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SCREENING EVALUATION: SYNTHETIC LIQUID FUELS MANUFACTURE. FINAL REPORT

PARSONS (RALPH M.) CO., PASADENA, CALIF

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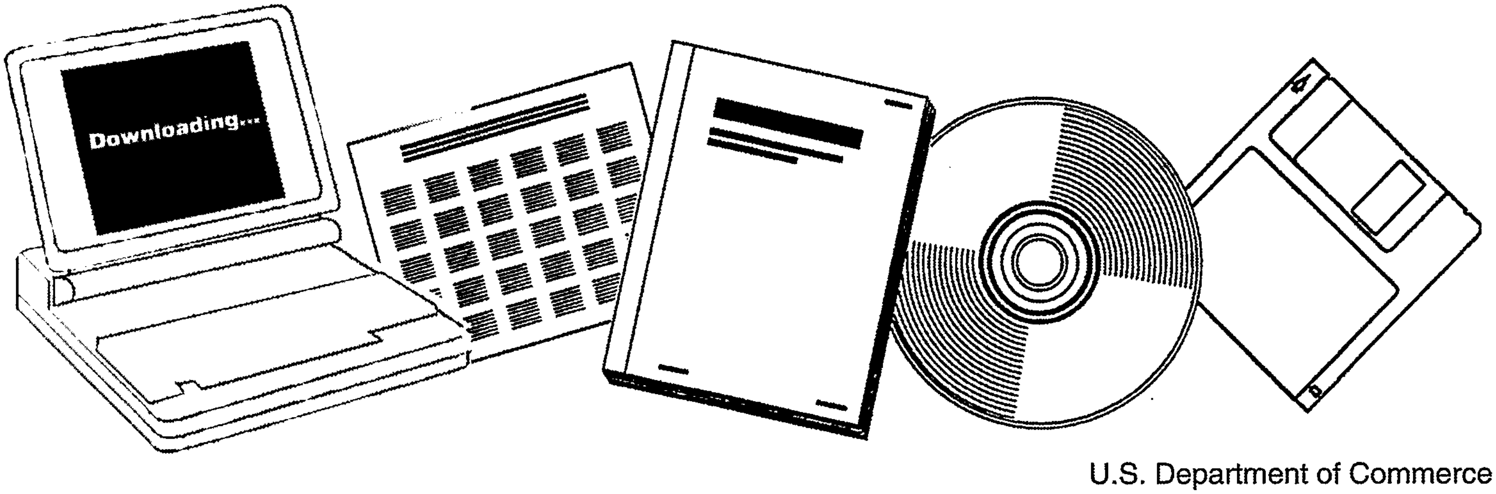

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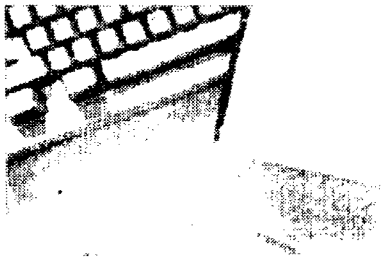

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SCREENING EVALUATION: SYNTHETIC LIQUID FUELS MANUFACTURE

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(Research Project 715)

Final Report

August 1977

Prepared by

THE RALPH M. PARSONS COMPANY
100 West Walnut Street
Pasadena, California 91124

PRINCIPAL INVESTIGATOR
T. K. Chow
D. W. Stanbridge

PROJECT MANAGER
G. A. White

Prepared for

Electric Power Research Institute
3412 Hillview Avenue
Palo Alto, California 94304

EPRI Project Managers
S. A. Vejtasa
N. Korens

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ABSTRACT

The objective of this screening study was to compare various available and proposed gasification processes in combination with one methanol process in conceptual commercial size plant designs for production of clean liquid fuel from coal. Additionally, a Fischer-Tropsch liquids plant design was included for direct comparison with a methanol case using one gasification process.

The gasification processes screened in this study were entrained coal gasifiers representing commercial (Koppers-Totzek), near commercial (Texaco) and advanced concepts (Foster-Wheeler), as well as a moving bed slagging gasifier (British Gas-Lurgi) now under development. The Chem Systems ebullating catalyst bed methanol process was used in combination with the above gasifiers. The British Gas-Lurgi gasifiers were used with the Fischer-Tropsch synthesis plant.

The results of this study indicate that there is an economic advantage to the advanced gasification systems for production of methanol. They also show that there is an economic advantage to production of methanol over that of Fischer-Tropsch liquids when the same type of gasifiers is used in both plants.

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SECTION 1

INTRODUCTION

Methanol and Fischer-Tropsch liquid products are potentially attractive fuels for power generation because of their favorable storage characteristics and their compatibility with environmental standards. As part of a continuing program to evaluate synthetic fuels from coal, the Electric Power Research Institute (EPRI) authorized The Ralph M. Parsons Company to perform a screening study of processes for the production of methanol and Fischer-Tropsch liquids. These synthetic liquid fuels are produced from synthesis gases which are generated by the gasification of coal. The scope of this study is limited to a screening evaluation of several gasification processes in combination with methanol and Fischer-Tropsch synthesis processes. The level of detail used in the screening study is not sufficient to allow comparison with more detailed studies on other fuels.

In an effort to evaluate a total plant concept for the production of liquid fuels from synthesis gases, facilities for coal gasification and syngas purification are included. The evaluation is based on a comparison of five cases; in four of the cases methanol is produced by the Chem Systems process and, in the remaining one case, Fischer-Tropsch liquids are produced. For methanol production, four gasification processes are used:

1. Foster-Wheeler (F-W)
2. Slagging-Gasifier (British Gas Council-Lurgi)
3. Kopper-Totzek (K-T)
4. Texaco

For Fischer-Tropsch products, the Slagging Gasifier system was selected.

In a screening type evaluation, it is beyond the scope of work to develop completely consistent heat and material balances for all cases considered. In the present study, certain input data was accepted from licensors whereas some balances were prepared by Parsons. In all cases, an effort was made to base synthesis gas production including gas composition and yield structure on an Illinois No. 6 coal. Sufficient work was performed to assure reasonable consistency between cases and where appropriate, a comment is offered where the degree of confidence may be stronger in one case than another but where adjustments in balances by Parsons were not warranted. Such comments are included only as a basis for the exercise of further judgment by the reader and are not intended as a criticism of the bases used in the preparation of such preliminary information.

SECTION 2

BASES FOR EVALUATION

The immediate objective of this screening study is to identify major differences or advantages in one process concept over others. Using this approach, it was hoped that future studies could be concentrated on those concepts having the greatest potential for capital reduction and improved thermal efficiency.

As a result of preliminary study work, it was agreed between Parsons and EPRI to select five cases for development in the study, four methanol cases and one Fischer-Tropsch case.

Case 1 - Foster-Wheeler Gasification - Chem Systems Methanol.
Case 2 - British Gas Council/Lurgi Slagger - Chem Systems Methanol.
Case 3 - Koppers-Totzek Gasification - Chem Systems Methanol.
Case 4 - Texaco Gasification - Chem Systems Methanol.
Case 5 - British Gas Council/Lurgi Slagger - Fischer-Tropsch Synthesis.

A plant capacity was selected based on the heating value of the methanol product. For an equivalent base point, 50,000 barrels per day of fuel oil having an assumed heating value of 6.3 million Btu's/bbl was selected. For the four methanol cases, the designs were based on producing exactly 315,000 MMBtu/day (HHV) of liquid methanol product. For Case 5, the same coal gasification section as in Case 2 was used to produce the same quantity of synthesis gas. The lower liquid yield of Fischer-Tropsch resulted in a lower liquid thermal output for the plant in

Case 5 although total vapor and liquid thermal yield is close to 315,000 MMBtu/day (HHV). Although it was preferred that all exported products would be in the form of liquids, it was recognized that two of the gasification processes as well as the Fischer-Tropsch process produce fuel gases in excess of the requirements for plant operation. It was agreed that a nominal quantity of export fuel gas would be acceptable. The plant designs were to be self sufficient in the production of all operating utilities including power.

In addition, it was agreed to use the 500 psig level for the design of the Chem Systems methanol system in view of the pressure level of the various gasification processes provided for this study.

Sources of data and other information are listed below that have been used in the preparation of this report. In the case of the gasifier designs, product distributions were supplied by the licensors in some instances as noted, including the required operating utilities; in other cases, heat and material balances and utility requirements were prepared by Parsons. In all cases for shift conversion and gas purification, all balances were prepared by Parsons. Information for liquid fuel synthesis was prepared as noted.

Data Source

1. Heat and Material Balance Data

   a. Gasifiers - Foster-Wheeler - supplied through EPRI by Foster-Wheeler

      British Gas Council-Lurgi Slagger - supplied through EPRI by Fluor.

      Texaco - supplied by Texaco supplemented with Parsons' computer program.

      Koppers-Totzek - Parsons' computer program.

Data Source (cont'd)

1. Heat and Material Balance Data (cont'd)

   b. Acid Gas Removal Units - Parsons in-house data based on Benfield's technology.

   c. Shift Section - Parsons' computer program.

   d. Cryogenic Section - data supplied by Union Carbide-Linde and incorporated into Parsons' computer program.

   e. Methanol Synthesis

      Chem Systems - supplied by Chem Systems

      ICI - Parsons' in-house data based on ICI's technology.

   f. Fischer-Tropsch

      In-house Parsons' and published Sasol's data used.

2. Cost Data For Screening-Type Estimate

   a. Coal preparation, gasification for Foster-Wheeler - supplied by EPRI.

   b. Coal preparation, gasification, tar separation, phenol recovery for British Gas Council-Lurgi Slagger - supplied by EPRI.

   c. Chem Systems methanol - Parsons' estimate based on Chem Systems' equipment sizing.

   d. Cryogenic Unit Cost - Union Carbide-Linde.

   e. All other estimates by Parsons based on curve-type estimating.

   f. Parsons' oxygen plant cost data, based on vendor's quotations, was used for all cases.

3. Coal Feed

   The coal analysis shown below was supplied to all licensors to predict product distribution from the gasifier and to set the requirements for oxygen and other utilities.

Coal Analysis

Type	Illinois No. 6
Proximate Analysis	**Weight Percent**
Moisture	4.2
Ash	9.6
Fixed Carbon	52.0
Volatile Matter	34.2
	100.00

Ultimate Analysis	Weight Percent	
	M.A.F.	As Received
Carbon	77.26	66.60
Hydrogen	5.92	5.10
Oxygen	11.14	9.60
Nitrogen	1.39	1.20
Sulfur	4.29	3.70
Ash	0	9.60
Moisture	0	4.20
	100.00	100.00

Heating Value - As Received

Higher Heating Value (HHV)	12,235 Btu/lb.
Lower Heating Value (LHV)	11,709 Btu/lb.

Heating Value - M.A.F.

Higher Heating Value (HHV)	14,194 Btu/lb.

SECTION 3

SUMMARY

A review of theoretical as well as practical considerations confirmed by a direct comparison of two cases showed that the yield of liquid fuel from coal is approximately 40 percent greater when producing methanol as compared to producing Fischer-Tropsch liquids, 56 percent of the heating value of the coal produced as methanol compared with 40 percent as Fischer-Tropsch liquid product. However, when the heating value of co-product fuel gas is added to that of the F-T liquids, the thermal recovery of heating value in fuel products in the two concepts is approximately the same at just over 56 percent of the heating value of the coal.

When producing methanol, the cost differences between gasification schemes developed in this study are small and are within the accuracy of the plant estimates with one possible exception. There are added costs for gasification at near atmospheric pressure when the product synthesis gas must be compressed for conversion to liquid fuels. This disadvantage results in both a marginally higher capital cost and a loss in thermal efficiency for the Koppers-Totzek system.

An earlier report* prepared by Parsons for EPRI, showed some potential economic advantage of the Chem Systems methanol process over others (ICI, Lurgi) and thus was used in association with the four gasification systems in this study.

(*Parsons' Job 5457-1, EPRI RP411 - detailed results unpublished because of the confidential nature of the data - covered by secrecy agreements).

Comparing the production of methanol with that of F-T liquid fuels, the following advantages favor methanol:

1. Lower capital cost.
2. Lower production cost per million Btu's of product fuel.
3. Produces only a liquid product with approximately the same heating value as the combined liquid and gas products from F-T.
4. Produces a single liquid product with a consistent product composition.

Differences in production costs are mainly a reflection of the differences in capital cost. Summaries of capital and production costs are given in Table 1 along with overall thermal efficiencies for the five cases based both on higher and lower heating values.

Within the accuracy of the plant estimates, it is reasonable to conclude that a plant for methanol production would be less costly than a plant of equivalent product heating value produced by Fischer-Tropsch processing. Comparing plant costs for the two products when using the same gasification system (Case 2 and Case 5), shows that the methanol plant is approximately 10 percent lower in capital. This difference is due entirely to the higher costs of the synthesis section of a plant to produce Fischer-Tropsch liquids.

TABLE 1

SUMMARY OF CAPITAL COSTS, PRODUCTION COSTS AND EFFICIENCIES

Note: Capital cost in millions of dollars, early 1977 basis:

	Case 1	Case 2	Case 3	Case 4	Case 5
	Foster-Wheeler Gasification/ Methanol	British Gas Council/Lurgi Slagging Gasification/ Methanol	Koppers-Totzek Gasification/ Methanol	Texaco Gasification/ Methanol	British Gas Council/Lurgi Slagging Gasification/ Fischer-Tropsch Synthesis
Total Plant Cost, Including Engineering	1,091	1,000	1,495	1,226	1,117
Contingency	164	150	224	184	168
Interest on Construction Loan	279	255	382	313	286
Royalty Allowance	6	6	9	7	6
Startup and Working Capital	183	169	232	195	183
Total Capital Requirement*	1,723	1,580	2,342	1,925	1,760
Production Cost: First Year - $/MMBtu (HHV)	5.92	5.70	7.81	6.57	6.24
Levelized - $/MMBtu (HHV)	5.18	5.00	6.75	5.70	5.44
First Year $/MMBtu (LHV)	6.73	6.49	8.89	7.48	6.76
Levelized - $/MMBtu (LHV)	5.88	5.69	7.69	6.49	5.89
Overall Thermal Efficiency, %					
HHV	55.5	57.0	52.4	58.2	56.0
LHV	51.0	52.3	48.1	53.5	54.0

*Relative accuracy of cost estimates with consideration to source of data and status of development of the gasification processes are judged to be as follows, in order with greatest accuracy first: - Cases 3, 4, 2, 1. Case 2 - Methanol estimate is considered to be of better accuracy than Case 5 - Fischer-Tropsch.

Although the comparison between gasification processes are not entirely consistent, they should not materially affect the study conclusions. For example, Koppers-Totzek made an allowance for 95 percent carbon conversion in the gasifier; all other designs are based on close to 100 percent. An adjustment to 100% basis for K-T would result in approximately the same numerical efficiency as the other cases as well as a proportionate reduction in capital cost. In the case of the slagging gasifier, the licensor elected to recycle to extinction the oils and tars produced in gasification. This type of operation may be successfully demonstrated in a current test program but it would not be surprising if additional oxygen were required. In the Texaco design, a relatively high slurry concentration was assumed which may require adjustment for a final design. These assumptions which were made for preliminary designs should be recognized but do not warrant adjustment at this time.

SECTION 4

SYNTHESIS PROCESSES FOR LIQUID FUELS

The chemistry of liquid fuel production in the form of methanol or Fischer-Tropsch liquids is based on using synthesis gas. A ratio of $2H_2/1CO$ represents a near stoichiometric ratio in a syngas used for the production of this type of liquid fuel. The consistency of the feed gas relationship is shown in Table 2 for the production of a variety of products.

Representative equations for the production of F-T liquids shown in Table 2 are typified by the coproduction of water. Also included below is the equation for the production of methanol which does not have a coproduct. For comparison, the methanation reaction for the production of methane is also shown.

Table 2

Synthesis Reactions

methane	$CO + 3H_2 \longrightarrow CH_4 + H_2O$
butane	$4CO + 9H_2 \longrightarrow C_4H_{10} + 4H_2O$
hexane	$6CO + 13H_2 \longrightarrow C_6H_{14} + 6H_2O$
decane	$10CO + 21H_2 \longrightarrow C_{10}H_{22} + 10H_2O$
cyclohexane	$6CO + 12H_2 \longrightarrow C_6H_{12} + 6H_2O$
ethylene	$2CO + 4H_2 \longrightarrow C_2H_4 + 2H_2O$
methanol	$CO + 2H_2 \longrightarrow CH_3OH$

Based on these equations, the theoretical recovery of heat in product fuels as a percentage of the heating value of the stoichiometric synthesis gas is shown in Table 3 below. Percentage-wise, the recovery of heating value in product fuels is approximately the same for all hydrocarbon products. For methanol, the recovery is higher. However, the heat of reaction in all cases is recoverable in the form of steam production. When this recoverable energy is added to the fuel values shown in Table 3, the theoretical thermal yields are in excess of 95 percent as shown below for a selected number of typical products.

Table 3

Theoretical Percentage Recovery of Fuel Value From Syngas - HHV

	Product as Liquid	Product as Gas
decane	76.8	77.3
hexane	76.9	77.5
butane	77.1	77.7
methane	-	78.1
cyclohexane	76.5	77.1
ethylene	-	82.5
methanol	85.1	

Theoretical Percentage Recovery of Thermal Energy From Syngas - HHV

	Methanol	Butane	Cyclohexane	Methane
Liquid Fuel	85.1	77.1	76.5	78.1
Heat of Reaction	11.5	18.3	18.5	19.4
Total Thermal Yield	96.6	95.4	95.0	97.5

Recoverable Thermal Values. Recovery of the calculated thermal values based on theoretical conversions of syngas to fuel products is difficult to approach in plant operation. For the production of methanol as well as for Fischer-Tropsch liquids, high conversions of syngas to products is achieved only by the use of gas recycle due to limitations of catalyst performance. Gas recycle systems are limited by the buildup of inert gases contained in the makeup syngas. The restrictions imposed by recycle systems and inert gas buildup are the cause of differences in costs for the production of methanol and Fischer-Tropsch liquids and in the loss in thermal efficiency for both processes. Chem Systems methanol process is aimed at minimizing recycle energy requirements.

In methanol synthesis, the conversion of syngas per pass through a reactor is relatively low due to temperature limitations on the catalyst. The exothermic heat released in the reaction is largely absorbed in a temperature rise across the catalyst bed. A recycle gas system is employed to dilute the reactants and to assist in absorbing a portion of the reaction heat. To obtain a high conversion, unconverted syngas is recycled back to the reactor inlet after removal of product methanol by condensation.

Complete recycle so as to obtain 100% conversion is not possible because the feed gas contains "inerts" (N_2, A, CH_4) in addition to the reactants (H_2, CO) and so it is necessary to "purge" the synthesis loop to maintain pressure balance by removing the inerts to balance inerts fed. This results in some loss of reactants (H_2, CO) with the purge. It is the purpose of process design and development to optimize the several operating factors. The main difference between methanol processes is in reactor design, heat recovery, recycle schemes and selected operating pressures and temperatures.

