMINAL)
	BLOCK FLOW DIAGRAM
JOB NUMBER	4118-13
DRAWING NUMBER	P3957-D

TABLE 12

THE M. W. KELLOGG COMPANY
A DIVISION OF PULLMAN INCORPORATED

DESCRIPTION LOW BTU GAS MATERIAL BALANCE

CUSTOMER EPA

DATE 3/18/73

BY FC

PAGE NO. 1 OF 2

JOB NO. 4118-13

DESCRIPTION	COAL TO GASIFICA-TION	ASH TO POND	COMPRESSED AIR TO GAS	STEAM & WATER TO GAS	LOW BTU GAS FOR POWER GEN	LOW BTU GAS PRO-DUCT	TAR, OIL, NAPHTHA	GAS LIQ-UOR TO PHE-NOLSOLVAN	TO SULFUR RECOVERY	PHENOLS BYPRODUCT	AMMONIA BYPRODUCT	WATER EFF-LUENT PHE-NOLSOLVAN	RECYCLE WATER
STREAM	1	2	3	4	5	6	7	8	9	10	11	12	13
	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH
CO_2					1,951	5,254			17,165				
H_2S					8	23			315				
C_2H_4					109	293			4				
CO					7,326	19,728			6				
H_2					9,628	25,928			89				
CH_4					2,298	6,729			22				
C_2H_6					164	441			4				
N_2 + AR			64,326		17,332	46,675			150				
O_2			17,101		-	-			-				
TOTAL DRY			81,427		39,017	105,069			17,755				
MOL. WT.			28.86		21.64	21.64			43.51				
TOTAL, #/HR			2,349,971		844,320	2,273,702			772,500				
	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR
H_2O	292,315				49,131	132,305		836,129				836,129	585,290
COAL (DAF)	1,173,846	16,577	14,823	1,256,927	-	-							
ASH	306,655	306,655			-	-							
TAR, OIL, NAPHTHA					11,790	31,749	88,018						
PHENOL					-	-		8,320		8,320			
SULFUR					-	-							
AMMONIA					-	-		13,780			13,780		
TOTAL #/HR	1,772,816	323,232	2,364,794	1,256,927	905,240	2,437,756	88,018	858,229	772,500	8,320	13,780	836,129	585,290
HHV 10^9 BTU	15.728	0.2221	-	-	3.414	9.1936	1.7247	-	0.1088	0.0133	0.1324	-	-
MMSCFD (DRY)	-	-	-	-	356.5	960.1	-	-					
HHV (BTU/SCF DRY)					230	230							

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THE M. W. KELLOGG COMPANY
A DIVISION OF PULLMAN INCORPORATED

TABLE 12 CONT'D.

DESCRIPTION LOW BTU GAS FLOW RATE

CUSTOMER EPA

DATE 3/18/73

BY FC

JOB NO. 4118-13

DESCRIPTION	COMPRESSED AIR	TAIL GAS TO STEAM GEN.	SULFUR BYPRODUCT
STREAM	14	15	16
	MPH	MPH	MPH
CO_2		17,165	
H_2S		16	
C_2H_4		4	
CO		6	
H_2		89	
CH_4		22	
C_2H_6		4	
N_2 + AR	563	1,372	
O_2	149	-	
TOTAL DRY	712	18,678	
MOL. WT.	28.86	42.57	
TOTAL, #/HR	20,559	795,150	
	#/HR	#/HR	#/HR
H_2O	139	5,387	
COAL (DAF)			
ASH			
TAR, OIL, NAPHTHA			
PHENOL			
SULFUR			10,114
AMMONIA			
TOTAL #/HR	20,698	800,537	10,114
HHV 10^9 BTU	-	0.0381	0.0403
MMSCFD (DRY)			

COOA 11 11-64

V. Discussion of Results and Recommendations

Using the specific processes shown in this report, gasoline, methanol, substitute natural gas and low Btu gas can be manufactured from coal via a SASOL-type plant which utilizes Lurgi air- and oxygen-blown gasification processes.

Cost ($/MMBtu) of manufacturing such products decreases in the following manner:

Gasoline > methanol > SNG > Low Btu Gas

Incidentally, the order shown also represents the degree of flexibility of these products. Gasoline and methanol being in liquid form are less expensive for transportation and storage. Gasoline having a higher heating value is a superior product because it is the only proven and widely used automotive fuel and has a greater market demand than methanol. SNG in turn is superior to low Btu gas because it can be transported as pipeline quality gas whereas low Btu gas cannot be transported long distances and its use is therefore restricted to close-coupled plants (e.g., utility boilers).