Analysis of an actual process design from a previous study gives the following results for a typical Chem Systems' methanol loop (1100 psi synthesis; based on a feed gas with 1.1% inerts and stoichiometric H_2/CO ratio).

Percent of heating value of feed gas recovered as:

Methanol product	81.5
Off gas	4.2
Subtotal	85.7
Reaction heat recovered	12.8
Total Recovered	98.5
Loss to cooler	1.6
Total	100.1

The loop design is seen to approach the ideal quite closely; the conversion of $H_2 + CO$ is over 97% and the total heat recovery is high. Not shown above, however, is the recycle energy required. The above loop produces methanol containing certain condensible by-products, the thermal value of which are included in the values shown. A typical complete analysis supplied by Chem Systems' is:

methanol	95.4 wt %
ethanol	1.0 wt %
isopropanol	1.0 wt %
higher alcohols	0.1 wt %
water	2.5 wt %

The methanol synthesis produces no CH_4 and hence the purge gas is taken only to balance "inerts" contained in the feed gas.

In Fischer-Tropsch synthesis, the same considerations as above apply with regard to conversion per pass, need for recycle of unconverted product and purging of inerts. Other factors complicate the design, however. The Fischer-Tropsch produces a wide range of products from methane and ethylene through gasoline and fuel oil ranging up to waxes.

The only detailed operating experience of practical interest is that of Sasol (South Africa) which runs two types of Fischer-Tropsch systems: fixed bed and fluidized bed. Typical conversions and yields are:

	Fixed Bed	Fluidized Bed
Percent net conversion of H_2 + CO	65	85
H_2/CO ratio feed gas	1.7	2.8
Wt % distribution of products		
C_1	7.8	13.1
C_2	3.2	10.2
C_3	6.1	16.2
C_4	4.9	13.2
C_5-C_{11}	24.8	33.4
C_{12}-C_{20}	14.7	5.1
$> C_{20}$	36.2	-
alcohols, ketones	2.3	7.8
acids	-	1.0
	100	100

The fixed bed (450°F, 370 psig) produces a heavier spectrum of products than the fluidized bed (625°F, 320 psig); the latter requires a higher than stoichiometric H_2/CO ratio (2.8 to 1 rather than 2 to 1) to maintain continuous operation of the fluidization process.

However, because of the formation of light gases in F-T synthesis and the high inerts in the feed gas (over 10% in the purified gas from the Lurgi process as an example) and because of the non-stoichiometric feed, the amount of purge must be increased and the overall conversion of $H_2 + CO$ is thereby limited. Typically, as shown, only 65-85% of the $H_2 + CO$ fed is converted and the thermal recovery in liquid products relative to the feed is as follows:

% Liquid Fuel Value	=	Theoretical Thermal Yield @ 100% Conversion	x	Fraction Feed Converted	x	Fraction of Products Heavier Than C_5	
Fixed Bed F-T		77	x	.65	x	.829	= 41%
Fluidized Bed F-T		77	x	.85	x	.605	= 40%

Assuming it were possible to obtain a conversion of 95% with a low-inert feed gas at near stoichiometric composition, the thermal recovery as liquid products would be about 77 x .95 x .829 = 60%. This is still considerably less than the 81% obtained with methanol.

In the original design of Sasol, the tail gas from the two F-T units was fed to a reforming unit (partial oxidation type) to reform the CH_4 back to H_2 and CO. This reforming step was necessary in order to increase the conversion of synthesis gas to liquid products. The reforming step introduces another inefficiency in the process that should be minimized. At Sasol, a partial solution has been to co-produce a fuel gas which is distributed to users in the area. This gas has a heating value of approximately 500 Btu/scf which could be upgraded to SNG (980 Btu/scf) by methanation and CO_2 removal.

Alternately, the tail gas could be used as boiler fuel for plant utility steam raising or power generation. At Sasol, it is preferred to sell fuel gas and to use coal for the steam/power plant because this gives better overall efficiency.

The Fischer-Tropsch process has considerable flexibility to alter product distribution by selecting different process conditions but invariably a wide boiling range of products is produced. For Sasol, all the products are useful. While the medium boiling oil cuts and part of the off gas are used as fuels, the rest of the product spectrum has numerous miscellaneous uses in the petroleum and chemical industry - in such products as gasoline, various types of waxes, solvents, varnishes and plastics. These are produced in an extensive refinery and by-product plant. The above considerations suggest that a F-T plant producing all liquid fuels has relatively low efficiency due to excessive thermal loss in re-reforming the CH_4 and other light products. A preferred scheme for fuels would appear to be to produce part liquid products and part SNG.

Although a detailed process design is not available for a F-T plant producing all liquids to enable a precise comparison to be made with the methanol process, the above considerations indicate that the thermal efficiency of producing methanol from coal will be higher than for producing F-T liquid fuels. This is based on theoretical and practical considerations. Also, Sasol have given results* of a comparative study producing methanol from natural gas at 58% efficiency compared with producing F-T liquids from natural gas at 49%. The efficiency of producing methanol using coal as feed is about 50-60%; similar lower efficiency is expected for F-T from coal since the front-end gasification and purification are essentially the same for each giving F-T efficiency of 42-50%. Lurgi** has suggested the wider range of difference, 58% for methanol from coal versus 42% for Fischer-Tropsch. It is important to note that F-T efficiency is higher than 42% if a gas co-product is permitted.

Product Properties

Both the methanol product and F-T synthesis liquids are sulfur and nitrogen free.

The methanol product is a narrow boiling range liquid with only small quantities of other materials - see analysis on Page 14.

The F-T products cover a wide boiling range from light LPG-type liquids up to waxes. The use of F-T products as fuels opens the possibility of selecting various blends and boiling range fractions.

Methanol has a lower heating value per pound and per gallon and hence product storage and pumps would be larger.

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*"New applications of the Fischer-Tropsch Process". J. C. Hoogendoorn, SASOL, 1975, Page 11.

**"Lurgi Information" 1/76, Page 28.

Higher Heating Values	Pure Methanol	Chem System Fuel Methanol	Heptane	Fischer-Tropsch C4+ Product
Btu/lb.	9,760	9,610	20,680	20,055
Btu/gallon	64,770	63,930	118,670	124,200
MMBtu/ton	19.5	19.22	41.4	40.1
MMBtu/barrel	2.72	2.68	4.98	5.22

Equipment

For this study, the plant capacity is based on producing 315,000 MMBtu/day of fuel product. This is equivalent to a production capacity of 50,000 bbl/day of fuel oil equivalent assuming 6.3 MMBtu/bbl of fuel oil. To produce 315,000 MMBtu/day HHV of liquid fuel, it would be necessary to produce about 16,400 TPD of methanol. This represents about 10-14 synthesis loops of the largest in operation (1200-1600 TPD) although designs have been proposed for up to 5000 TPD in a single loop. For Fischer-Tropsch, if the thermal output is all produced as a liquid, the product requirement would be about 63,000 BPD. At Sasol, the fixed bed reactors each produce about 550 BPD; so over 100 of these units at that commercial size would be required. According to Sasol, it is impractical to scale-up the present reactors to larger capacities so a large capacity plant would need a rather large number of F-T units. Unfortunately, this is the type of reactor which produces the heavier products preferred for fuel use. It is interesting to note that in the 1930's the capacity per unit was even smaller and some of the units in Germany had 40-70 reactors in 2 or 3 stages to get reasonable conversions.

The other type of reactor in operation at Sasol is the fluidized bed type, the original design capacity per reactor being 2,000 BPD. Recent information by Sasol indicates that the expansion now in the design stage ("Sasol II") will use fluidized bed reactors of capacity 5,000 BPD (original reactors x 2.5). For this newer type, about 12 would be required for the present study. However, this type of reactor is more suited to produce lighter products - Sasol's main objective is to produce gasoline rather than fuel oil.

The space velocities used on F-T and methanol are comparable, around 500-600 scf feed gas/cf catalyst volume. However, methanol requires a higher pressure level for optimum operation from 500-1100 psi (Chem Systems), 700-1000 psi (Lurgi), up to 1100-1500 psi (ICI). Levels of heat recovery are similar, approximately 250 psi steam can be raised indirectly (Chem Systems) whereas at Sasol the steam levels are 246 psi fixed bed, 175 psi fluidized bed. For methanol, the latest designs by ICI which include heating 1500 psi boiler feedwater result in a preferred overall economy for their process; in Lurgi's case up to 570 psi steam is raised directly in the "boiling-water" reactor.

F-T process requires more exotic metallurgy - protective lining is required to prevent erosion in the fluidized bed system and for protection against by-product acids which are condensed in the effluent cooling system.

Methanol Processes

Methanol processes are offered by several companies: Vulcan-Cincinnati, ICI, Lurgi, Mitsubishi, Nissui-Topsoe, and Selas-Polimex. In recent years, contract awards have been dominated by ICI and Lurgi. Recent methanol plant construction has been based on copper catalysts in the "so-called" low pressure methanol process. ICI have about 14 plants which have been started up since 1970 and Lurgi have at least 8 in operation or under construction. The ICI process uses a quench-type reactor and Lurgi uses a heat-exchanger type reactor with boiling water in the jacket and catalyst in the tubes. Significantly, Lurgi's recent success has been, at least partly, based on claims for higher process efficiency due to this "isothermal" steam-recovery type of reactor. Licensees of ICI have countered with various designs, some involving improved heat recovery from the methanol loop.

Methanol production is a highly developed, competitive field with several proven processes as indicated above. A recent development in this field, supported by EPRI is the Chem Systems' Liquid Phase Methanol Synthesis. This uses an inert liquid to absorb the heat of reaction and thereby reduce the gas recycle substantially.

A previous study by Parsons for EPRI* compared the Chem Systems' methanol process with ICI's for synthesis at 1100 psi. The overall methanol process is similar for both ICI and Chem Systems with the exception of the methanol synthesis loop. Since the ICI process, as well as Lurgi's are highly developed commercial processes of good efficiency, room for further improvement is relatively small.

However, the above study indicated that the Chem Systems' process, although commercially unproven, did show potentially somewhat higher thermal efficiency and lower capital cost than the ICI system.

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*Parsons' Job 5457-1, EPRI RP411 - detailed results unpublished because of the confidential nature of the data - covered by secrecy agreements.

Therefore, it was proposed for this present study to select Chem Systems for the methanol process. Chem Systems had suggested that the process may show better economics at 500 psi rather than 1100 psi synthesis. A brief separate study was made to evaluate this suggestion. Comparative capital costs for the Chem Systems' loop at 500 psi and 1100 psi were made based on data provided for the previous study. Assuming the upstream plant producing syngas is at the 500 psi level, a syngas compressor from 500 psi to 1100 psi was added to the 1100 psi loop. It is found that the heat recovery of the loop is slightly poorer at the lower pressure, however, total energy required for feed and recycle compression is lower. Capital cost is considerably higher for the lower pressure as shown in Table 4.

TABLE 4

COMPARISON OF 1100 PSI AND 500 PSI CHEM SYSTEMS METHANOL SYNTHESIS

	Chem Systems 1100 psi Synthesis	Chem Systems 500 psi Synthesis
Recovery as a percentage of the thermal value of feed gas to loop as:		
Methanol	81.5	80.8
Purge Gas	4.2	5.1
	85.7	85.9
Waste Heat Boiler	11.0	10.9
Boiler Feedwater	1.8	1.2
Total	98.5	98.0
Loss to Cooler	1.6	2.1
	100.1	100.1
Capital Cost for Nominal 1300 TPD loop	9.0	15.0
Feed Gas Compressor	1.7	-
Total	10.7 MM $	15.0 MM $
Power Consumption		
Make-up Compressor	6,150	-
Recycle Compressor	510	2,020
Oil Pump	440	990
	7,100 HP	3,010 HP

Savings in HP for lower pressure loop

= 4,090 HP = 3,050 KW

(@ 10,000 Btu/kwh = 30.5 MMBtu/hr.)

Extra equipment cost: 4.3 MM $

Effect on production cost $/MMBtu: -

Extra cost due to 4.3 MM $ capital	+ 0.12 $/MMBtu
Reduction due to increased overall efficiency	- 0.12 $/MMBtu
Net effect	0

This brief study indicates a stand-off between the extra capital cost and the energy savings for Chem Systems loops at 500 psi and 1100 psi with gas supplied at 500 psi.

SECTION 5

GASIFICATION PROCESSES

The four gasification processes under study cover a wide range of operating conditions and gasifier types. Briefly the processes are as follows:

1. Foster-Wheeler (F-W)

   Pulverized coal is introduced to a lock-hopper system and then injected into the gasifier second stage using transport gas (portion of product gas recycled). The second stage gasifier is an entrained bed type operating at 300-400 psig with an exit temperature of 1700°F. The char carried out with the gas is separated in the char separator and passed to drums from which it is fluidized with steam and fed to the first stage gasifier. Oxygen is fed to the first stage gasifier and the gas passes directly to the second stage located above the first. Slag from the first stage gasifier is removed via a quench and lock-hopper system.

   The F-W process is in a relatively early stage of pilot plant development.

2. Slagging-Gasifier (British Gas Council-Lurgi) (BGC-Slagger)

   Sized coal is introduced through a coal lock-hopper system into the top of the gasifier via a distributor system. The coal descends slowly in the gasifier bed ("moving bed") and is successively dried, devolatized and gasified. Oxygen and steam are fed at the bottom and slag is withdrawn to a quench chamber and a slag lock-hopper. The gasifier operates at 300-350 psig with temperatures over 2000°F in the ash zone and outlet gas temperature of 800-1100°F. Outlet gas is quenched by a circulating liquor stream.

The slagging-gasifier has a demonstration unit in operation at Westfield, Scotland.

3. Koppers-Totzek (K-T)

Pulverized coal is injected with steam and oxygen horizontally into the gasifier operating essentially at atmospheric pressure. The feed ratios are such that the gasification temperature is around 2700°F. Molten slag falls to a slag quench tank and gas passes to a waste heat boiler and thence to gas cleanup system. A single gasifier may have several "heads" injecting the coal-steam-oxygen mixture, these "heads" being arranged directly opposite so that flames from opposite "burners" impinge.

The K-T process is commercially developed with many gasifiers in operation and several under construction.

4. Texaco

Pulverized coal is slurried in water and pumped into a high pressure (600-700 psig) gasifier together with oxygen.

The mixture enters the top and is fired downwards. Most of the gas at 2500°F is withdrawn through a side nozzle to the waste heat boiler system and thence to the gas cleanup. Part of the gas passes to a lower quench chamber. The slag is removed through a slag hopper beneath the quench chamber.

The Texaco process has undergone extensive pilot plant work at Montebello, California and a demonstration unit is under construction in Germany.

Composition of Gasifier Product Gas

The produced syngas compositions are given in Table 5 for the four gasification systems.

The gas compositions reflect the different operating temperatures used in the different gasification systems. The gas from the BGC-Slagger at 800-1000°F is directly from the coal devolatizing zone and, in addition to the gasification products from the steam-O_2 zone i.e. H_2, CO and CO_2, it contains a spectrum of hydrocarbons from methane through naphtha, oils and phenols up to tars. In the data presented by EPRI, the N+O+T+P (naphtha, oil, tar, phenol) is recovered and recycled to extinction in the gasifier.

The F-W gas is at a higher temperature (1700°F) and is produced from a devolatizing zone. In the data presented by EPRI, no hydrocarbons heavier than methane are produced although the yield of methane is similar to the BGC-Slagger. The next highest gas temperature (2500°F) is produced in the Texaco reactor. The gas contains a low concentration of methane and no heavier hydrocarbons. Finally, the highest temperature (2700°F) gas is from the K-T unit which shows no higher hydrocarbons and an essentially negligible methane content.

All the sulfur in the coal appears as H_2S and COS in the gas, different ratios of H_2S/COS reflect the different operating conditions used in the four gasification processes.

In the case of K-T and Texaco, all the N_2 in the coal as well as from the oxygen appear in the gas. In the case of F-W and BGC-Slagger, partial conversion to NH_3 occurs and this ammonia is removed and does not appear in a purified gas.

The ratios of H_2/CO and CO/CO_2 depend on the amount of steam utilized and process conditions. Highest contents of CO correspond with the lowest steam users, K-T and BGC-Slagger.

TABLE 5

COMPOSITION OF GASIFIER PRODUCT GAS**

(Coal Feed - Illinois No. 6)

Mole % (Dry Basis)	Foster-Wheeler		BGC-Slagger		Koppers-Totzek		Texaco*	
	Generator Gas	After Acid Gas Removal	Generator Gas	After Acid Gas Removal	Generator Gas	After Acid Gas Removal	Generator Gas	After Acid Gas Removal
CH_4	7.02	8.40	7.65	7.99	--	--	0.10	0.11
C_2H_4	--	--	0.23	0.24	--	--	--	--
C_2H_6	--	--	0.33	0.34	--	--	--	--
H_2	35.16	42.05	29.96	31.28	35.20	38.68	35.06	39.58
CO	40.93	48.95	57.18	59.70	54.71	60.12	52.35	59.10
CO_2	14.38	Nil	1.91	Nil	7.63	Nil	10.12	Nil
H_2S	1.14	Nil	1.36	Nil	1.23	Nil	1.22	Nil
COS	0.12	2 ppm	0.06	2 ppm	0.14	2 ppm	0.07	2 ppm
N_2	0.51	0.61	0.44	0.46	1.09	1.20	1.07	1.22
NH_3	0.74	--	0.89	--	--	--	--	--
Total	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
H_2/CO	--	0.86	--	0.52	--	0.64	--	0.67

*After adjustment of oxygen feed purity from Texaco data.

**All based on 98% O_2, 2% N_2.

The preferred synthesis gas for methanol production would have low "inerts" (CH_4+N_2+A). In this respect, the Texaco and K-T gases are advantageous. However, the high "inert" processes, F-W and BGC-Slagger, have higher gasification efficiency overall and therefore must also be considered.

Oxygen Usage in Gasification

The oxygen usages are as follows:

	F-W	BGC-Slagger	K-T	Texaco
O_2 ton/ton coal (as received)	0.58	0.46	0.76	0.82
O_2 ton/ton MAF coal	0.68	0.53	0.88	0.95
O_2 purity	98%	98%	98%	98%

The oxygen usage for the K-T and Texaco process are higher than for F-W and BGC-Slagger. This is due to more complete conversion to CO and H_2 (less CH_4, etc.) caused by operating at a higher temperature and in the case of Texaco due to use of water-slurry.

Steam Used and Steam Recovery in Gasification

The steam used and recovered are approximately as follows:

	F-W	BGC-Slagger	K-T	Texaco
HP steam recovered, T/T coal*	1.50	-	1.79	1.90
Jacket steam recovered, T/T coal	-	0.14	0.13	-
Steam used, T/T coal	0.60 (process)	0.30 (process)	0.27 (process)	0.20 (preheat)
Net steam export T/T	0.90	-	1.65	1.70
Net steam import T/T	-	0.16	-	-

*Based on equivalent saturated steam at 1500 psi.