With respect to the technology involved, gasoline-from-coal plant has been operated commercially by SASOL for more than twenty years. At present a SNG-from-coal facility using Lurgi air-and oxygen-blown gasification processes is yet to be built. Although several are in design stages, large-scale methanol from coal plants are only in planning discussion. Thus, the technology of gasoline-from-coal is ahead of the other coal conversion processes in that a commercial plant is in operation.

Methanol unquestionably can be manufactured from coal cheaper than gasoline; however, the applicability of methanol as a fuel should be explored carefully. Studies have shown that a methanol-gasoline mixture of up to 10% methanol by volume burns more efficiently than gasoline in automobiles and the emissions of unburned hydrocarbons, carbon monoxides, nitrogen oxides are reduced drastically (5). In view of the current U.S. crude shortage of about 8%, methanol can be explored as an additive to gasoline provided the conversion cost of the engine for burning gasoline-methanol mixture is insignificant. Alternatively, the best use of methanol may be to displace fuel oil and natural gas from utility and industrial boilers. The fuel oil thus displaced could be converted to gasoline.

SNG or low Btu gas can be produced cheaper than gasoline or methanol. However, either SNG or low Btu gas is basically a different form of fuel and has a different applicability than gasoline and methanol. Direct comparison of the costs should only be made after one establishes usefulness of the products as well as market demand.

It should be noted that both this report and Task 13 Preliminarly Report did not include any optimization studies because of the limited scope of the task. For example, when the desired end product is SNG or low Btu gas, Lurgi gasification technology is probably the best. For gasoline-or methanol-from-coal plant where methane is an undesirable product in the raw gas synthesis, however, the use of other gasification processes such as Kopper-Totzek gasification unit (which produces practically negligible methane) should be investigated for possible savings.

The use of fuel gas from coal gasification for steam and power generation is undesirable if low sulfur coal can be burned directly for this purpose. If high sulfur fuels are used for auxiliary steam and power generation, installation of a stack gas scrubbing unit may be required.

Production of a singular specialized product may not be the best utilization of the SASOL-type process as reflected by SASOL facility which includes the manufacture of a wide spectrum of chemicals, plastics, oil, gas and fertilizers.

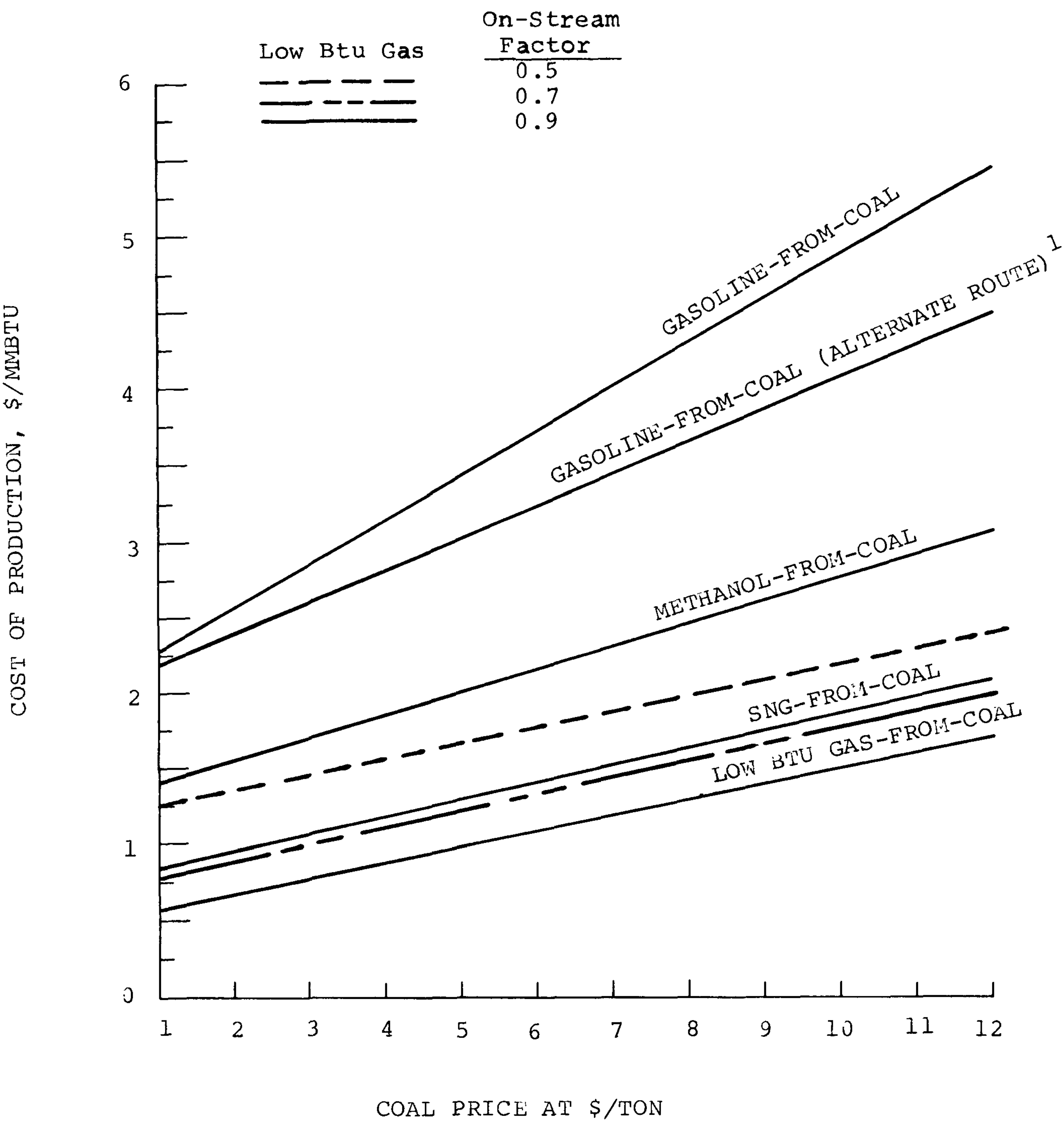
In any event, the cost of manufacturing gasoline and methanol from coal can be lowered by proper optimization of the process. The gasoline-from-coal plant via a SASOL-type process merits a much more detailed optimization study since it is the only available gasoline-from-coal technology today. Such study should include the latest operating technique of the SASOL plant as well as the Synthol process by M.W. Kellogg.

The costs presented in this report should be considered as having budget type accuracy. Also, the values presented in the appendices for coal cost, hourly wage, interest rate, etc., are assumed to be typical for a general evaluation but would need to be refined for a specific application. Figure 5 presents the cost of the products for various coal costs as well as the sensitivity of on-stream factors on the low Btu gas cost.

The basis of the modified Panhandle Eastern accounting procedure used in deriving the cost figures is given in Appendix E. This short cut method is intended to be used for the financing of utility plant and may not be appropriate for the gasoline- or methanol-from-coal plants, both of which are of chemical and refinery type operation. In the absence of a comparative accounting procedure for chemical and refinery plant, the modified Panhandle Eastern accounting procedure is used throughout the report to generate compatible production costs of gasoline and methanol. An estimated incremental increase of 1% in the fixed total capital charge will raise the unit gasoline and methanol costs by 5% (with by-product credits). For example, using a sinking fund method with a 15 year plant life, the fixed capital charge including depreciation, interim replacements, insurance, tax and cost of capital is 18.22% (6). The resulting gasoline and methanol unit costs will be $4.05/MMBtu and $2.34/MMBtu respectively, corresponding to an increase of about 30% over the previous prices derived from the modified Panhandle Eastern accounting procedure. An estimated incremental increase of 1% in the fixed total capital charge will raise the unit SNG cost by 5% and unit low Btu gas by 4% (with by-product credits). Using the same sinking fund method, the unit SNG and low Btu gas costs will be \$1.50/MMBtu and \$1.10/MMBtu respectively.

FIGURE 5: EFFECT OF COAL PRICE ON THE PRODUCTION COST OF GASOLINE, METHANOL, SNG & LOW BTU GAS

Mine-Mouth Coal Cost
On-Stream Factor = 0.9
For Gasoline, Methanol & SNG



1. Maximize gasoline production by including the conversion of tar, oil, naphtha

VI. Reference

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2. Govaarts, J. H., and Schutte, C. W. (SASOL), "The Use of Low Grade Coal for the Production of Oil, Gas, Fertilizers and Chemicals," Eigth World Energy Conference, Bucharest, June 28-July 2, 1971, paper No. 3.3-187.

3. El Paso Natural Gas Company application to Federal Power Commission for Burnham Coal Gasification Complex in New Mexico, November 7, 1972.

4. "The Supply - Technical Advisory Task Force - Synthetic Gas - Coal", Final Report, April 1973.

5. Reed, T. B., and R. M. Lomer, Science, Volume 182, December 28, 1973, Number 4119.

6. "Evaluation of SO_2-Control Process", Task No. 5 Final Report, submitted to Environmental Protection Agency, Office of Air Programs, Division of Control Systems by M. W. Kellogg Co., Contract No. CPA 70-68. October 15, 1971.

7. "Steam-Electric Plant Factors", 1973 Edition, National Coal Association, Washington, D.C.