All processes except BGC-Slagger recover more steam than used in operation. The high oxygen, high temperature processes K-T and Texaco, recover more heat from the effluent gas. All processes use effluent boilers except BGC-Slagger which quenches the effluent gas due to its relatively low temperature.

Gas Production Pressure

The pressures of the gasifiers are approximately:

	F-W	BGC-Slagger	K-T	Texaco
psig	360	350	6	650

Texaco is the only process that does not require synthesis gas compression for liquid fuel synthesis at 500 psi. Koppers-Totzek requires large compressors.

Conversions Obtained

	F-W	BGC-Slagger	K-T	Texaco
Specified %C converted	99.3	99.7	95.0	~ 100

Although Koppers-Totzek show lowest carbon conversion, the process data is based on commercial practice. Other conversions may be optimistic.

Gasification Efficiencies

As a means of illustrating differences between gasification processes, the following efficiencies are calculated, all on a HHV basis:

"Cold Gas" efficiency, (synthesis gas) $$E_1 = \frac{\text{Heating value of } H_2 \text{ plus CO produced}}{\text{Heating value of coal}}$$

"Cold Gas" efficiency, (fuels gas) $$E_2 = \frac{\text{Heating value of } H_2\text{, CO and } C_1{+}C_2 \text{ produced}}{\text{Heating value of coal}}$$

"Process" efficiency, (fuel gas plus steam) $$E_3 = \frac{\text{Heating value of } H_2\text{, CO, } C_1{+}C_2 \pm \text{ net steam produced or required}}{\text{Heating value of coal}}$$

Results are:

	F-W	BGC-Slagger	K-T	Texaco
"Cold Gas" efficiency (H_2+CO), E_1	66.7	70.8	75.7	77.3
"Cold Gas" efficiency (H_2+CO+C_1+C_2), E_2	85.9	92.7	75.7	77.6
"Process" efficiency (including steam), E_3	93.3	91.4	89.2	91.6

The K-T and Texaco gasifiers show higher conversion to H_2+CO,(E_1), which is required for making synthesis gas. However, when the methane and ethane and ethylene are also included, (E_2), the F-W and BGC-Slagger show considerably higher efficiency. This reflects the lower oxygen usage but this also results in the recovery of less heat by steam generation. When heat recovery is also included, (E_3), all processes are more comparable. Note that K-T would be higher if the carbon conversion were assumed closer to 100% like the other three processes rather than 95%.

If an equivalent of power usage is included, K-T is penalized most because of high gas compression and high O_2 usage and shows the lowest efficiency; Texaco is next lowest being penalized by high O_2 consumption, with F-W next and BGC-Slagger showing the highest "cold gas" efficiency.

These efficiencies are, of course, only gasification efficiencies truly applicable only for fuel gas production. For synthesis gas production, to obtain overall efficiency, the complete plant must be designed and fuel and utility balanced. This is done in the individual case design sections that follow.

The gas analyses in Table 5 were submitted to Chem Systems for design of the methanol synthesis section.

Adjustment of Gas Composition for Methanol Synthesis

Chem Systems reduced the four gasifier cases to a single case at the 500 psi level and required that the methanol loop be fed with synthesis gas of the following composition.

	% mol	
CO	32.90	
H_2	65.83	
CO_2	0.17	
CH_4	0.62	1.11% "inerts"
N_2+A	0.49	
	100.00	

The H_2/CO ratio is 2.0, the stoichiometric ratio for methanol synthesis; the small quantity of CO_2 is said to be required for catalyst activity and is shown in the mass balance to pass unconverted into the purge gas along with the "inerts" CH_4 and N_2+A.

For Cases 3 and 4, Koppers-Totzek and Texaco respectively, this gas analysis is matched closely after partial shift and CO_2 removal to give the following synthesis gas composition.

	Case 3 K-T		Case 4 Texaco	
CO	32.87		32.82	
H_2	65.77		65.68	
CO_2	0.17		0.17	
CH_4	-	1.19	0.11	1.32
N_2+A	1.19		1.21	
	100.00		100.00	

In Cases 1 and 2, Foster-Wheeler and British Gas Council/Lurgi Slagger respectively, the gas after partial shift and CO_2 removal still contains substantial quantities of methane (about 8.4%), far in excess of the 0.62% specified. A cryogenic unit is included in Cases 1 and 2 to remove the surplus CH_4 and heavier hydrocarbons. Linde (Union Carbide) advises that the following typical results for a cold box in this service can be obtained.

H_2 recovery to syngas	100% approx.
CO recovery to syngas	90% approx.
CH_4 level at outlet	0.6% (94% removal) approx.

Overall, the cryogenic unit splits the feed into a methanol synthesis gas at 500 psi and a tail gas consisting of CH_4 and some CO at low pressure which is used as fuel gas. For this service, the cold box requires a feed pressure of 700 psi to provide the necessary refrigeration and also requires recompression of a CO-rich portion.

Since all CO_2 is removed ahead of the cold box for cryogenic separation, a small quantity of CO_2 is added to the synthesis gas ahead of the synthesis loop.

The compositions obtained for synthesis gas after partial shift, CO_2 removal and cryogenic separation for Case 1, Foster-Wheeler and Case 2, British Gas Council-Lurgi Slagger matches the above analysis used by Chem Systems.

SECTION 6

CASE 1

FOSTER-WHEELER GASIFICATION - CHEM SYSTEMS METHANOL

The overall plant is shown on Process Block Flow Diagram 5604-FS-1-A. A total of 24,566 tons/day of coal are used, all being gasified. Fuel gas recovered from process is used in the utility plant and excess fuel gas exported as a product. Plant thermal efficiency is as follows:

Input

Coal

24,566 TPD x 2,000 lb/T x 12,235 Btu/lb. = 601,130 MMBtu/day (HHV)

Output

Methanol

16,392 TPD x 2,000 lb/T x 9,610 Btu/lb. = 315,000 MMBtu/day (HHV)

Excess fuel gas

23.05 MMSCFD x 800.2 Btu/scf = 18,444 MMBtu/day (HHV)

Total 333,444 MMBtu/day (HHV)

$$\text{Efficiency} = \frac{333,444}{601,130} \times 100 = \underline{\underline{55.5\%}} \text{ (HHV)}$$

$$\text{Efficiency} = \frac{276,700 + 16,812}{575,286} \times 100 = \underline{\underline{51.0\%}} \text{ (LHV)}$$

Coal Preparation

Coal (1-1/2" x 0) is unloaded and conveyed to the storage stack. Reclaimed coal is conveyed to the pulverizer feed hopper. Coal is pulverized in a system using nitrogen as the conveying medium.

Pulverized coal is then air conveyed to the gasifier feed surge bin.

Gasification

Refer to Process Flow Diagram 5604-FS-1-B, Gasification, Case 1, Foster-Wheeler.

Pulverized coal from the feed surge bin is fed to one of the coal feed hoppers which is then pressurized with nitrogen and fed to the high pressure coal feed hopper. The coal is then entrained in the transport gas, which is a portion of the product gas recycled and fed to the gasifier Stage II. Here, the coal is gasified in the presence of hot gas from the gasifier Stage I. Gas carrying entrained char flows to the char separator where char is removed and then fed by a lock system and using steam as fluidizing and conveying medium into the gasifier - Stage I. Here, the char reacts with steam and oxygen as the gas flows to the second stage as described above. Slag from the second stage is removed via a slag quench and lock-hopper system. A circulating water system cools the slag and conveys it to a settler where ash/slag is removed for disposal. The water is recycled via a cooling tower.

Hot gas at 1700°F from the char separator is cooled while generating high pressure steam and then scrubbed in a water stream for final particulate matter removal. These solids are returned to the gasifier in Stage I. The gas is further cooled and ammonia removed in a water wash tower.

Acid Gas Removal

Refer to Process Flow Diagram 5604-FS-1-C, Acid Gas Removal.

The Benfield Hi-Pure System is used for removal of H_2S and CO_2. The gas is contacted by lean carbonate solution and then by lean amine solution in separate sections of the absorber. The rich carbonate solution is regenerated in a stripper at low pressure and the rich amine solution is similarily regenerated in a regenerator using steam heated reboilers. CO_2 and H_2S, the "acid" gas from the regenerator, passes to the sulfur recovery unit.

CO Shift

Refer to Process Flow Diagram 5604-FS-1-D, CO Shift Conversion.

Part of the gas is passed to a two-stage CO shift section and part is bypassed. Of the feed to the shift, part of the gas is blended with all the steam and fed to the first stage and the rest of the gas is blended to the second stage feed. In the shift section, the H_2/CO ratio is increased from the gasifier level up to that required for methanol production by the reaction.

$$CO + H_2O = H_2 + CO_2$$

The reaction is exothermic and heat is recovered as high pressure steam. From the second stage, the gas is cooled to remove excess steam as condensate and then passes to the CO_2 removal section.

CO_2 Removal

Refer to Process Flow Diagram 5604-FS-1-E, CO_2 Removal.

The Benfield process is used for CO_2 removal from the gas from the shift section. The gas is cooled by passing through the reboiler where the heat of condensation is used to vaporize steam used in the regenerator.

The gas is contacted in the absorber by lean carbonate solution and CO_2 is removed by the reaction:

$$K_2CO_3 + CO_2 + H_2O = 2KHCO_3$$

The "lean" solution is rich in carbonate and low in bicarbonate while the "rich" solution is mostly converted to bicarbonate. This "rich" solution is regenerated in the stripper. By the combined affect of low pressure and stripping steam, CO_2 is removed and carbonate formed, reversing the above reaction. The lean solution is returned to the absorber. CO_2 is vented to atmosphere.

Compression

Gas from shift and CO_2 removal is blended with the gas that has bypassed those sections to produce a synthesis gas with a H_2/CO ratio close to 2 which is compressed from 225 psig to 700 psig.

CO_2, H_2O Removal

Final traces of CO_2 and H_2O are removed in a standard cryogenic feed purification unit which may be caustic scrubbing followed by alumina driers or preferably molecular sieves.

Cryogenic Unit

The dry, CO_2-free gas is fed to the cold box where a separation is made into synthesis gas (H_2/CO ratio = 2, CH_4+N_2 about 1%) and a tail gas which is used as fuel gas and consists mainly of methane with some CO.

Some CO_2 is blended back into the synthesis gas to attain the specified 0.17% CO_2 content required by methanol synthesis.

The gas at 500 psi is now ready for methanol synthesis.

Methanol Synthesis

Refer to Process Flow Diagram 5604-FS-1-F, Methanol Synthesis.

Synthesis gas feed is heated in exchange with reactor effluent gas and passed through a zinc oxide bed for final trace H_2S removal. The gas enters the ebullating bed catalytic reactor together with the recycle gas and recycle oil. Conversion of $2H_2+CO$ to methanol is an exothermic reaction and the temperature rises 15°F from 431°F inlet (start-of-run) or 467°F inlet (end-of-run). At the top of the reactor, a phase separation is made and the oil is separated. It passes to the oil surge drum and is recirculated to the reactor via the waste heat boilers where in cooling back 15°F it raises steam at approximately 300 psi.

The vapor overhead from the reactor is cooled in successive stages. First, it is split and the two streams used to heat the feed gas and the recycle. Then the gases are recombined and a final cooler cools the gas and condenses the methanol. In a separator at 120°F, the fuel methanol is withdrawn as a product to storage.

The gas is recycled via a compressor with a portion purged from the system to withdraw inerts. This purged gas is used as fuel gas elsewhere in the methanol complex.

Also from the separator, an oil stream is skimmed and recyled to the reactors.

Approximate Arrangement

The plant consists of pulverizers feeding 8 Foster-Wheeler gasifiers. The gas is handled by four main trains of acid gas removal with some multi-units (for example absorbers), 4 shift trains and 4 trains of CO_2 removal. Then follows compression and several trains of molecular sieve and cryogenic units for synthesis gas preparation.

Finally, 7 trains of methanol synthesis are required all with multiple reactors. Two sulfur plants are included.

Utility System

The plant is balanced so that the steam generated is sufficient for use in the steam turbine for plant compressors, power generation and other steam requirements. A total of 5.7×10^6 lb/hr. of steam at 1500 psig, 950°F is produced.

About 54% is produced (saturated) in the Foster-Wheeler gasification section, about 6% is produced (saturated) in the shift section and the balance 40% is produced in the utility boiler/superheater system which also superheats the total steam.

The HP steam is used in back pressure and condensing turbines to drive the main plant turbines and provide process steam at 420 psig and 50 psig levels.

The large turbines are:

	HP
Synthesis Gas and Cryogenic Compressors	260,000
Oxygen Plant Compressors	359,000
Other Compressors (methanol; transport gas)	27,000
Power Generator	72,000
Total	718,000

Steam from methanol synthesis 1.6×10^6 lb/hr., is produced at 285 psig, saturrated. Part is used as shift process steam and part is superheated and used for power generation (25,000 HP) before final use at 50 psig in the acid gas removal reboilers and miscellaneous users. Total power generated is 97,000 HP.

Condensate is recovered from vacuum condensers, steam reboilers and shift condensate and returned to the deaerator.

Make-up boiler water required is 1.8×10^6 lb/hr. corresponding to the steam converted in gasification, shift sections and losses.

CAPITAL SUMMARY - CASE 1

	MM $
Coal Preparation and Grinding	60
Gasification	134
Acid Gas Removal	114
CO Shift	30
CO_2 Removal	88
Oxygen	250
Cryogenic Purification and Cold Boxes	21
Methanol Synthesis	122
Final Desulfurization - Zinc Oxide	10
Raw Gas Compression	51
Sulfur Recovery	20
Utility Boilers, Superheaters and Power Generators	49
	949
Utilities and General Offsites at 15%	142
Total Plant Cost, Including Engineering	1,091
Contingency at 15%	164
Total Plant Investment	1,255
Interest on Construction Loan	279
Royalty Allowance	6
Startup and Working Capital	183
Total Capital Requirement	1,723 MM $

PRODUCTION COST* - CASE 1

Illinois Coal	-	25$/ton at 12,235 Btu/lb. HHV (1.02 $ per MM Btu)
Coal	-	24,566 TPD
Overall Efficiency	-	55.5% HHV
Liquid Methanol Product	-	315,000 MM Btu/day HHV
Fuel Gas Product	-	18,444 MM Btu/day HHV
Total Product	-	333,444 MM Btu/day HHV
Total Capital Requirement	-	1,723 million dollars
Operating Factor	-	90%, 328.5 days/year

	First Year Cost $/MMBtu	Levelized Cost $/MMBtu
Coal	1.84	1.84
Water, Catalyst & Chemicals	0.07	0.07
Other Operating Costs	0.82	0.82
Total Operating Cost	2.73	2.73
Capital Charges	3.19	2.45
Production Cost	5.92 HHV	5.18 HHV
	6.73 LHV	5.88 LHV

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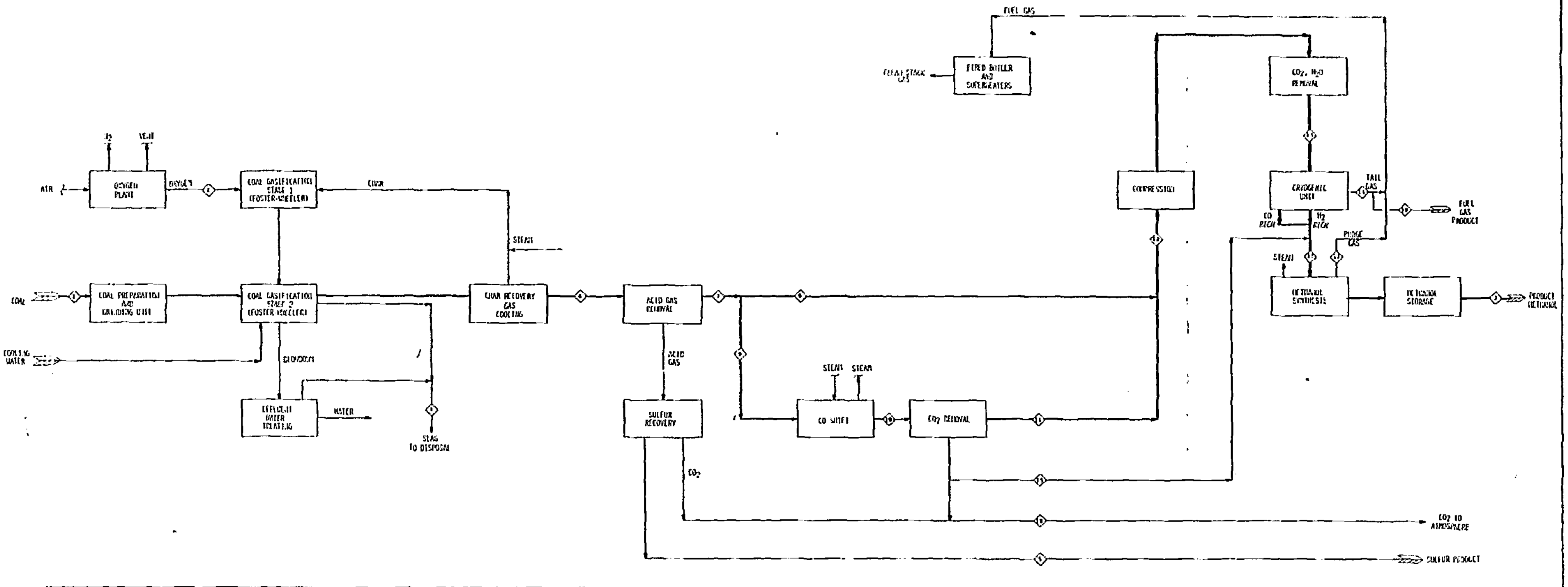
*Calculated in accordance with EPRI's Economic Premises, Feb. 11, 1977-utility financing.