Appendix A

Gasoline-From-Coal Via Synthol Process
Total Capital Requirement

		1975 M$
Total Direct & Indirect Cost of Plant (Incl. Contractor & Eng. Fees, Tax & Licenses)		505,000
Contingency		47,000
Total Plant Investment		552,000
Interest During Construction Interest Rate (9.0%) x Total Plant Investment x 1.875 years average period		93,000
Plant Start-Up Cost 40% of Operating Cost for 1/2 year		9,000
Working Capital	M$	
Coal @ $3.60/ton** (64 day supply)	7,900	
Catalyst & Chemicals (60 day supply)	1,000	
Receivables less Payable (1/24 of Annual Revenue from Gasoline @ $3.05/MMBtu)	5,200	
Total Working Capital		14,000
Total Capital Requirement		668,000

**Mine-Mouth Coal Cost.
This figure is taken from the average coal cost in New Mexico as reported by Steam-Electric Plant Factors, 1973 Edition, National Coal Association with escalation of 10% per year to 1975.

Appendix A (Cont'd.)

Gasoline-From-Coal Via Synthol Process
Annual Operating Cost

On-stream factor = 0.9

		1975, M$/year
1.	Raw Materials	
	Coal @ $3.60/ton	40,000
2.	Purchased Utilities	
	Power	---
	Raw Water	500
3.	Labor	
	A. Operating Labor @ $8/hr	10,800
	B. Maintenance Labor (1.5% of Total Plant Investment)	8,300
	C. Supervision (0.15 of A + B)	2,900
4.	Supplies	
	A. Operating Catalyst & Chemicals	8,000
	B. Maintenance (1.5% of Total Plant Investment)	8,300
5.	Administration & General Overheads	
	60% of Total Labor Including Supervision	13,200
6.	Tax & Insurance at 2.7% of Total Plant Investment	15,000
7.	Total Operating Cost (Without By-Product Credits)	107,000

Appendix A (Cont'd.)

Gasoline-From-Coal Via Synthol Process
Annual Operating Cost

On-stream factor = 0.9

			MS/Yr.	1975, M$/Year
8.	By-Product Credits			
	A. Tar, Oil, Naphtha	($8/Barrel)	34,770	
	B. Phenols	($70/Ton)	3,770	
	C. Ammonia	($50/Ton)	5,670	
	D. Sulfur	($10/LT)	480	
	E. Higher Alcohols	($100/Ton)	3,040	
	F. Acetone	($150/Ton)	1,570	
	G. M.E.K.	($200/Ton)	530	
	H. Diesel Oil	($10.5/Barrel)	4,250	
	I. Waxy Oil	($7.5/Barrel)	2,280	
	J. LPG	($6.5/Barrel)	4,270	
	Total By-Product Credit		60,600	
9.	Net Operating Cost (With By-Product Credit)			46,400

Unit Costs Base Case

10.	Gasoline Cost (with By-product Credits)	
	A. $/MMBtu	3.05
	B. $/Barrel	15.11
11.	Gasoline Cost (without By-product Credits)	
	A. $/MMBtu	4.55
	B. $/Barrel	22.52

Unit Costs Alternate Case

12.	Gasoline Cost (with By-product Credits)	
	A. $/MMBtu	2.76
	B. $/Barrel	13.70

Appendix A (Cont'd.)

Gasoline-From-Coal Cost

For 20 year Average Price
Without Escalation (Based on
Shortcut Method on Panhandle
Eastern Accounting Procedure)**

Gasoline Cost

= (Net Operating Cost + 0.1198 x Total Capital Requirement + 0.0198 x Working Capital)/Gasoline Production

Gasoline Production

= 41.5×10^6 MMBtu/Year (8.375 MM Barrels)

$$\text{Gasoline Cost} = \frac{46.4 + 0.1198 \times 668 + 0.0198 \times 14}{41.5}$$

= \$3.05/MMBtu (with By-Product Credits)

= \$15.11/Barrel (with By-Product Credits)

$$\text{Gasoline Cost} = \frac{107 + 0.1198 \times 680 + 0.0198 \times 14}{41.5}$$

= \$4.55/MMBtu (Without By-Product Credits)

= \$22.52/Barrel (Without By-Product Credits)

**Final Report of the Supply-Technical Advisory Task Force - Synthetic Gas From Coal, April, 1973

Appendix A (Cont'd.)