	1	2	3	4	5
	FEED COAL	OXYGEN	METHANOL	TOTAL SLAG/ASH	SULFUR
COAL TPD	21,546	-	-	-	-
OXYGEN TPD (100%)	-	14,354	-	-	
METHANOL TPD			16,392	-	-
SLAG ASH TPD	-	-	-	2 477	-
SULFUR TPD	-	-	-	-	909

	6	7	8	9	10	11	12	13	14	15	16	17	18	19
	RAW GAS	CLEAN GAS	SHIFT, CO2 BYPASS	SHIFT FEED	SHIFT EFFLUENT	CO2 EFFLUENT	COMPRESSOR	CRYOGENIC FEED	TAIL GAS	CO2 [illegible]	SYNTHESIS GAS	PURGE	CO2 VENT	FUEL GAS PRODUCT
MOL %														
H2	35 16	42 04	42 04	42 04	58 67	82 40	58 49	58 56		-	65.83	46 65	-	-
CO	40 87	48.94	48 94	48 94	6 27	8 68	32 37	32 41	28 13	-	32 90	22 92	-	28 13
CO2	14.18	0 02	0 02	0 02	28 70	0 36	0 15	-	-	100	0 17	3 83	100	-
CH4	7 02	8 39	8 39	8 39	5 93	8 37	8 38	8 40	70 30	-	0 62	14 10	-	70 30
N2+A	0 56	0 61	0 61	0 61	0 43	0 61	0.61	0 63	1 57	-	0 48	11 10	-	1 57
H2S+COS	1 26	-	-	-	-	-	-	-	-	-	-	-		-
CnHm	-	-	-	-	-	-	-	-	-	-	-	1 39		-
NH3	0 74	-	-	-	-	-	-	-	-	-	-	-	-	-
TOTAL (DRY)	100 00	100 00	100 00	100 00	100 00	100 00	100 00	100 00	100 00	100	100 00	100 00	100	100 00
MMSCFD (DRY)	1621	1356	799	557	782	559	1356	1356	151	2	1207	53	456	21
PSIG	310	305	305	305	330	275	775	700	5	470	690	665	0	50

ELECTRIC POWER RESEARCH INSTITUTE
CASE 1 - FOSTER WHEELER GASIFICATION
CHEM SYSTEMS METHANOL
PROCESS BLOCK FLOW DIAGRAM

THE RALPH M PARSONS COMPANY
PASADENA, CALIFORNIA

JOB NO. 5604-1
DWG NO. 5604-FS-1-A
REV 0

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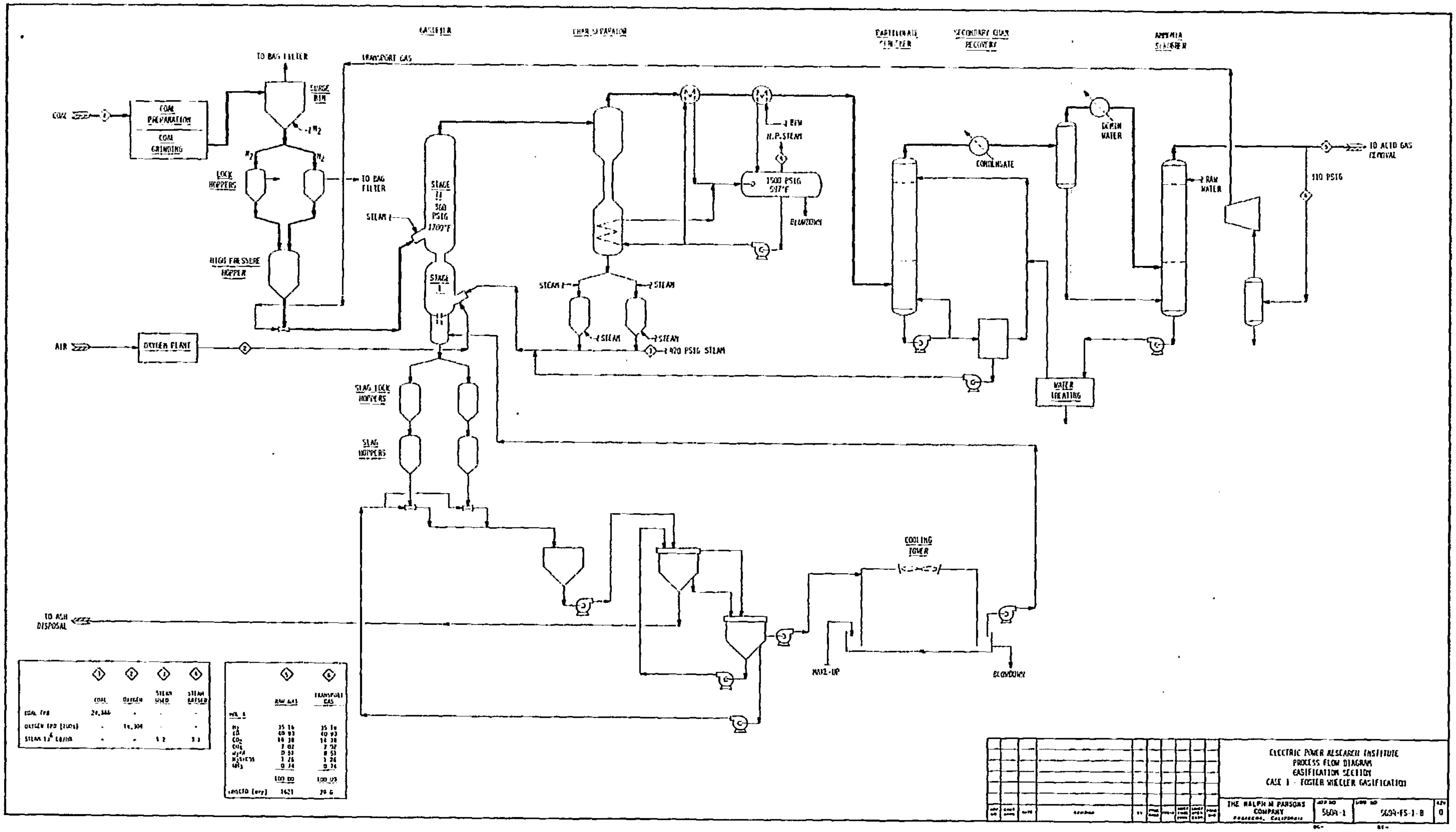

	①	②	③	④
	COAL	OXYGEN	STEAM USED	STEAM RAISED
COAL TPD	20,344	-	-	-
OXYGEN TPD (100%)	-	14,304	-	-
STEAM 10^6 LB/HR	-	-	1.2	3.3

	⑤	⑥
	RAW GAS	TRANSPORT GAS
MOL %		
H_2	35.16	35.16
CO	40.93	40.93
CO_2	14.38	14.38
CH_4	7.02	7.02
N_2+A	0.51	0.51
H_2S+COS	1.26	1.26
NH_3	0.74	0.74
	100.00	100.00
MMSCFD (dry)	1621	79.6

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PROCESS FLOW DIAGRAM
GASIFICATION SECTION
CASE 1 - FOSTER WHEELER GASIFICATION

THE RALPH M PARSONS COMPANY
PASADENA, CALIFORNIA

JOB NO	DWG NO	REV
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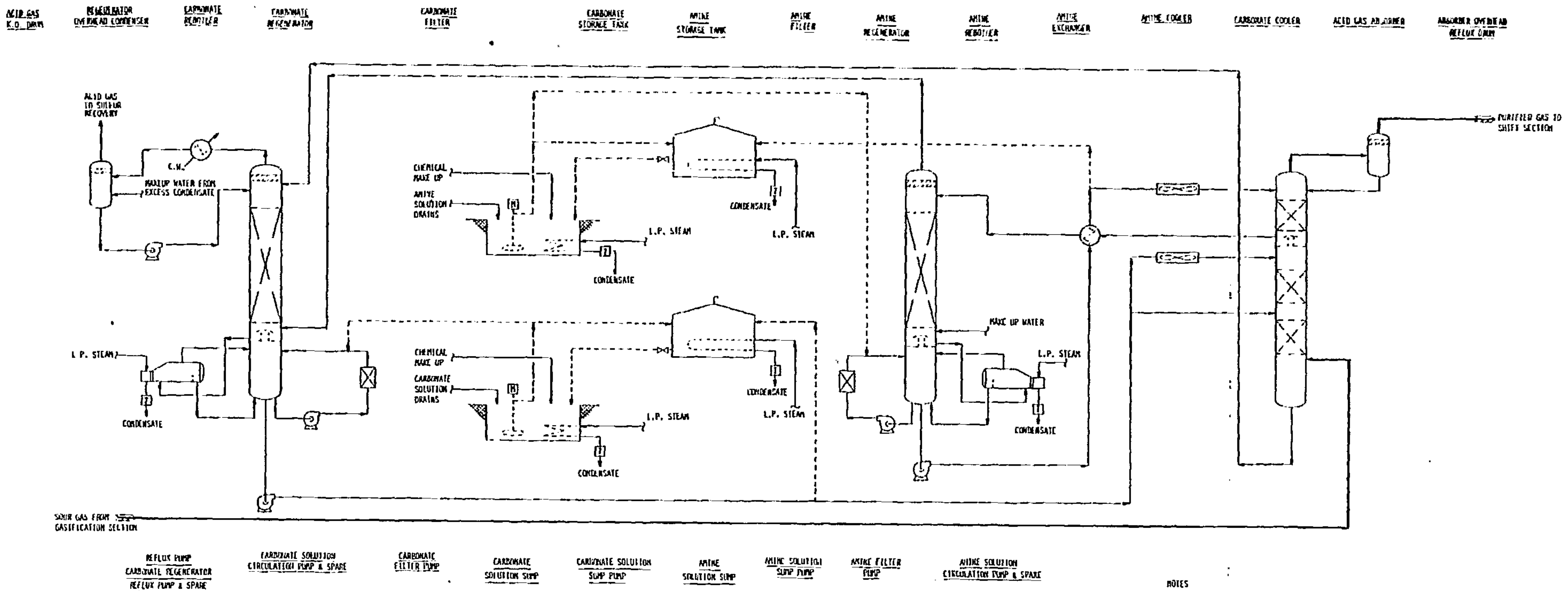


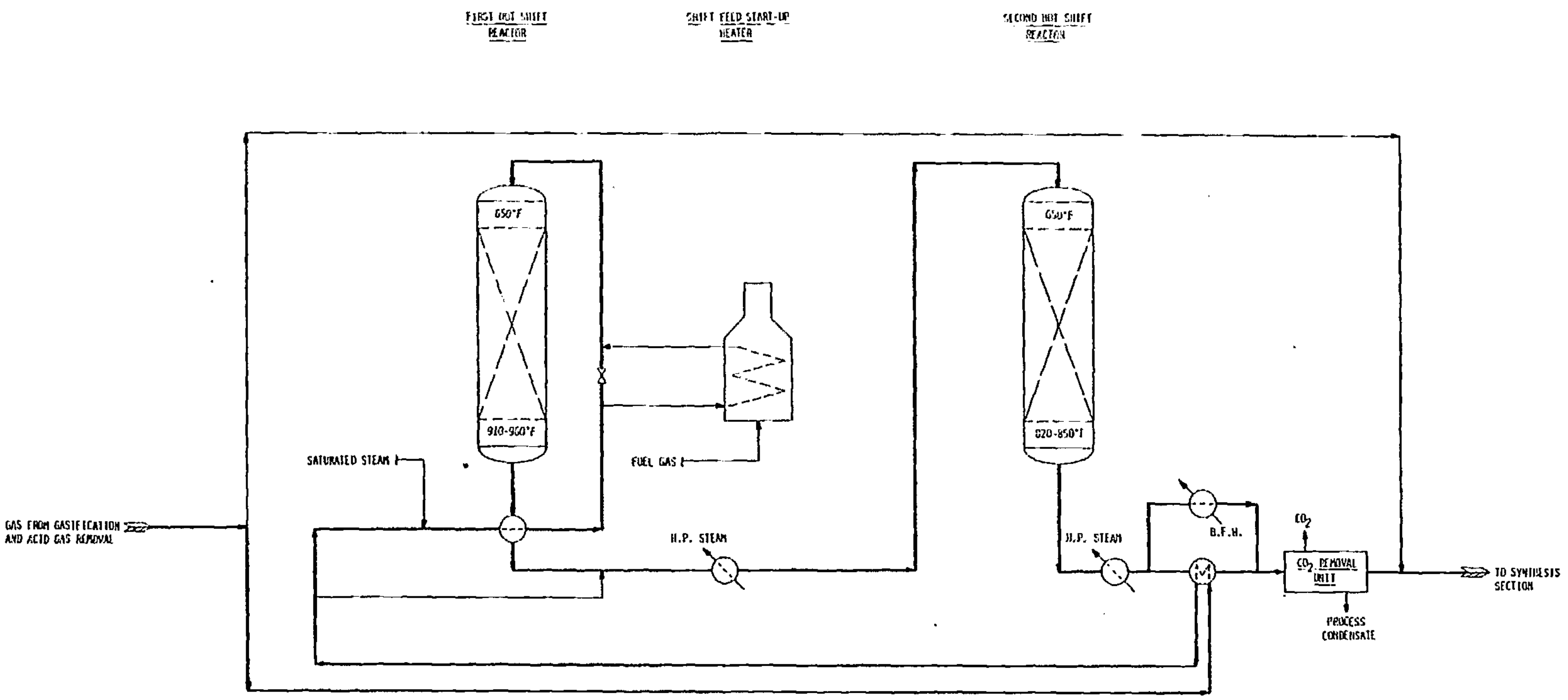
NOTES

1. THIS DRAWING IS A SIMPLIFIED VERSION OF BENFIELD HI PURE PROCESS IN ACCORDANCE WITH DISCLOSURE AGREEMENTS.
2. SEE BLOCK FLOW DIAGRAM FOR EACH CASE FOR FLOWS, COMPOSITIONS AND PRESSURES.

ELECTRIC POWER RESEARCH INSTITUTE
PROCESS FLOW DIAGRAM
ACID GAS REMOVAL

THE RALPH M PARSONS COMPANY
PASADENA, CALIFORNIA

JOB NO	DWG NO	REV
5604-1	5604-FS-1-C	0



NOTE. SEE BLOCK FLOW DIAGRAM FOR EACH CASE FOR FLOWS, COMPOSITIONS AND PRESSURES.

ELECTRIC POWER RESEARCH INSTITUTE
PROCESS FLOW DIAGRAM
SHIFT SECTION

THE RALPH M PARSONS COMPANY PASADENA, CALIFORNIA	JOB NO 5604-1	DWG NO 5604-FS-1-D	REV 0

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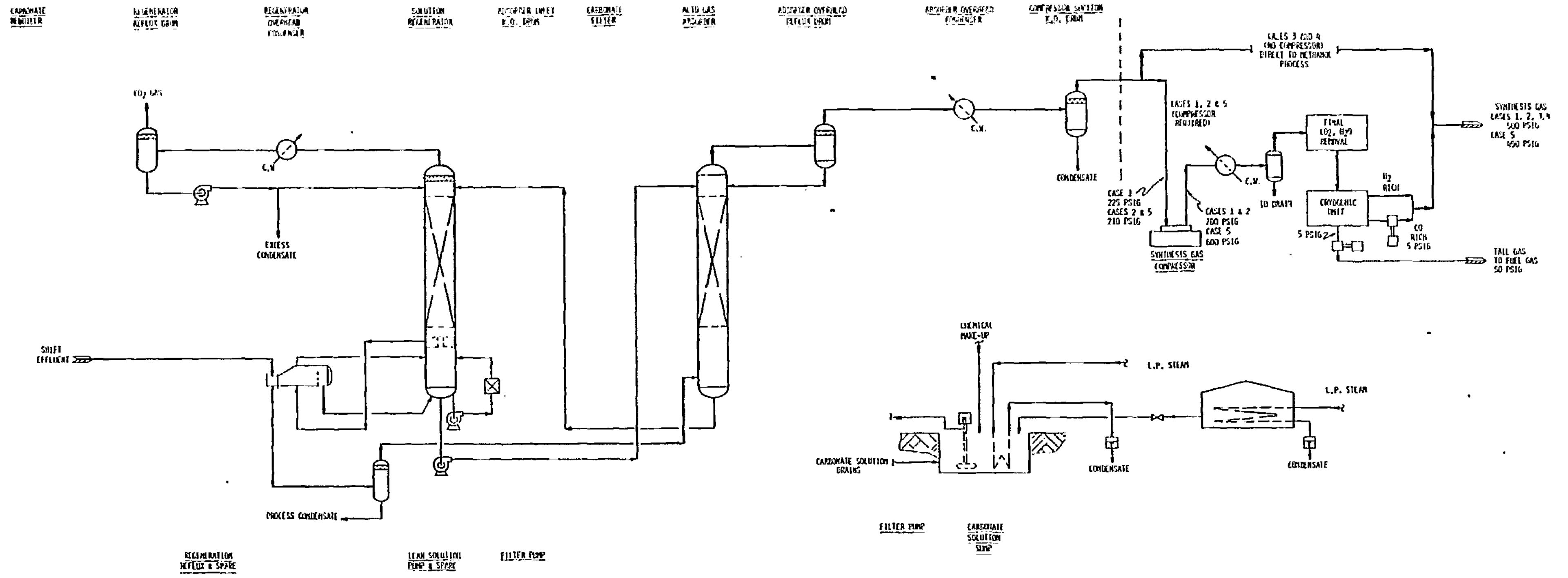

NOTES.

1. THIS DRAWING IS A SIMPLIFIED VERSION OF THE BENFIELD PROCESS IN ACCORDANCE WITH DISCLOSURE AGREEMENT.
2. SEE BLOCK FLOW DIAGRAM FOR EACH CASE FOR FLOWS, COMPOSITIONS AND PRESSURES.

ELECTRIC POWER RESEARCH INSTITUTE
PROCESS FLOW DIAGRAM
CO_2 REMOVAL

THE RALPH M PARSONS COMPANY
PASADENA, CALIFORNIA

JOB NO	DWG NO	REV
5604-1	5604-FS-1-E	0

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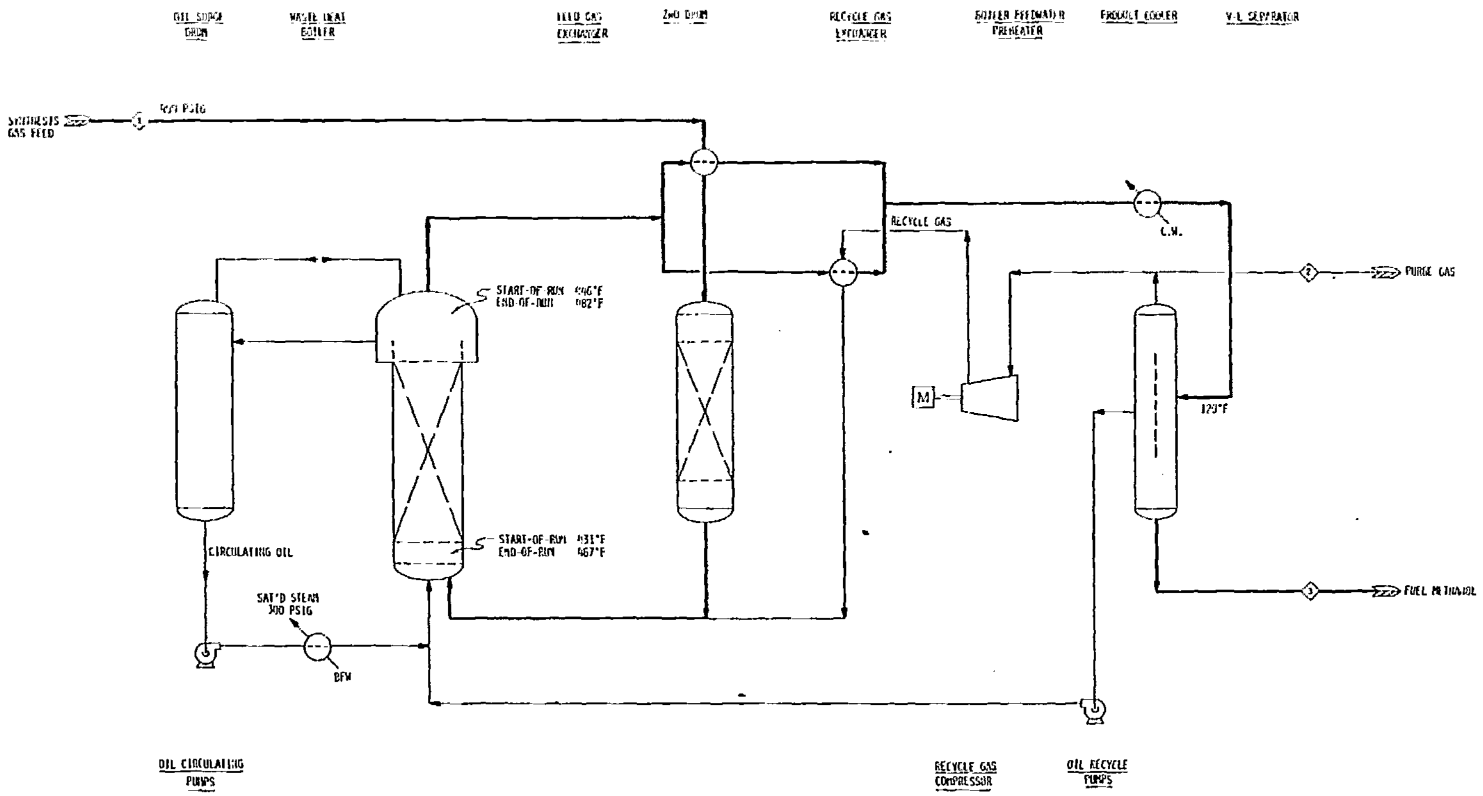


	⟨1⟩ SYNTHESIS GAS FEED	⟨2⟩ PURGE GAS	⟨3⟩ METHANOL PRODUCT
	MOL %	MOL %	MOL %
H_2	65.83	46.47	-
CO	32.90	22.83	-
CO_2	0.17	3.82	-
CH_4	0.62	14.65	-
N_2+A	0.49	11.06	-
CH_3OH	-	1.39	94.48
H_2O	-	0.28	4.36
BY PRODUCTS	-	-	1.16
	100.02	100.00	100.00
MMSCFD (total)	1207	53	-
TPD	-	-	16,392

ELECTRIC POWER RESEARCH INSTITUTE
PROCESS FLOW DIAGRAM
CHEM-SYSTEM
LIQUID PHASE METHANOL

THE RALPH M PARSONS COMPANY
PASADENA, CALIFORNIA

JOB NO 5609-5
DWG NO 5609-FS-1-F
REV 0

SECTION 7

CASE 2

BRITISH GAS COUNCIL/LURGI SLAGGING GASIFICATION - CHEM SYSTEMS METHANOL

The overall plant is shown on Process Block Flow Diagram 5604-FS-2-A. A total of 22,918 tons/day of coal are used, all being gasified. Fuel gas recovered from the process is used in the utility plant and some surplus fuel gas is available as a product.

Plant thermal efficiency is as follows:

Input

Coal

22,918 TPD x 2,000 lb/T x 12,235 Btu/lb. = 560,803 MMBtu/day HHV

Output

Methanol

16,392 TPD x 2,000 lb/T x 9,610 Btu/lb. = 315,000 MMBtu/day (HHV)

Excess fuel gas

5.29 MMSCFD x 850 Btu/scf = 4,500 MMBtu/day (HHV)

Total = 319,500 MMBtu/day (HHV)

$$\text{Efficiency} = \frac{319{,}500}{560{,}803} \times 100 = \underline{\underline{57.0\%}} \text{ (HHV)}$$

$$\text{Efficiency} = \frac{276{,}700 \times 4{,}108}{536{,}693} \times 100 = \underline{\underline{52.3\%}} \text{ (LHV)}$$

Coal Preparation

Coal is unloaded and conveyed to the stacking system. It is reclaimed and fed to the gasification coal hopper. The coal is assumed to be supplied sized at 1-1/2" x 1/4" suitable for gasification.

Gasification

Refer to Process Flow Diagram 5604-FS-2-B. Gasification, Case 2 - British Gas Council/Lurgi Slagging Gasification.

Sized coal is fed to the gasifiers through the pressurized lock system. The coal descends through the "moving bed" gasifier and is dried, devolatized, gasified and combusted. The gases rise counter currently to the descending coal. At the bottom oxygen and steam are fed, together with recycle tar, oil, naphtha and phenols.

Slag is removed via a water quench chamber and lock system and water-conveyed to separating tanks. The ash/slag is separated for disposal and the water recycled via a cooling tower.

The gas from the gasifier is immediately quenched in a scrubber by gas liquor and then cooled in successive stages before a final water wash for ammonia removal.

Tar Separation

Refer to to Process Flow Diagram 5604-FS-2-C, Tar Separation, Case 2 - British Gas Council/Lurgi Slagging Gasification.

Dusty tar liquor from the gasifier quench pot and oily gas liquor from the gas cooling section are separated in a series of separators and settling vessels into several streams.

One water stream is recycled to the gasifier quench section and another is the "gas liquor" which passes to the phenol section. Tar is produced as the heaviest material from each separator. An oil fraction is also recovered. The tar and the oil together with phenols from the phenol section are remixed and recycled back to the gasifier for disposal by gasification.

Phenol Extraction

Refer to Process Flow Diagram 5604-FS-2-D, Phenol Extraction, British Gas Council/Lurgi Slagging Gasification.

The gas liquor from the tar separation area needs treating in several stages before disposal or reuse.

The first step is to remove the phenols by using the Lurgi Phenolsolvan process. The phenol is extracted from water in an organic solvent. Excess solvent is removed from the water ("raffinate") by stripping with nitrogen; tne solvent is then recovered from the nitrogen by absorption and distillation. The extract stream, solvent containing phenols, is fractionated to recover solvent as an overhead and reject the phenols as a net bottom product. Phenols are returned to the tar separation area for mixing with tar and oil and then recycled to the gasifier. The phenol-free water is sent to the process water treating unit.

Acid Gas Removal, CO Shift, CO_2 Removal, Compression, Cryogenic Separation and Methanol Synthesis

These sections are similar to those described for Case 1. See Drawings 5604-FS-1-C, 5604-FS-1-D, 5604-FS-1-E and 5604-FS-1-F.

Approximate Arrangement

The plant consists of coal handling for sizing coal (no pulverizing) to feed 14 British Gas Council-Lurgi, slagging gasifiers with 5 trains of tar oil separation and 2 trains of Phenolsolvan. The gas is handled in 4 trains of acid gas removal, 4 shift trains and 7 trains of CO_2 removal. After compression and several trains of cryogenic separation, there are 7 trains of methanol synthesis, all with multiple unit reactors. Two sulfur plants are included.

Utility System

The plant is balanced so that the steam generated provides sufficient steam for the turbines for plant compressors, power generation and other steam requirements. A total of 4.1×10^6 lb/hr. of steam at 1500 psig, 950°F is produced.

About 14% is produced (saturated) in the shift section and the balance 85% is produced in the utility boiler/superheater system which also superheats the total steam.

The HP steam is used in back pressure and condensing turbines to drive the main plant turbines and to provide 0.6×10^6 lb/hr. of gasifier process steam at 400 psig.

The large turbines are:

	HP
Synthesis Gas and Cryogenic Compressor	271,000
Oxygen Plant Compressors	258,000
Methanol Recycle Compressor	23,000
Power Generators	71,000
Total	623,000

Steam from the methanol synthesis 1.5×10^6 lb/hr., produced at 285 psig saturated, is used as shift process steam.

Steam from back pressure turbines at 50 psig and 100 psig is used in the acid gas removal reboilers, Phenolsolvan unit and other miscellaneous users.

Condensate is recovered from vacuum condensers, steam reboilers, and shift condensate and returned to the deaerator.

Make-up boiler water required is 1.4×10^6 lb/hr. corresponding to steam converted in gasification, shift sections and losses.

CAPITAL SUMMARY - CASE 2

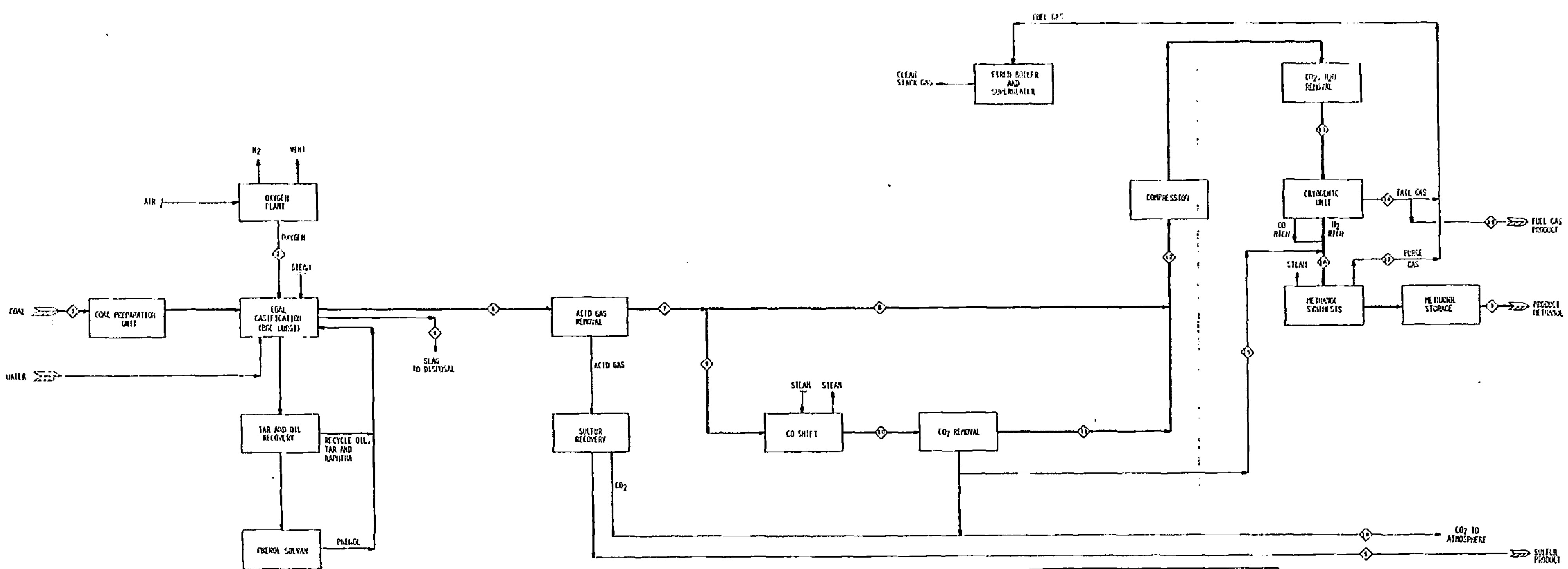
	MM $
Coal Preparation	28
Gasification	101
Tar Recovery & Phenol Extraction	69
Acid Gas Removal	21
CO Shift	39
CO_2 Removal	147
Oxygen	183
Cryogenic Clean-up and Boxes	21
Methanol Synthesis	122
Final Desulfurization - Zinc Oxide	10
Raw Gas Compression	53
Sulfur Recovery	20
Utility Boilers, Superheaters and Power Generator	56
	869
Utilities and General Offsites at 15%	130
Total Plant Cost, Including Engineering	1,000
Contingency at 15%	150
Total Plant Investment	1,150
Interest on Construction Loan	255
Royalty Allowance	6
Startup and Working Capital	169
Total Capital Requirement	1,580 MM $

PRODUCTION COST* - CASE 2

Illinois Coal	-	25$/ton at 12,235 Btu/lb. HHV (1.02 $ per MM Btu)
Coal	-	22,918 TPD
Overall Efficiency	-	57.0% HHV
Liquid Methanol Product	-	315,000 MM Btu/day HHV
Fuel Gas Product	-	4,500 MM Btu/day HHV
Total Product	-	319,500 MM Btu/day HHV
Total Capital Requirement	-	1,580 million dollars
Operating Factor	-	90%, 328.5 days/year

	First Year Cost $/MM Btu	Levelized Cost $/MM Btu
Coal	1.79	1.79
Water, Catalyst & Chemicals	0.07	0.07
Other Operating Costs	0.79	0.79
Total Operating Cost	2.65	2.65
Capital Charges	3.05	2.35
Production Cost	5.70 HHV	5.00 HHV
	6.49 LHV	5.69 LHV

*Calculated in accordance with EPRI's Economic Premises, Feb. 11, 1977 - utility financing.



	6	7	8	9	10	11	12	13	14	15	16	17	18	19
	RAW GAS	CLEAN GAS	SHIFT, CO2 BYPASS	SHIFT FEED	SHIFT EFFLUENT	CO2 EFFLUENT	COMPRESSION	CRYOGENIC FEED	TAIL GAS	CO2 RECYCLE	SYNTHESIS GAS	PURGE	CO2 VENT	FUEL GAS PRODUCT
MOL %														
H_2	29.96	31.21	31.21	31.21	53.14	78.38	58.48	58.57	-	-	65.83	46.44	-	-
CO	57.18	59.69	59.69	59.69	8.42	12.37	32.36	32.48	28.08	-	32.90	22.12	-	28.08
CO_2	1.91	0.02	0.02	0.02	32.12	0.25	0.15	-	-	100	0.67	3.43	100	-
CH_4	7.65	7.98	7.98	7.98	5.47	7.96	7.97	7.95	66.47	-	0.62	14.10	-	66.47
C_2H_6	0.33	0.35	0.35	0.35	0.24	0.35	0.35	0.35	3.12	-	-	-	-	3.12
C_2H_4	0.21	0.29	0.29	0.29	0.16	0.21	0.26	0.24	2.11	-	-	-	-	2.11
N_2+A	0.41	0.46	0.46	0.46	0.31	0.46	0.46	0.44	0.22	-	0.49	11.10	-	0.22
$COS+H_2S$	1.42	-	-	-	-	-	-	-	-	-	-	-	-	-
C_6H_5OH	-	-	-	-	-	-	-	-	-	-	-	1.33	-	-
NH_3	0.93	-	-	-	-	-	-	-	-	-	-	-	-	-
TOTAL (DRY)	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100	100.00	100.00	100	100.00
MMSCFD (DRY)	1416	1356	574	782	1153	781	1350	1354	152	3	1207	53	775	5
PSIG	225	245	220	220	220	210	210	700	50	450	750	915	0	50

	1	2	3	4	5
	FEED COAL	OXYGEN	METHANOL PRODUCT	SLAG-ASH	SULFUR
COAL, TPD	22,916		-	-	-
OXYGEN, TPD (100%)	-	10,534	-	-	-
METHANOL, TPD	-	-	16,392	-	-
SLAG-ASH, TPD	-	-	-	2,215	-
SULFUR, TPD	-		-	-	848

ELECTRIC POWER RESEARCH INSTITUTE
CASE 2 - BRITISH GAS COUNCIL/LURGI SLAGGING
GASIFICATION; CHEM SYSTEMS METHANOL
PROCESS BLOCK FLOW DIAGRAM

THE RALPH M PARSONS COMPANY
PASADENA, CALIFORNIA

JOB NO 5604-1
DWG NO 5604-FS-2-A
REV 0

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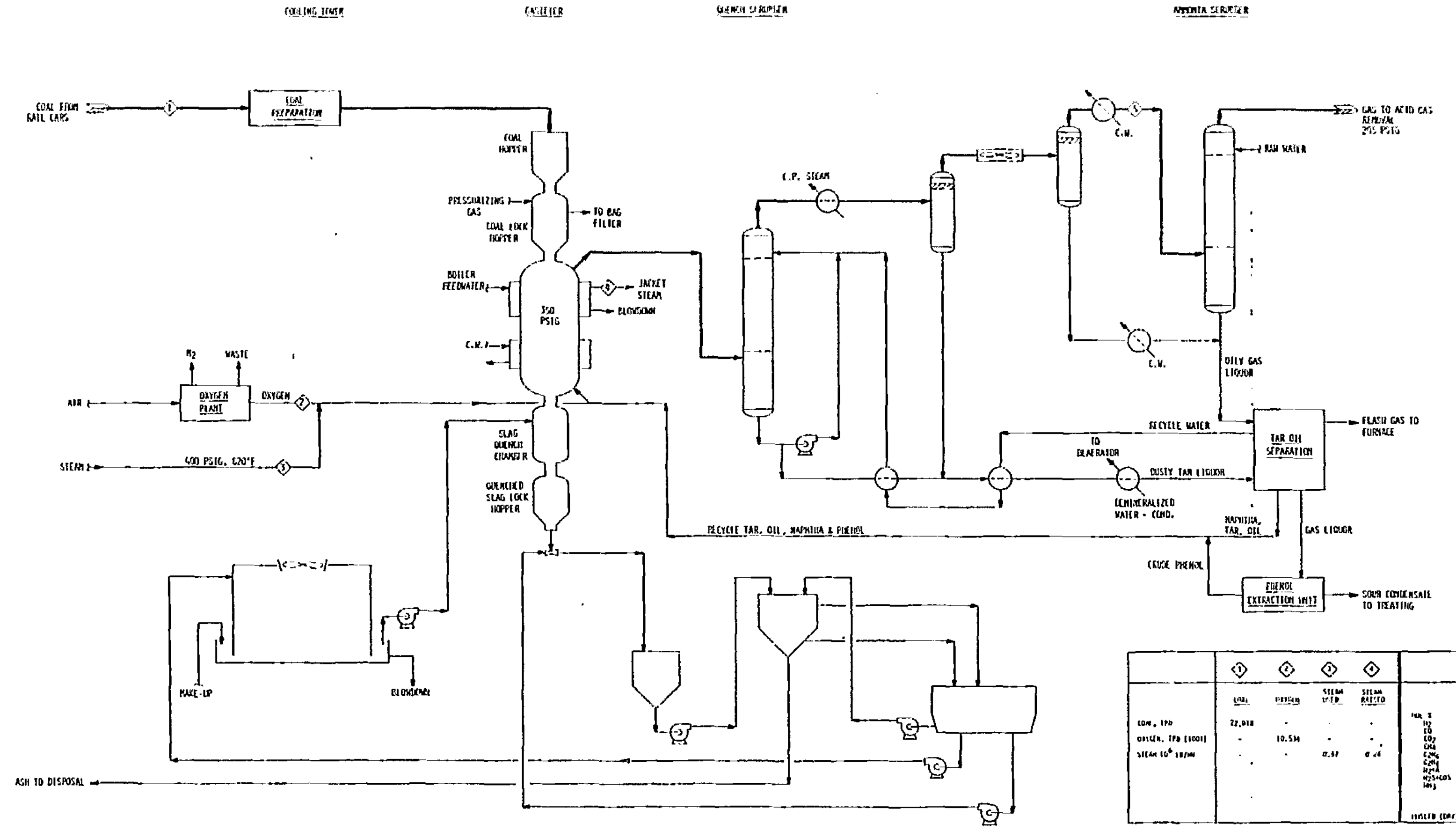
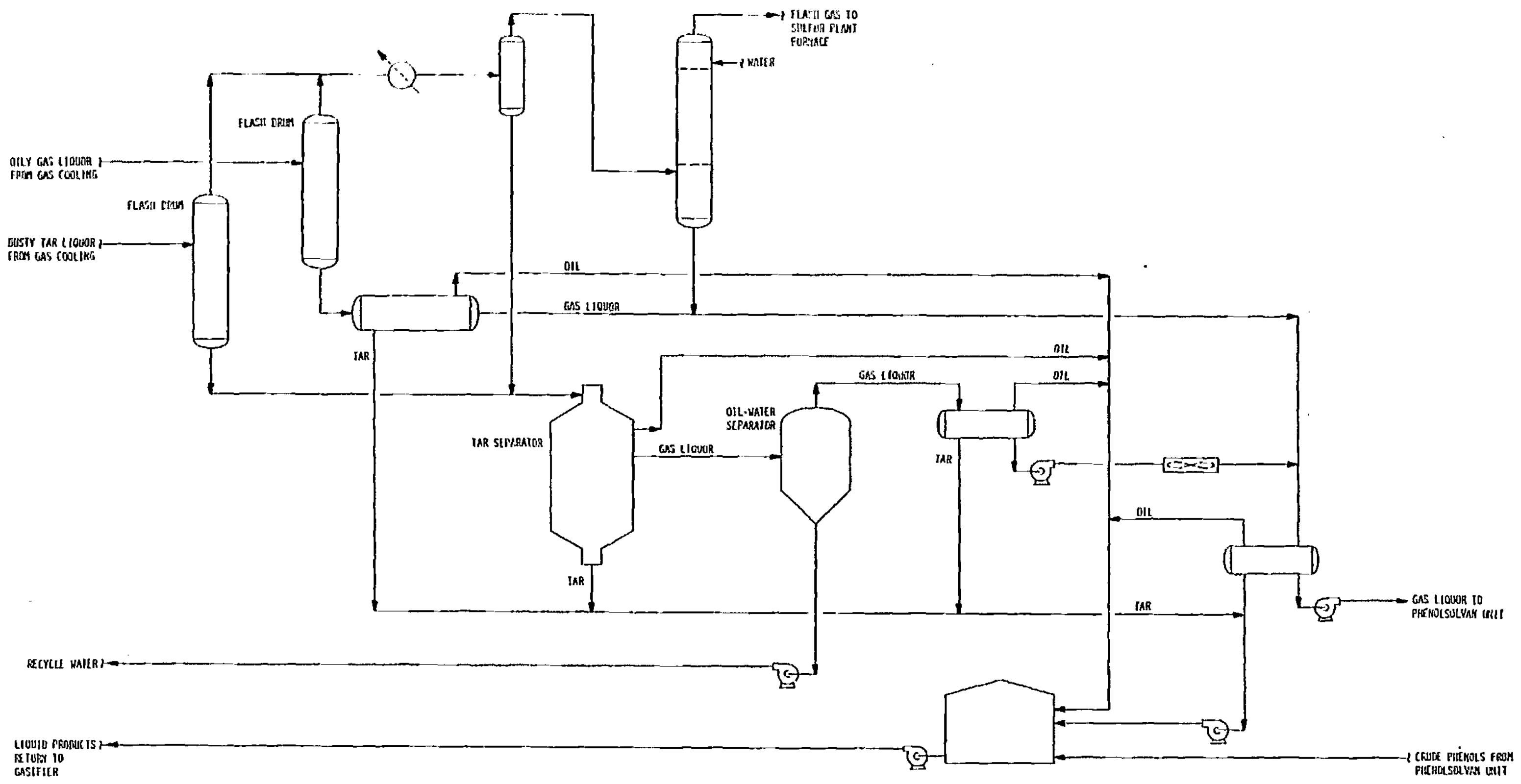