Alternate Gasoline-From-Coal Cost
(By Further Processing the Tar
Oil, Naphtha to Gasoline Product)

Gasoline Production = 25495 + 13230 x 0.8 BPD

= 11.85×10^6 MM Barrel/Year

= 58.7×10^6 MMBtu/Year

Annual Operating Cost (Deletion of Tar, Oil, Naphtha By-Product Credits)

= \$81.2 Million/Year

Total Capital Required = \$552 + \$93 + \$16 + \$14 Million

= \$675 Million

$$\text{Alternate Gasoline Cost} = \frac{81.2 + 675 \times 0.1198 + \$14 \times 0.0198}{58.7}$$

= \$2.76/MMBtu (With Byproduct Credits)

= \$13.70/Barrel (With Byproduct Credits)

Appendix B

Methanol-From-Coal
Total Capital Requirement

		1975, M$
Total Direct& Indirect Cost of Plant (Incl. Contractor & Engr. Fees Tax & Licenses)		472,000
Contingency		42,000
Total Plant Investment		514,000
Interest During Construction Interest Rate (9.0%) x Total Plant Investment x 1.875 Average Year Period		87,000
Plant Start-Up Cost 40% of Operating Cost for 1/2 year		11,000
Working Capital	M$	
Coal @ $3.60/ton** (64 days supply)	7,320	
Catalyst & Chemical (60 days supply)	800	
Receivables Less Payable (1/24 of annual Revenue from Methanol @$1.80/MMBtu	5,460	
Total Working Capital		14,000
Total Capital Requirement		626,000

** Mine-Mouth Coal Cost. This figure is taken from the average coal cost in New Mexico as reported by Steam-Electric Plant Factors, 1973 Edition, National Coal Association with escalation of 10% per year to 1975.

Appendix B (Cont'd.)

Methanol-From-Coal
Annual Operating Cost
On Stream Factor = 0.9

				1975, M$/Yr.
1.	Raw Materials			
	Coal at $3.60/ton			37,600
2.	Purchased Utilities			
	Power			--
	Water			400
3.	Labor			
	A. Operating Labor at $8/hr			10,800
	B. Maintenance Labor (15% of Total Plant Investment)			7,700
	C. Supervision (0.15 of A+B)			2,800
4.	Supplies			
	A. Operating Catalyst & Chemicals			6,000
	B. Maintenance (1.5% of Total Plant Investment)			7,700
5.	Administration & General Overheads			
	60% of Labor Including Supervision			12,800
6.	Tax & Insurance at 2.7% of Total Plant Investment			13,900
7.	Total Operating Cost (Without By-Product Credits)			99,700
8.	By-Product Credits		M$/Yr.	
	A. Tar Oil, Naphtha	($8/barrel)	32,500	
	B. Phenols	($70/ton)	3,600	
	C. Ammonia	($50/ton)	5,400	
	D. Sulfur	($10/LT	400	
	E. Higher Alcohols	($100/ton)	1,900	
	Total By-Product Credits			43,800
9.	Total Net Operating Cost			55,900

Unit Costs

10.	Methanol Cost (with By-product Credits)	
	A. $/MMBtu	1.80
	B. $/Barrel	4.90
11.	Methanol Cost (without By-product Credits)	
	A. $/MMBtu	2.42
	B. $/Barrel	6.58

Appendix B (Cont'd.)

Methanol-From-Coal Cost

For 20 year average price without escalation
(Based on Shortcut Method on Panhandle
Eastern Accounting Procedure)

Cost of Methanol

$$= \frac{\text{(Net Operating Cost + 0.1198 x Total Capital Requirement + 0.0198 x Working Capital)}}{\text{Methanol Production}}$$

Methanol Production

$= 72.75 \times 10^6$ MMBtu/year (26.75 MMBarrels)

Cost of Methanol

$$= \frac{55.9 + 0.1198 \times 626 + 0.0198 \times 14}{72.75}$$

= \$1.80/MMBtu (With By-product Credits)

= \$4.90/Barrel (With By-Product Credits)

Cost of Methanol

$$= \frac{99.7 + 0.1198 \times 635 + 0.0198 \times 14}{72.75}$$

= \$2.42/MMBtu (Without By-Product Credit)

= \$6.58/Barrel (Without By-Product Credit)

Appendix C

Substitute Natural Gas Production Total Capital Requirement

		1975,M$
Total Direct & Indirect Cost of Plant (Incl. Contractor & Engr. Fees, Tax & Licenses)		365,000
Contingency		34,000
Total Plant Investment		399,000
Interest During Construction (Interest Rate (9.0%) x Total Plant Investment x 1.875 years average period)		67,000
Plant Start-up Cost 40% of Operating Cost for 1/2 year		7,000
Working Capital **	M$	
Coal @ 3.60/ton (64 day supply)	6,150	
Catalyst & Chemicals (60 days supply)	500	
Receivables Less Payables - 1/24 of annual revenue at $1.13/MMBtu)	3,900	
Total Working Capital		11,500
Total Capital Requirement		484,000

** Mine-Mouth Coal Cost. This figure is taken from the average coal cost in New Mexico as reported by Steam-Electric Plant Factors, 1973 Edition, National Coal Association with escalation of 10% per year to 1975.