	①	②	③	④		⑤
	COAL	OXYGEN	STEAM USED	STEAM RAISED		RAW GAS
					MOL %	
COAL, TPD	22,018	-	-	-	H_2	29.96
OXYGEN, TPD (100%)	-	10,534	-	-	CO	57.16
STEAM 10^6 LB/HR	-	-	0.57	0.46	CO_2	1.91
					CH_4	7.65
					C_2H_6	0.33
					C_2H_4	0.23
					N_2+A	0.44
					H_2S+COS	1.42
					NH_3	0.89
						100.00
					MMSCFD (DRY)	1416

ELECTRIC POWER RESEARCH INSTITUTE
PROCESS FLOW DIAGRAM
GASIFICATION SECTION
CASE 2 - BRITISH GAS COUNCIL/LURGI
SLAGGING GASIFIER

THE RALPH M. PARSONS COMPANY PASADENA, CALIFORNIA	JOB NO. 5604 1	DWG NO. 5604-FS-2-B	REV 0



ELECTRIC POWER RESEARCH INSTITUTE
PROCESS FLOW DIAGRAM
TAR, OIL SEPARATION
CASE 2 - BRITISH GAS COUNCIL-
LURGI SLAGGING GASIFIER

THE RALPH M PARSONS
COMPANY
PASADENA, CALIFORNIA

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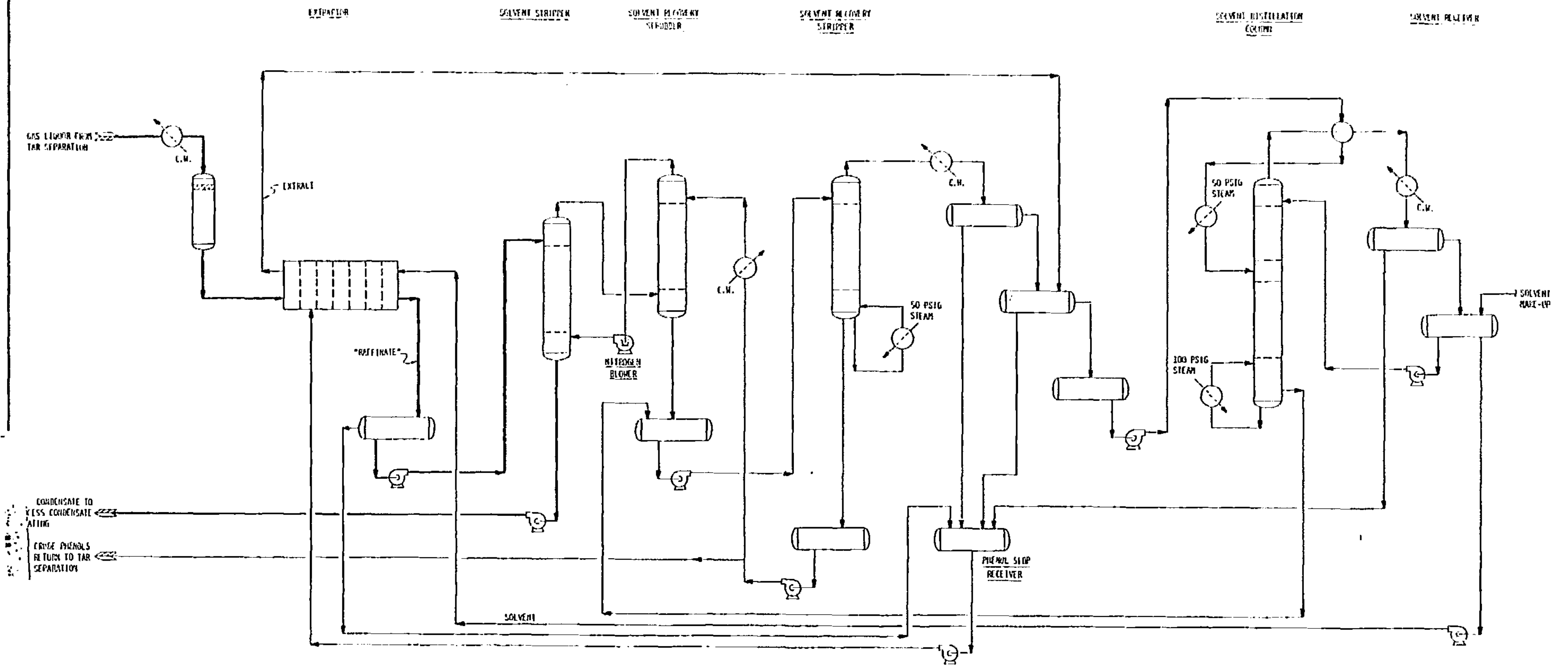
										5604-1	5604-FS-2-C	0



ELECTRIC POWER RESEARCH INSTITUTE
PROCESS FLOW DIAGRAM
PHENOL EXTRACTION
CASE 2 - BRITISH GAS COUNCIL
LURGI SLAGGING GASIFIER

THE RALPH M. PARSONS COMPANY
PASADENA, CALIFORNIA

JOB NO	DWG NO	REV
5604-1	5604-FS-2-D	0

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SECTION 8

CASE 3

KOPPERS-TOTZEK GASIFICATION - CHEM SYSTEMS METHANOL

The overall plant is shown on Process Block Flow Diagram 5604-FS-3-A. A total of 24,574 tons/day of coal are used; 20,702 tons/day are gasified and 3,872 tons/day are used in the utility plant to supplement fuel gas.

Plant thermal efficiency is as follows:

Input

Coal

24,574 TPD x 2,000 lb/T x 12,235 Btu/lb. = 601,325 MMBtu/day (HHV)

Output

Methanol

16,392 TPD x 2,000 lb/T x 9,610 Btu/lb. = 315,000 MMBtu/day (HHV)

$$\text{Efficiency} = \frac{315{,}000}{601{,}325} \times 100 = \underline{\underline{52.4\%}} \text{ (HHV)}$$

$$\text{Efficiency} = \frac{276{,}700}{575{,}470} \times 100 = \underline{\underline{48.1\%}} \text{ (LHV)}$$

Coal Preparation

Coal is unloaded and conveyed to the stacking system. It is reclaimed and passed to the pulverizing section where it is pulverized to about 70% through 200 mesh. The coal dust is then conveyed with nitrogen to the gasifier feed system.

Gasification

Refer to Process Flow Diagram 5604-FS-3-B, Gasification, Case 3, Koppers-Totzek Gasification. Each gasifier has 4 "burners" or "heads" each fed with coal, steam and oxygen. The coal is fed by a screw feeder and at the mixing head it is entrained by the mixture of steam and oxygen.

A very high temperature flame zone, around 3,500°F is followed by endothermic reactions between carbon and steam which lower the temperature to about 2,700°F.

The gasifier is jacketed raising low pressure steam; the gas passes to a heat recovery section where high pressure, superheated steam is raised.

Slag falls into a quench tank and thence it is removed for disposal in a water stream.

After the boiler section, the gas is water-scrubbed in two stages followed by a final cooler. Water from the scrubbers and coolers,as well as the slag tank, overflows to a clarifier. Sludge is separated for disposal and water is recycled via a cooling tower to the scrubber and cooler. The cooled gas is at essentially atmosphere pressure; it is then compressed to 600 psia.

Acid Gas Removal, CO Shift, CO_2 Removal and Methanol Synthesis

These sections are essentially similar to Case 1. See Flow Sheets 5604-FS-1-C, 5604-FS-1-D, 5604-FS-1-E, and 5604-FS-1-F.

No cryogenic unit is required.

Approximate Arrangement

The plant consists of 5 pulverizers feeding 26 Koppers-Totzek, four-headed, gasifiers. After compression, the gas is handled by four main trains of acid gas removal with some multiple units for absorbers, 4 shift trains and 6 trains of CO_2 removal.

Finally, 7 trains of methanol synthesis all with multiple units of reactors are required. Two sulfur plants are included.

Utility System

The plant is balanced so that the steam generated is sufficient for the steam turbine drivers of the plant compressors, power generation and other steam requirements. A total of 5.8×10^6 lb/hr. of steam at 1500 psig, 950°F is produced.

Of the total steam about 41% is produced (superheated) in the Koppers-Totzek gasification section, about 12% is produced (saturated) in the shift section and the balance 47% is produced in the utility boiler/superheater system which also superheats the shift steam.

The HP steam is used in back pressure and condensing turbines to drive the main plant turbines and to provide 1.4×10^6 lb/hr. of shift process steam at 600 psig.

The large turbines are:

	HP
K-T Raw Gas Compressor	440,000
Oxygen Plant Compressors	306,000
Methanol Recycle Compressors	23,000
Power Generator	51,000
Total	820,000

Heat recovered in the methanol synthesis section is used partly as boiler feedwater preheat but mostly to produce steam, 1.4×10^6 lb/hr., 285 psig (saturated). This steam is superheated and used for power generation (62,000 HP) before final use at 50 psig in the acid gas removal reboilers, process steam for K-T gasifiers and miscellaneous uers. Total power generation is 113,000 HP.

Condensate is recovered from vacuum condensers, steam reboilers and shift condensate and returned to the deaerator.

Make-up boiler water required is 1.3×10^6 lb/hr. corresponding to steam converted in gasification, shift sections and losses.

CAPITAL SUMMARY - CASE 3

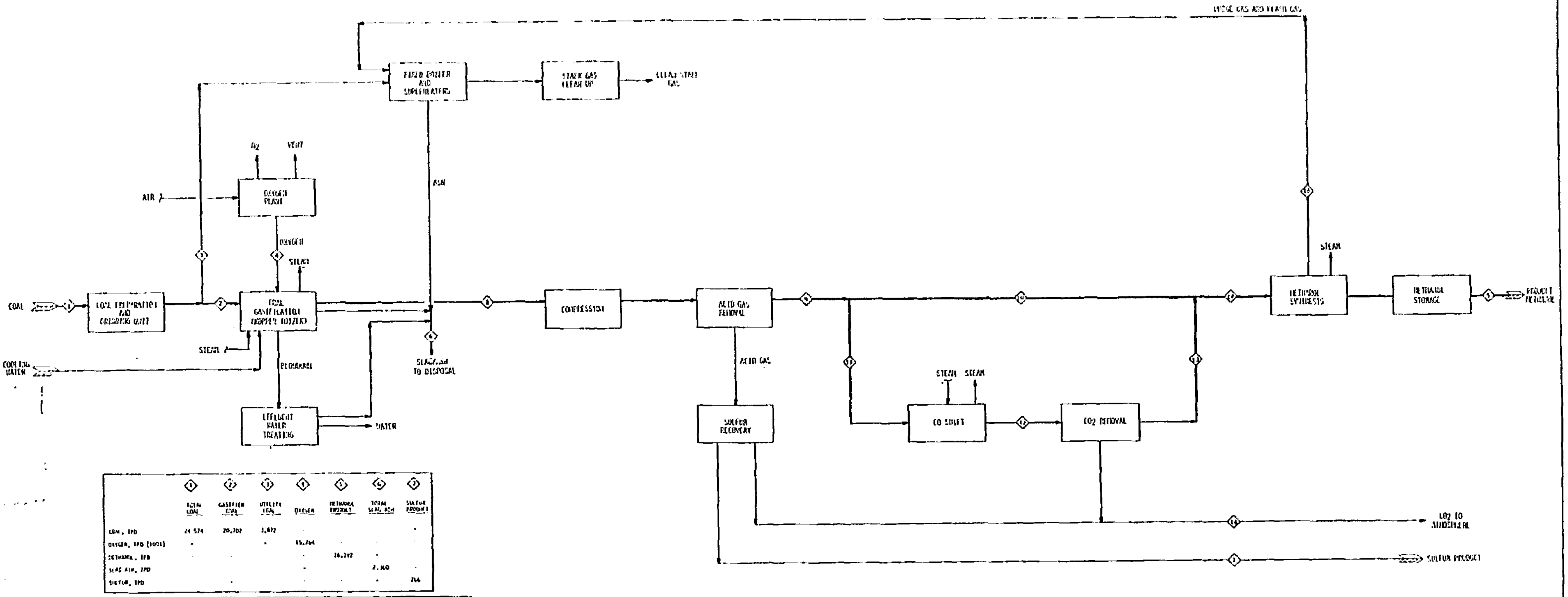
	MM $
Coal Preparation and Grinding	60
Gasification	390
Acid Gas Removal	54
CO Shift	35
CO_2 Removal	129
Oxygen	258
Methanol Synthesis	122
Final Desulfurization - Zinc Oxide	10
Raw Gas Compression	86
Sulfur Recovery	20
Boilers, Superheaters, Power Generators, and Stack Gas Clean-up	136
	1,300
Utilities and General Offsites at 15%	195
Total Plant Cost, Including Engineering.	1,495
Contingency at 15%	224
Total Plant Investment	1,719
Interest on Construction Loan	382
Royalty Allowance	9
Startup and Working Capital	232
Total Capital Requirement	2,342

PRODUCTION COST* - CASE 3

Illinois Coal	-	25$/ton at 12,235 Btu/lb. HHV (1.02 $ per MM Btu)
Coal	-	24,574 TPD
Overall Efficiency	-	52.4% HHV
Liquid Methanol Product	-	315,000 MMBtu/day HHV
Total Capital Requirement	-	2,342 million dollars
Operating Factor	-	90%, 328.5 days/year

	First Year Cost $/MMBtu	Levelized Cost $/MMBtu
Coal	1.95	1.95
Water, Catalyst & Chemicals	0.08	0.08
Other Operating Costs	1.19	1.19
Total Operating Cost	3.22	3.22
Capital Charges	4.59	3.53
Production Cost	7.81 HHV	6.75 HHV
	8.89 LHV	7.69 LHV

*Calculated in accordance with EPRI's Economic Premises, Feb. 11, 1977 - utility financing.