Appendix C (Cont'd.)

SNG Annual Operating Cost
Stream Factor = 0.9

		1975, M$/year
1.	Raw Material	
	Coal at $3.60/ton	31,590
2.	Purchased Utilities	
	Power @ 0.8¢/KWH	---
	Raw Water	300
3.	Labor	
	A. Operating Labor at $8.00/hr	5,980
	B. Maintenance Labor (1.5% of total plant investment)	5,990
	C. Supervision (15% of A + B)	1,780
4.	Supplies	
	A. Operating Catalyst & Chemicals	3,000
	B. Maintenance @ 1.5% of Total Plant Investment	5,990
5.	Administration & General Overhead	
	60% of total labor including supervision	8,250
6.	Taxes & Insurance at 2.7% of total Plant Investment per Year	10,773
7.	Total Operating Cost (Without By-Product Credits)	73,650
8.	By-Product Credits	

	M$/Yr.
A. Tar, Oil, Naphtha ($8/Barrel)	31,960
B. Crude Phenols ($70/ton)	2,880
C. Ammonia ($50/ton)	3,420
D. Sulfur ($10/LT)	440

		1975, M$/year
	Total By-Product Credit	38,700
9.	Net Annual Operating Cost	35,000

Unit Costs

10.	Substitute Natural Gas (with By-Product Credits)	
	A. $/MMBtu	1.13
11.	Substitute Natural Gas (without By-Product Credits)	
	A. $/MMBtu	1.60

Appendix C (Cont'd.)

SNG Gas Cost

For 20-year Average Gas Price Without Escalation (Based on Short-Cut Method on Panhandle Eastern Accounting Procedure)

Gas Price

$$= \text{(Net Operating Cost} + 0.1198 \times \text{Total Capital Requirement} + 0.0198 \times \text{Working Capital)} / \text{Gas Production}$$

Gas Production

$$= 82.4 \times 10^{6} \text{ MMBtu/year}$$

Gas Price

$$= \frac{35.0 + 0.1198 \times 484.0 + 0.0198 \times 11.0}{82.4}$$

$$= \$1.13/\text{MMBtu (With By-Product Credit)}$$

Gas Price

$$= \frac{73.650 + 0.1198 \times 494.0 + 0.0198 \times 11.0}{82.4}$$

$$= \$1.60/\text{MMBtu (Without By-Product Credit)}$$

Appendix D

Low Btu Gas Production
Total Capital Requirement

		1975, M$
Total Direct & Indirect Cost of Plant (Incl. Contractor & Engr. Fees Tax & Licenses)		218,000
Contingency		20,000
Total Plant Investment		238,000
Interest During Construction (Interest Rate (9%) x Total Plant Investment x 1.875 Years Average Period)		40,000
Plant Start-Up Cost 40% of Operating Cost for 1/2 year		5,400
Working Capital **	M$	
Coal at $3.60/ton (64 day supply)	4,900	
Catalyst and Chemicals (60 day supply)	300	
Receivable Less Payable (1/24 of Annual Revenue at $0.8/MMBtu)	2,600	
Total Working Capital		7,800
Total Capital Requirement		291,000

** Mine-Mouth Coal Cost. This figure is taken from the average coal cost in New Mexico as reported by Steam-Electric Plant Factors, 1973 Edition, National Coal Association with escalation of 10% per year to 1975.

Appendix D (Cont'd.)

Low Btu Gas Production
Annual Operating Cost
On Stream Factor = 0.9

			1975M$/Year
1.	Raw Materials		
	Coal at $3.60/ton (Catalyst & Chemicals included with supplies)		25,160
2.	Purchased Utilities		
	Power		--
	Raw Water		200
3.	Labor		
	A. Operating Labor at $8/hr		3,990
	B. Maintenance Labor at 1.5% of Total Plant Investment		3,570
	C. Supervision @ 15% of A+B		1,130
4.	Supplies		
	A. Operating Catalyst & Chemicals		2,000
	B. Maintenance @ 1.5% of total plant investment		3,570
5.	Administration & General Overhead		
	(60% of total Labor Including Supervision)		5,210
6.	Tax & Insurance at 2.7% of Total Plant Investment per year		6,430
7.	Total Operating Cost (Without By-Product Credits)		51,260
8.	By-Product Credits	M$/Yr.	
	A. Tar, Oil, Naphtha ($8/barrel)	18,880	
	B. Crude Phenols ($70/ton)	2,300	
	C. Ammonia ($30/ton)	2,710	
	D. Sulfur ($10/LT)	350	
	Total By-Product Credit		24,240
9.	Net Annual Operating Cost		27,000