	① TOTAL COAL	② GASIFIER COAL	③ UTILITY COAL	④ OXYGEN	⑤ METHANOL PRODUCT	⑥ TOTAL SLAG, ASH	⑦ SULFUR PRODUCT
COAL, TPD	24,574	20,702	3,872	-			-
OXYGEN, TPD (100%)	-		-	15,764	-	-	-
METHANOL, TPD	-	-		-	18,392	-	
SLAG ASH, TPD				-		2,160	-
SULFUR, TPD		-		-	-	-	766

	⑧ RAW GAS	⑨ CLEAN GAS	⑩ SHIFT, CO_2 BYPASS	⑪ SHIFT FEED	⑫ SHIFT EFFLUENT	⑬ CO_2 EFFLUENT	⑭ METHANOL FEED	⑮ PURGE	⑯ CO_2 VENT
MOL %									
H_2	35.20	38.68	38.68	38.68	58.24	65.32	65.77	45.75	-
CO	54.70	60.11	60.11	60.11	9.02	33.21	32.07	22.49	-
CO_2	7.63	0.02	0.02	0.02	31.92	0.28	0.17	3.77	100
CH_4		-	-	-	-	-		-	-
N_2+A	1.09	1.19	1.19	1.19	0.81	1.19	1.99	26.58	-
H_2S+COS	1.37	-	-	-		-	-	-	-
CH_3OH		-	-	-	-	-	-	1.46	-
TOTAL (DRY)	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100
MSCFD (DRY)	1325	1206	506	700	1028	702	1236	54	422
PSIG	0	575	575	575	445	490	490	450	0

ELECTRIC POWER RESEARCH INSTITUTE
CASE 3 - KOPPERS-TOTZEK GASIFICATION
CHEM SYSTEMS METHANOL
PROCESS BLOCK FLOW DIAGRAM

THE RALPH M. PARSONS COMPANY
PASADENA, CALIFORNIA

JOB NO 5604-1

DWG NO 5604-FS-3-A

REV 0

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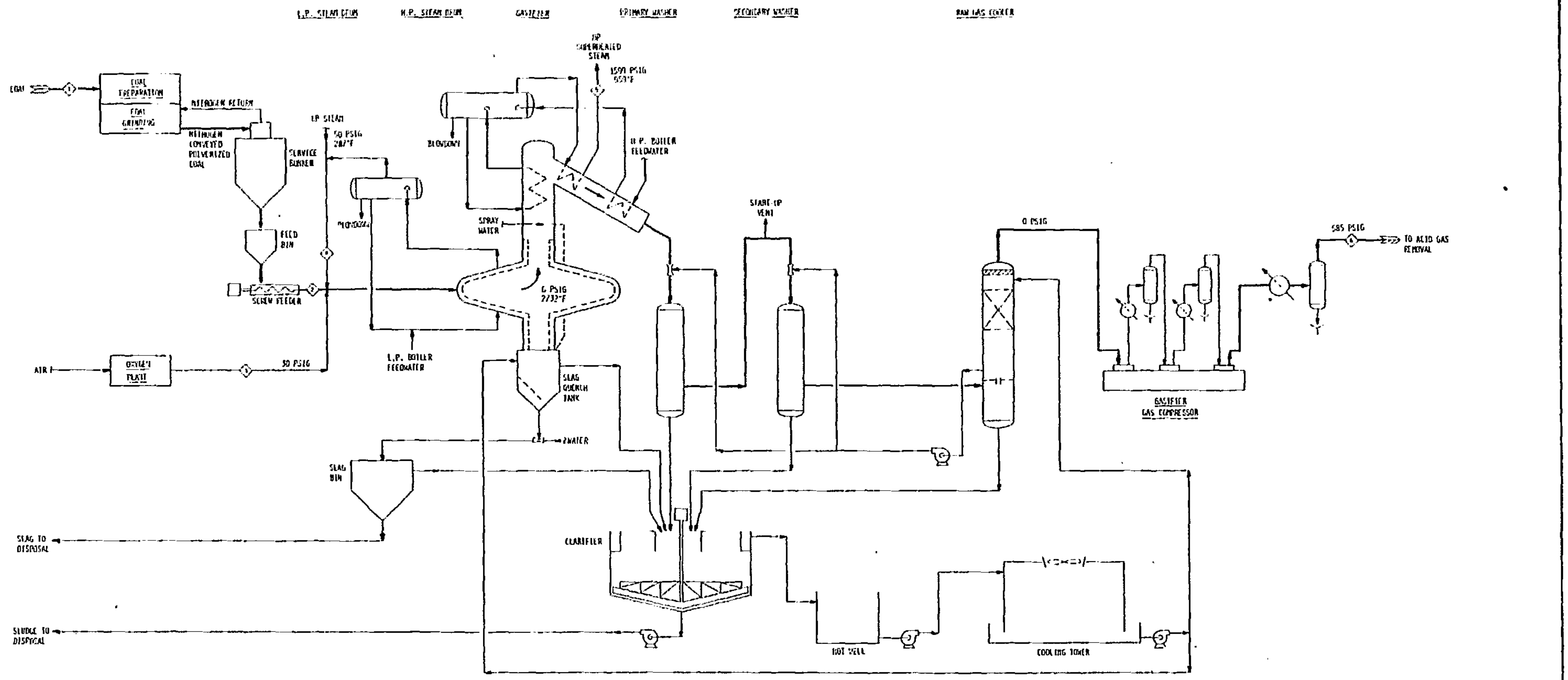

	①	②	③	④	⑤
	COAL	DRIED COAL	OXYGEN	LP STEAM USED	S/H STEAM GENERATED
COAL, TPD	20,202	20,237	·	-	·
OXYGEN, TPD		·	15,764	-	·
STEAM 10^6 LB/HR				0.47	2.33

	⑥
	RAW GAS
VOL %	
H_2	35.20
CO	54.73
CO_2	2.63
CH_4	-
N_2+A	1.03
H_2S+COS	1.17
TOTAL (DRY)	100.00
MMSCFD (DRY)	1375

ELECTRIC POWER RESEARCH INSTITUTE
PROCESS FLOW DIAGRAM
GASIFICATION SECTION
CASE 3 - KOPPERS-TOTZEK GASIFICATION

THE RALPH M PARSONS COMPANY
PASADENA, CALIFORNIA

JOB NO 5609-1
DWG NO 5609-TS-3-D
REV 0

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SECTION 9

CASE 4

TEXACO GASIFICATION - CHEM SYSTEMS METHANOL

The overall plant is shown on Process Block Flow Diagram 5604-FS-4-A. A total of 22,100 tons/day of coal are used; 20,282 tons/day are gasified and 1,818 tons/day are used in the utility plant to supplement fuel gas.

Plant thermal efficiency is as follows:

Input

Coal

22,100 TPD x 2,000 lb/T x 12,235 Btu/lb. = 540,787 MMBtu/day (HHV)

Output

Methanol

16,392 TPD x 2,000 lb/T x 9,610 Btu/lb. = 315,000 MMBtu/day (HHV)

$$\text{Efficiency} = \frac{315{,}000}{540{,}787} \times 100 = \underline{\underline{58.2\%}} \text{ (HHV)}$$

$$\text{Efficiency} = \frac{276{,}700}{517{,}538} \times 100 = \underline{\underline{53.5\%}} \text{ (LHV)}$$

Coal Preparation

Coal is unloaded and conveyed to the stacking system. It is reclaimed and passed to the grinding section. In wet grinding equipment, coal is ground to about 50% through 100 mesh (about 20 mesh grind) and produced in the form of a water slurry.

Gasification

Refer to Process Flow Diagram 5604-FS-4-B, Gasification, Case 4, Texaco Gasification.

The coal-water slurry is pumped at high pressure through a preheater, which uses LP steam, into the gasifier. Here it reacts with oxygen at 650 psig and 2500°F. Most of the gas, about 90-95%, is withdrawn from the side of the gasifier and passes through a waste heat boiler and feedwater heaters raising high pressure saturated steam. To lower the temperature at the inlet of the boiler, a gas recycle quench is used. The remaining 5-10% of the hot gas flows downwards carrying the bulk of the slag and is quenched in a water section before rejoining the main flow of gas in the heat recovery section. A succession of scrubbing stages remove particulate material from the gas. Water from these scrubbing stages and from the gasifier's slag quench-chamber flows to a clarifier.

Fine ash and carbon slurry from the clarifier bottom is recycled to the coal preparation section and clarified water is returned to the scrubbers.

Acid Gas Removal, CO Shift, CO_2 Removal, Methanol Synthesis

These sections are essentially similar to Case 1 - see Flow Sheets 5604-FS-1C, 5604-FS-1D, 5604-FS-1E and 5604-FS-1F.

No synthesis gas compression is required, the gas is produced in the Texaco gasifiers at 650 psig and eventually used in the methanol synthesis at 500 psig.

No cryogenic unit is required.

Approximate Arrangement

The plant consists of 5 pulverizers feeding 16 Texaco gasifiers. After heat recovery and cooling, the gas is processed in four main trains of acid gas removal with some multiple units for absorbers, 4 shift trains and 6 trains of CO_2 removal.

Finally, 7 trains of methanol synthesis all with multiple units of reactors are required. Two sulfur plants are included.

Utility System

The plant is balanced so that the steam generated provides the steam turbine, power generation and other steam requirements. A total of 4.1×10^6 lb/hr. of steam at 1500 psig, 950°F is produced.

Of the total steam, about 79% is produced (saturated) in the Texaco gasification section, about 17% is produced (saturated) in the shift section and the balance 4% is produced in the utility boiler/superheater system which also superheats the total steam.

The HP steam is used in back pressure and condensing turbines to drive the main plant turbines and to provide 1.35 x 10^6 lb/hr. of shift steam.

	HP
Texaco Recycle Compressor	14,000
Oxygen Plant Compressors	446,000
Methanol Recycle Compressors	23,000
Power Generator	40,000
	523,000

Steam from the methanol synthesis 2.0 x 10^6 lb/hr. is produced at 285 psig saturated. Part is used in heat exchangers and part is superheated and used for power generation (72,000 HP) before use at 50 psig in the acid gas removal reboilers and miscellaneous users. Total power generated is 112,000 HP.

Condensate is recovered from vacuum condensers, steam reboilers and shift condensate and returned to the deaerator.

Make-up boiler water required is 0.7 x 10^6 lb/hr. to account for steam converted in gasification and shift sections and losses.

CAPITAL SUMMARY - CASE 4

	MM $
Coal Preparation, Grinding and Slurrying	50
Gasification	265
Acid Gas Removal	70
CO Shift	35
CO_2 Removal	125
Oxygen	296
Methanol Synthesis	122
Final Desulfurization - Zinc Oxide	10
Sulfur Recovery	20
Boilers, Superheaters, Power Generators and Stack Gas Clean-up	73
	1,066
Utilities and General Offsites at 15%	160
Total Plant Cost, Including Engineering	1,226
Contingency at 15%	184
Total Plant Investment	1,410
Interest on Construction Loan	313
Royalty Allowance	7
Startup and Working Capital	195
Total Capital Requirement	1,925

PRODUCTION COST * - CASE 4

Illinois Coal - 25$/ton at 12,235 Btu/lb. HHV (1.02 $ per MM Btu)

Coal - 22,100 TPD

Overall Efficiency - 58.2% HHV

Liquid Methanol Product - 315,000 MMBtu/day HHV

Total Capital Requirement - 1,925 million dollars

Operating Factor - 90%, 328.5 days/year

	First Year Cost $/MMBtu	Levelized Cost $/MMBtu
Coal	1.75	1.75
Water, Catalyst & Chemicals	0.07	0.07
Other Operating Costs	0.98	0.98
Total Operating Cost	2.80	2.80
Capital Charges	3.77	2.90
Production Cost	6.57 HHV	5.70 HHV
	7.48 LHV	6.49 LHV

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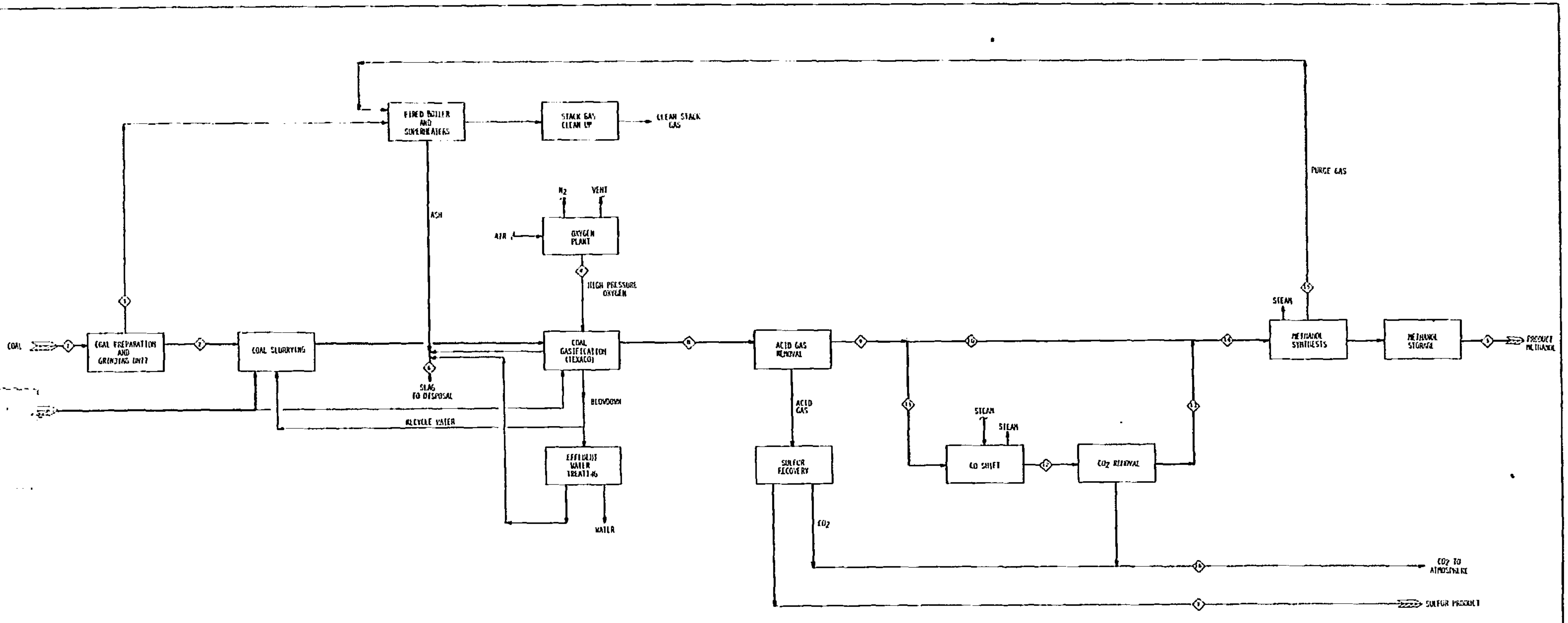
*Calculated in accordance with EPRI's Economic Premises, Feb. 11, 1977 - utility financing.



	① TOTAL COAL	② GASIFIER COAL	③ UTILITY COAL	④ OXYGEN	⑤ METHANOL PRODUCT	⑥ TOTAL SLAG-ASH	⑦ SULFUR PRODUCT
COAL, TPD	22,300	20,282	1 818	-	-		-
OXYGEN, TPD (100%)	-	-		16,423			-
METHANOL TPD	-	-	-	-	16,352	-	-
SLAG ASH, TPD	-		-	-	-	2,122	-
SULFUR, TPD	-	-	-	-	-	-	250

	⑧ RAW GAS	⑨ CLEAN GAS	⑩ SHIFT CO2 BYPASS	⑪ SHIFT FEED	⑫ SHIFT EFFLUENT	⑬ CO2 EFFLUENT	⑭ METHANOL FEED	⑮ PURGE GAS	⑯ CO2 VENT
MOL %									
H_2	35.06	39.57	39.57	39.57	58.69	85.59	65.68	49.45	-
CO	52.35	59.03	59.03	59.03	8.73	12.61	32.82	21.85	-
CO_2	10.12	0.02	0.02	0.02	31.44	0.28	0.17	1.67	100
CH_4	0.10	0.11	0.11	0.11	0.08	0.51	0.15	2.35	-
N_2+A	1.07	1.28	1.28	1.28	0.83	1.21	1.21	23.28	-
H_2S+COS	1.33	-	-		-	-	-	-	-
CH_3OH	-		-	-	-	-		1.32	-
TOTAL (DRY)	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100
MMSCFD (DRY)	1363	1207	684	523	1001	486	1107	56	453
PSIG	600	575	575	575	495	433	440	450	0

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CASE 4 - TEXACO GASIFICATION
CHEM SYSTEMS METHANOL
PROCESS BLOCK FLOW DIAGRAM

THE RALPH M PARSONS COMPANY
PASADENA, CALIFORNIA

JOB NO 5604-1
DWG NO 5604-FS-4-A
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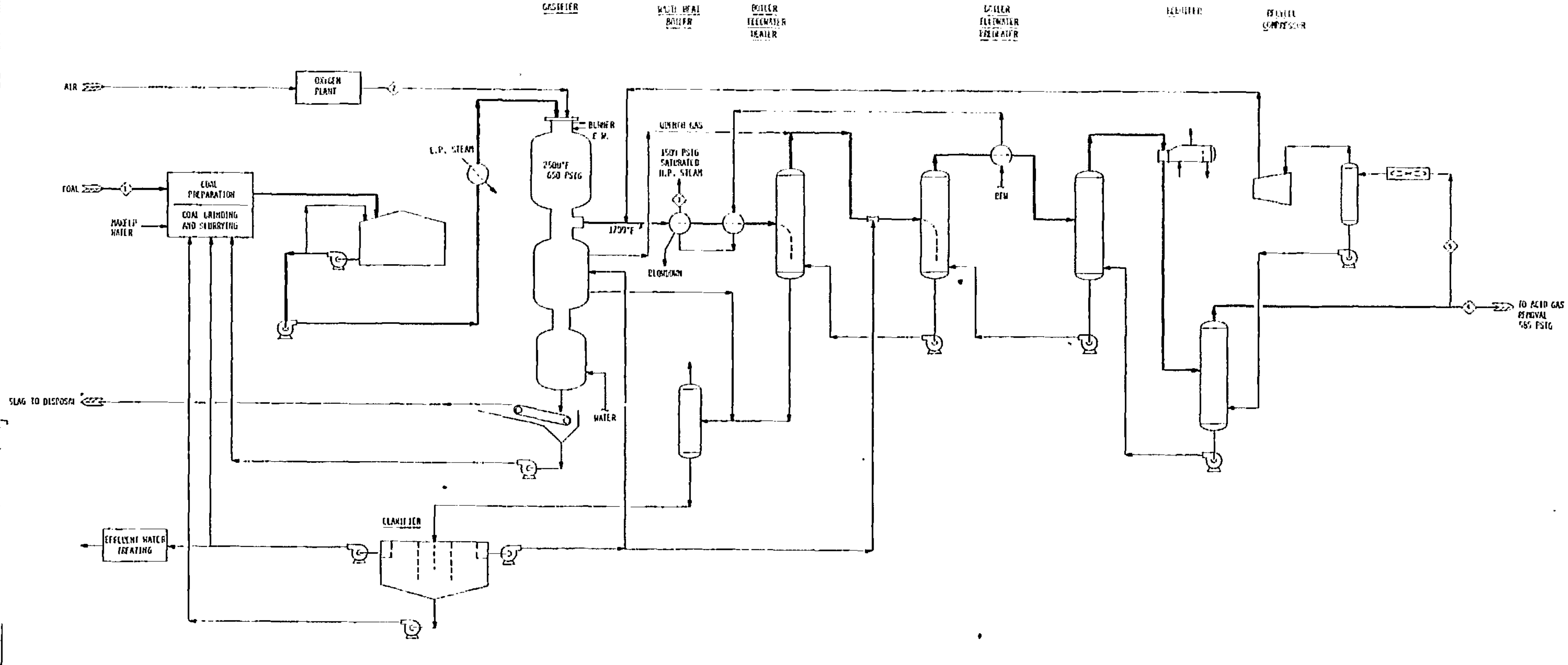

	① COAL	② OXYGEN	③ STEAM GENERATORS
COAL, TPD	20,783	-	-
OXYGEN, TPD (100%)	-	16,423	-
STEAM 10^6 LB/HR	-	-	3.22

	④ RAW GAS	⑤ RECYCLE GAS
MOL %		
H_2	35.06	35.06
CO	52.35	52.35
CO_2	10.12	10.12
CH_4	0.10	0.10
N_2+A	1.07	1.07
H_2S+COS	1.33	1.33
	100.03	100.03
MMSCFD (DRY)	1363	318

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PROCESS FLOW DIAGRAM
GASIFICATION SECTION
CASE 4 - TEXACO GASIFICATION

THE RALPH M PARSONS COMPANY
PASADENA, CALIFORNIA

JOB NO 5604-1
DWG NO 5604-FS-4-B
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SECTION 10

CASE 5

BRITISH GAS COUNCIL/LURGI SLAGGING GASIFICATION/FISCHER-TROPSCH SYNTHESIS

The overall plant is shown on Process Flow Diagram 5604-1-FS-5-A. A total of 22,918 tons/day of coal are used, all being gasified. Fuel gas recovered from the process is used in the utility plant and excess is export fuel product.