Unit Cost

10.	Low Btu Gas (with By-Product Credits)	
	A. $/MMBtu	0.86
11.	Low Btu Gas (without By-Product Credits)	
	A. $/MMBtu	1.20

Low Btu Gas Production
Gas Cost

For 20-year Average Gas Price Without Escalation (Based on Short-Cut Method on Panhandle Eastern Accounting Procedure)

Gas Price = (Net Operating Cost + 0.1198 x Total Capital Requirement + 0.0198 x Working Capital)/Gas Production

Gas Production = 72.5×10^6 MMBtu

$$\text{Gas Price} = \frac{(27 + 0.1198 \times 291 + 0.0198 \times 7.8)}{72.5}$$

= \$0.86/MMBtu (With By-Product Credits)

$$\text{Gas Price} = \frac{(51.26 + 0.1198 \times 296 + 0.0198 \times 7.8)}{72.5}$$

= \$1.20/MMBtu (Without By-Product Credit)

Appendix D (Cont'd.)

Low Btu Gas Production
Annual Operating Cost

Alternate Case with On Stream Factor = 0.7

			1975M$/Year
1.	Raw Materials		
	Coal at $3.60/ton (Catalyst & Chemicals included with supplies)		19,570
2.	Purchased Utilities		
	Power		--
	Raw Water		200
3.	Labor		
	A. Operating Labor at $8/hr		3,990
	B. Maintenance Labor at 1.5% of Total Plant Investment		3,570
	C. Supervision @ 15% of A+B		1,130
4.	Supplies		
	A. Operating Catalyst & Chemicals		2,000
	B. Maintenance @ 1.5% of total plant investment		3,570
5.	Administration & General Overhead		
	(60% of total Labor Including Supervision)		5,210
6.	Tax & Insurance at 2.7% of Total Plant Investment per year		6,430
7.	Total Operating Cost (Without By-Product Credits)		45,670
8.	By-Product Credits	M$/Yr.	
	A. Tar, Oil, Naphtha ($8/barrel)	14,700	
	B. Crude Phenols ($70/ton)	1,800	
	C. Ammonia ($30/ton)	2,100	
	D. Sulfur ($10/LT)	300	
	Total By-Product Credit		18,900
9.	Net Annual Operating Cost		26,800

	Unit Cost
10. Low Btu Gas (with By-Product Credits)	
A. $/MMBtu	1.10
11. Low Btu Gas (without By-Product Credits)	
A. $/MMBtu	1.44

Low Btu Gas Production
Gas Cost

Alternate Case with On Stream Factor = 0.70

For 20-year Average Gas Price Without Escalation (Based on Short-Cut Method on Panhandle Eastern Accounting Procedure)

Gas Price = (Net Operating Cost + 0.1198 x Total Capital Requirement + 0.0198 x Working Capital)/Gas Production

Gas Production = 56.4×10^6 MMBtu

$$\text{Gas Price} = \frac{(26.8 + 0.1198 \times 291 + 0.0198 \times 7.8)}{56.4}$$

= \$1.10/MMBtu (with By-Product Credits)

$$\text{Gas Price} = \frac{(45.67 + 0.1198 \times 296 + 0.0198 \times 7.8)}{56.4}$$

= \$1.44/MMBtu (without By-Product Credit)

Appendix E

Description of the Panhandle Eastern Accounting Procedures**

Basis:

- 20-year project life
- 5% straight line depreciation on Total Capital Requirement excluding Working Capital
- 48% federal income tax rate
- Debt/equit ratio of 75%/25%
- 9% percent interest on debt
- 15% percent return on equity

Derived Parameters:

- Rate Base = Total Capital Requirement less Accrued Depreciation (includes 1/2 depreciation for given year)
- Percent Return on Rate Base = Fraction Debt x Percent Interest + Fraction Equity x Percent Return on Equity

Calculated Cash Flows in Given Year:

- Return on Rate Base = Rate Base x (Percent Return on Rate Base ÷ 100)
- Return on Equity = (Fraction Equity x Rate Base) x (Percent Return on Equity ÷ 100)
- Federal Income Tax = Return on Equity x (Percent Tax Rate ÷ [100 - Percent Tax Rate])
- Depreciation = 0.05 x (Total Capital Requirement - Working Capital)
- Total Revenue Requirement in Given Year = Return on Rate Base + Federal Income Tax + Depreciation + Total Net Operating Cost

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Appendix E. (Cont'd.)

Costs of Production:

- In given year: Total Revenue Requirement ÷ Annual Production
- 20-year average: Total Revenue Requirement Over Project Life ÷ (20 x Annual Production)

Derivation of General Cost Equation

Definition of Terms:

C = Total Capital Requirement, Million \$
W = Working Capital, Million \$
N = Total Net Operating Cost, Million \$
G = Annual Production, 10^{12} Btu/year
d = Fraction Debt
i = Percent Interest on Debt
r = Percent Return on Equity
p = Percent Return on Rate Base
n = Year, 1 to 20
RR_n = Total Revenue Requirement in n^{th} Year

Calculate Rate Base in n^{th} Year:

Depreciable Investment = C-W
Accrued Depreciation @ Mid-Point of Year = 0.05(n-0.5)(C-W)
Rate Base = C - 0.05(n-0.5)(C-W)

Calulate Percent Return on Rate Base:

$$\begin{aligned} p &= (d)i + (1-d)r \\ &= 0.75 \times 9 + 0.25 \times 15 \\ &= 10.5 \end{aligned}$$

Calculate Cash Flows in n^{th} Year:

Return on Rate Base = $0.01\ p[C-0.05(n-0.5)(C-W)]$

Return on Equity = $0.01\ r\ (1-d)\ [C-0.05\ (n-0.5)(C-W)]$

Federal Income Tax = $\frac{48}{52}\ 0.01\ r\ (1-d)[C-0.05(n-0.5)(C-W)]$

Depreciation = $0.05(C-W)$

Total Net Operation Cost = N (excluding escalation)

Total Revenue Requirement (RR_n) =

$$0.01\ p\ [C-0.05(n-0.5)(C-W)]$$
$$+\frac{48}{52}\ 0.01\ r\ (1-d)\ [C-0.05(n-0.5)(C-W)]$$
$$+\ 0.05\ (C-W) + N$$

$$RR_n = N + 0.05\ (C-W)$$
$$+\ 0.01\ [p + \frac{48}{52}\ (1-d)\ r]\ [C-0.05(n-0.5)(C-W)]$$

Calculate Production Cost in n^{th} Year:

Gas Cost in n^{th} Year = RR_n/G (\$/MMBtu)

Calculate 20-Year Total Revenue Requirement (excluding escalation):

$$\sum_{n=1}^{20} RR_n = 20N + (C-W)$$
$$+\ 0.01\ [p + \frac{48}{52}(1-d)\ r]\ [20C-0.05(200)(C-W)]$$

$$\text{Total RR} = 20N + (C-W)$$
$$+\ 0.1[p + \frac{48}{52}(1-d)\ r](C+W)$$

Appendix E (Cont'd.)

Calculate 20-Year Average Production Cost Without Escalation:

Average Production Cost

$$= \text{Total RR} / (20 \times G)$$

$$= \frac{N + 0.05\ (C-W) + 0.005\ \left[p + \frac{48}{52}\ (1-d)\,r\right]\ (C+W)}{G}$$

$$= \frac{N + 0.05\ (C-W) + 0.005\ \left[10.5 + \frac{48}{52}\ (0.25)\ (15)\right]\ (C+W)}{G}$$

$$= \frac{N + 0.05\ (C-W) + 0.0698\ (C+W)}{G}$$

or

$$\text{Average Production Cost (\$/MMBtu)} = \frac{N + 0.1198\ C + 0.0198\ W}{G}$$

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7. AUTHOR(S) F.K. Chan		8. PERFORMING ORGANIZATION REPORT NO.
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16. ABSTRACT The report gives results of a study to assess costs and feasibility of manufacturing gasoline, methanol, SNG, and low-Btu gas from coal, using the SASOL-type process. This process is based on a SASOL plant which has been operated commercially for more than 20 years for the manufacture of gasoline, fertilizers, and other chemicals from coal in South Africa. The SASOL plant has been modified slightly to suit the product spectrum of the projected plants. Capital investments for plants producing various end products are estimates based on published or in-house information on a mine-mouth plant using Western U.S. coal. The capital investment is expressed in 1975 dollars with no forward escalation. The total capital requirement and the unit production cost are based on a shortcut version of the Panhandle Eastern Accounting Procedure, recommended for coal conversion facilities. The capital investment, as well as the unit production cost, decreases as follows: gasoline > methanol > SNG > low-Btu gas. The SASOL-type process utilizes Lurgi air- and oxygen-blown gasification systems exclusively; comparison with other gasification systems is not included.

17. KEY WORDS AND DOCUMENT ANALYSIS		
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