Plant thermal efficiency is as follows:

Input

Coal

22,918 TPD x 2,000 lb/T x 12,235 Btu/lb. = 560,803 MMBtu/day (HHV)

Output

Fischer-Tropsch liquids

5,573 TPD x 2,000 lb/T x 20,055 Btu/lb. = 223,545 MMBtu/day (HHV)

Excess fuel gas

106.62 MMSCFD x 850 Btu/scf = 90,634 MMBtu/day (HHV)

Total = 314,179 MMBtu/day (HHV)

$$\text{Efficiency} = \frac{314{,}179}{560{,}803} \times 100 = \underline{\underline{56.0\%}} \quad \text{(HHV)}$$

$$\text{Efficiency} = \frac{207{,}238 + 82{,}780}{536{,}693} \times 100 = \underline{\underline{54.0\%}} \quad \text{(LHV)}$$

The bulk of this plant design is exactly the same as Case 2. The same coal rate to gasification, synthesis gas production and clean-up section design are used.

The methanol synthesis of Case 2 is replaced by Fischer-Tropsch synthesis for Case 5; the thermal yield of Case 5 is somewhat less than Case 2. Of the total output 98.6% is as liquid in the methanol Case 2 but only 71.2% as liquid in the F-T Case 5.

Coal Preparation, Gasification, Tar Separation, Phenol Extraction, Acid Gas Removal, CO Shift, CO_2 Removal, Compression and Cryogenic Separation.

These sections are all similar to Case 2, see Drawings 5604-FS-2-B, 5604-FS-2-C and 5604-FS-2-D.

The capacities and flows are all the same; the difference is that syngas is required at 400 psi for Fischer-Tropsch synthesis instead of 500 psi for methanol synthesis.

Fischer-Tropsch Synthesis

Refer to Flow Diagram 5604-FS-5-B.

The Fischer-Tropsch synthesis as depicted is an adaption of the so-called ARGE synthesis (Ruhr-Chemie/Lurgi) as operated by Sasol for many years in South Africa. The feed gas is heated by exchange with effluent, passed over zinc oxide guard for sulfur removal, and mixed with recycle gas which is also heated by exchange with reactor effluent. The gas enters the top of the "fixed bed" reactor and passes over iron catalyst which is inside tubes. Outside the tubes, steam is produced at about 300 psi. The reactor is essentially isothermal; the effluent gas is used in heat exchange with the inlet streams before entering the condenser system.

Various hydrocarbons are withdrawn from the equipment in successive stages. First, some wax is drawn off the reactor bottom then hot condensate off the gas exchangers and finally, cold condensate off the coolers.

Caustic is circulated round the coolers and an aqueous phase is separated which contains alcohols and neutralized acids.

At Sasol, a very complicated fractionation and chemicals separation plant is used but for a fuels plant only stabilization and separation of aqueous phases would be required.

For the purpose of this study, data from Sasol was used and the following yield structure assumed. Liquid fuel product was taken as C_4 and heavier.

Yield Breakdown:

		Wt %	% Olefins in Cut
	(C_1	7.8	-
gases 17.1 wt %	(C_2	3.2	23
	(C_3	6.1	64
	(C_4	4.9	51
	(C_5-C_{11}	24.8	50
liquid 82.9 wt %	(C_{12}-C_{20}	14.7	40
	(C_{20}	36.2	15
	(alcohols, ketones	2.3	-
	(acids	-	
		100	

The conversion was estimated at 95% based on H_2 feed. This compares with 97% conversion for methanol. The 95% conversion is higher than the 65% conversion obtained at Sasol. This latter low yield is mainly due to feeding the synthesis with a high methane gas (14% CH_4). At Sasol, this results in a large purge which is used as town gas in the local area. With the low methane feed (0.6% CH_4) used in this design, higher conversions are obtained while maintaining the same "inert" CH_4 level in the purge stream (about 28%). This design is therefore illustrative only for the purpose of a screening comparison with methanol and is not based on a licensor design.

Approximate Arrangement

The plant consists of coal handling for sizing coal (no pulverizing) to feed 14 British Gas Council-Lurgi, slagging gasifiers with 5 trains of tar oil separation and 2 trains of Phenolsolvan. The gas is processed in 4 trains of acid gas removal, 4 shift trains and 7 trains of CO_2 removal. After compression and several trains of cryogenic separation, there are several trains of Fischer-Tropsch synthesis, with multiple reactors, totalling about 80 reactors of the Sasol fixed-bed type. Two sulfur plants are included.

Utility System

The plant is balanced so that the generated steam is sufficient for the steam turbines for plant compressors, power generation and other steam requirements. A total of 3.5×10^6 lb/hr. of steam at 1500 psig, 950°F is produced.

Of the total steam about 17% is produced (saturated) in the shift section and the balance, 83%, is produced in the utility boiler/superheater system which also superheats the shift steam.

The HP steam is used in back pressure and condensing turbines to drive the main plant turbines and to provide 0.6×10^6 lb/hr. of gasifier process steam at 400 psig.

The large turbines are:

	HP
Synthesis Gas and Cryogenic Compressor	243,000
Oxygen Plant Compressors	258,000
Fischer-Tropsch Recycle Compressor	33,000
Power Generators	16,000
Total	550,000

Steam from the Fischer-Tropsch synthesis, 2.45×10^6 lb/hr., produced 285 psig saturated, is used partly as shift process steam and partly, after superheating, in a power generator (40,000 HP). Total power generated is 56,000 HP.

Steam from back pressure turbines at 50 psig and 100 psig is used in the acid gas removal reboilers, Phenolsolvan unit, Fischer-Tropsch section and other miscellaneous users.

Condensate is recovered from vacuum condensers, steam reboilers and shift condensate and returned to the deaerator.

Make-up boiler water required is 1.4×10^6 lb/hr., corresponding to steam converted in gasification, shift sections and losses.

CAPITAL SUMMARY - CASE 5

	MM $
Coal Preparation	28
Gasification	101
Tar Recovery & Phenol Extration	69
Acid Gas Removal	21
CO Shift	39
CO_2 Removal	147
Oxygen	183
Cryogenic Clean-up and Boxes	21
Fischer-Tropsch Synthesis	235
Final Desulfurization - Zinc Oxide	10
Raw Gas Compression	48
Sulfur Recovery	20
Utility Boilers, Superheaters and Power Generators	49
	971
Utilities and General Offsites at 15%	146
Total Plant Cost, Including Engineering	1,117
Contingency at 15%	168
Total Plant Investment	1,285
Interest on a Construction Loan	286
Royalty Allowance	6
Startup and Working Capital	183
Total Capital Requirement	1,760 MM $

PRODUCTION COST* - CASE 5

Illinois Coal	-	25$/ton at 12,235 Btu/lb. HHV (1.02 $ per MM Btu)
Coal	-	22,918 TPD
Overall Efficiency	-	56.0% HHV
Liquid Fischer-Tropsch product	-	223,545 MMBtu/day HHV
Fuel Gas Product	-	90,634 MMBtu/day HHV
Total Product	-	314,179 MMBtu/day HHV
Total Capital Requirement	-	1,760 million dollars
Operating Factor	-	90%, 328.5 days/year

	First Year Cost $/MMBtu	Levelized Cost $/MMBtu
Coal	1.82	1.82
Water, Catalyst & Chemicals	0.07	0.07
Other Operating Costs	0.89	0.89
Total Operating Cost	2.78	2.78
Capital Charges	3.46	2.66
Production Cost	6.24 HHV	5.44 HHV
	6.76 LHV	5.89 LHV

*Calculated in accordance with EPRI's Economic Premises, Feb. 11, 1977 - utility financing.

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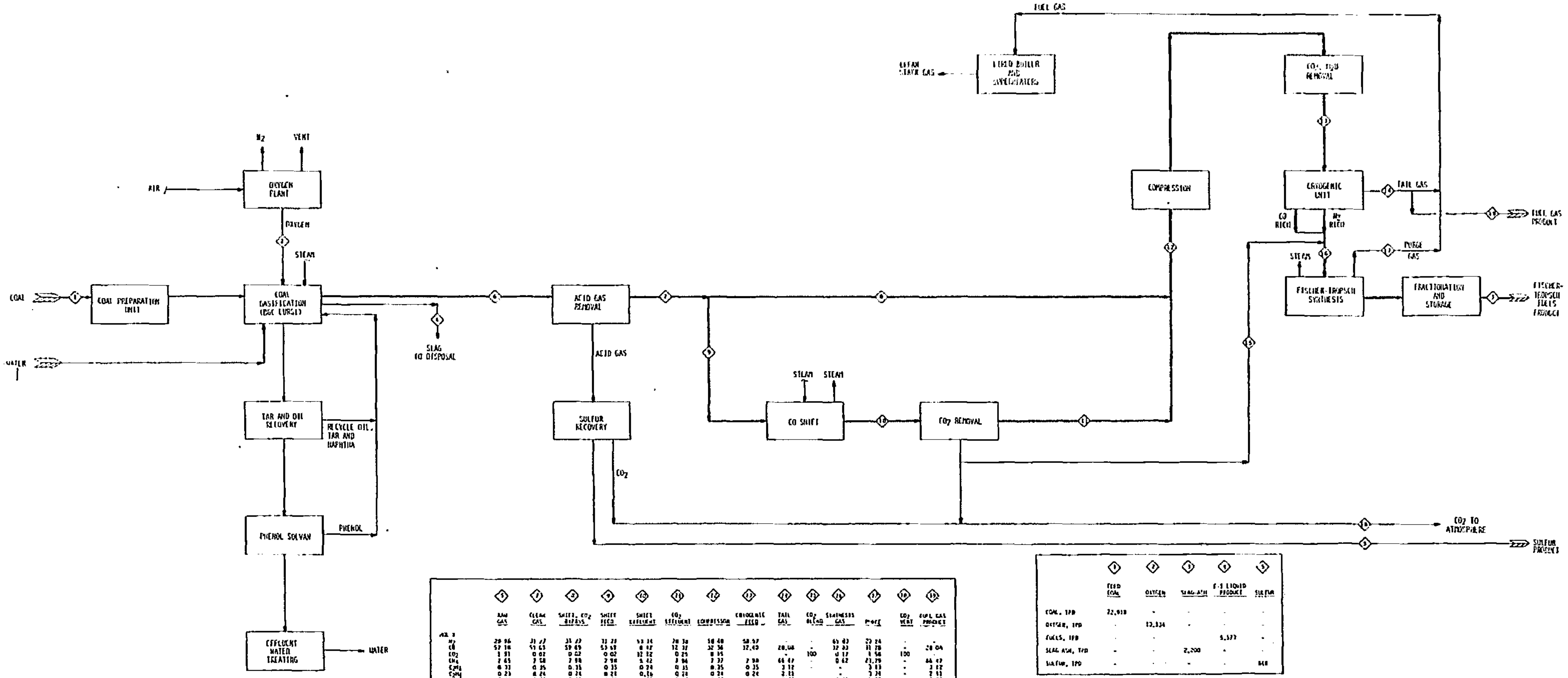

	6	7	8	9	10	11	12	13	14	15	16	17	18	19
	RAW GAS	CLEAN GAS	SHIFT, CO2 BYPASS	SHIFT FEED	SHIFT EFFLUENT	CO2 EFFLUENT	COMPRESSOR	CRYOGENIC FEED	TAIL GAS	CO2 BLEED	SYNTHESIS GAS	PURGE	CO2 VENT	FUEL GAS PRODUCT
MOL %														
H_2	28.96	31.27	31.27	31.27	53.16	78.38	58.40	58.57	-	-	65.83	23.24	-	-
CO	57.18	51.63	59.65	55.61	0.42	12.32	32.36	32.40	28.06	-	33.33	11.78	-	28.06
CO_2	3.91	0.02	0.02	0.02	32.32	0.25	0.35		-	100	0.17	1.56	100	-
CH_4	7.65	7.58	7.98	7.98	5.42	7.94	7.37	7.98	66.47	-	0.62	41.79	-	66.47
C_2H_6	0.33	0.35	0.35	0.35	0.24	0.35	0.35	0.35	3.12	-	-	3.83	-	3.12
C_2H_4	0.23	0.24	0.24	0.21	0.16	0.24	0.24	0.24	2.13		-	3.34	-	2.11
N_2+A	0.44	0.44	0.44	0.46	0.31	0.44	0.66	0.46	0.23	-	0.40	4.35	-	0.23
H_2S+COS	1.42	-	-	-	-	-	-	-	-	-	-			-
NH_3	0.03	-	-	-	-	-	-	-	-	-	-		-	-
C_3/C_3+	-	-	-	-	-	-	-	-	-	-	-	5.36	-	-
TOTAL (DRY)	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100	100.00	100.00	100	100.00
MMSCFD (DRY)	1416	1356	374	982	1353	384	1370	1356	152	2	1297	136	395	107
PSIA	225	225	270	270	270	310	310	600	50	330	330	345	0	50

	1	2	3	4	5
	FEED COAL	OXYGEN	SLAG-ASH	F-T LIQUID PRODUCT	SULFUR
COAL, TPD	22,910	-	-	-	-
OXYGEN, TPD	-	13,334	-	-	-
FUELS, TPD	-	-		5,577	-
SLAG ASH, TPD	-	-	2,200	-	-
SULFUR, TPD	-	-	-	-	848

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CASE 5 - BRITISH GAS COUNCIL/LURGI SLAGGING
GASIFICATION, FISCHER-TROPSCH SYNTHESIS
PROCESS BLOCK FLOW DIAGRAM

THE RALPH M PARSONS
COMPANY
PASADENA, CALIFORNIA

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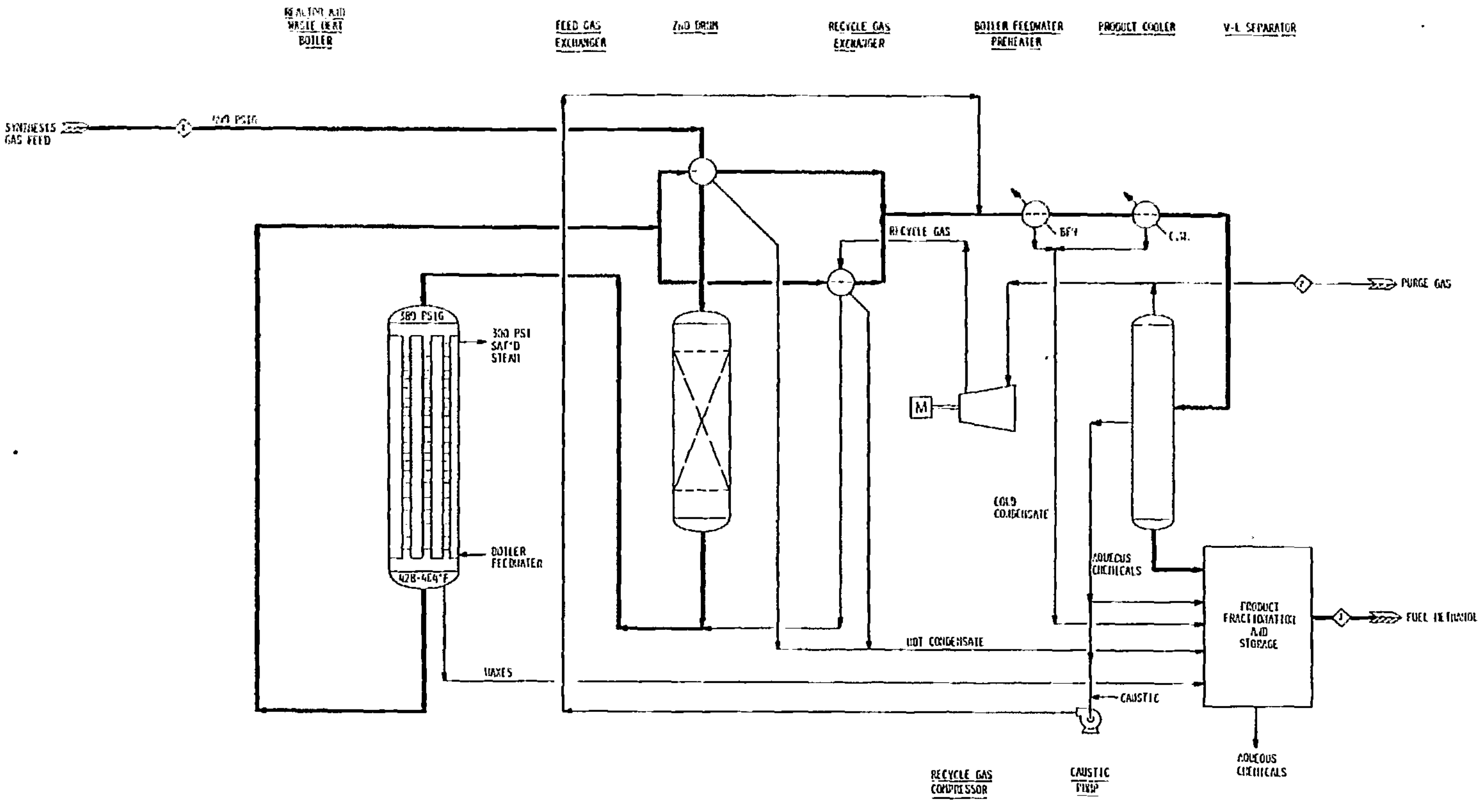

MOL %	① SYNTHESIS GAS	② PURGE
H_2	65.83	29.24
CO	32.90	31.30
CO_2	0.17	1.50
CH_4	0.62	23.79
C_2H_6	-	3.13
C_2H_4	-	0.94
C_3H_8	-	1.89
C_3H_6	-	3.48
N_2+A	0.49	4.15
TOTAL (DRY)	100.00	100.00
MMSCFD (DRY)	1207	1.16

WT %	③ TOTAL LIQUID
C_4	5.91
C_5-C_{11}	29.92
$C_{12}-C_{20}$	17.73
C_{20}	43.67
ALC., KET	2.77
TOTAL (DRY)	100.00
TPD	5573
TOTAL °API	58.7
BPD	42450

ELECTRIC POWER RESEARCH INSTITUTE
PROCESS FLOW DIAGRAM
FISCHER-TROPSCH SYNTHESIS

THE RALPH M PARSONS COMPANY
PASADENA, CALIFORNIA

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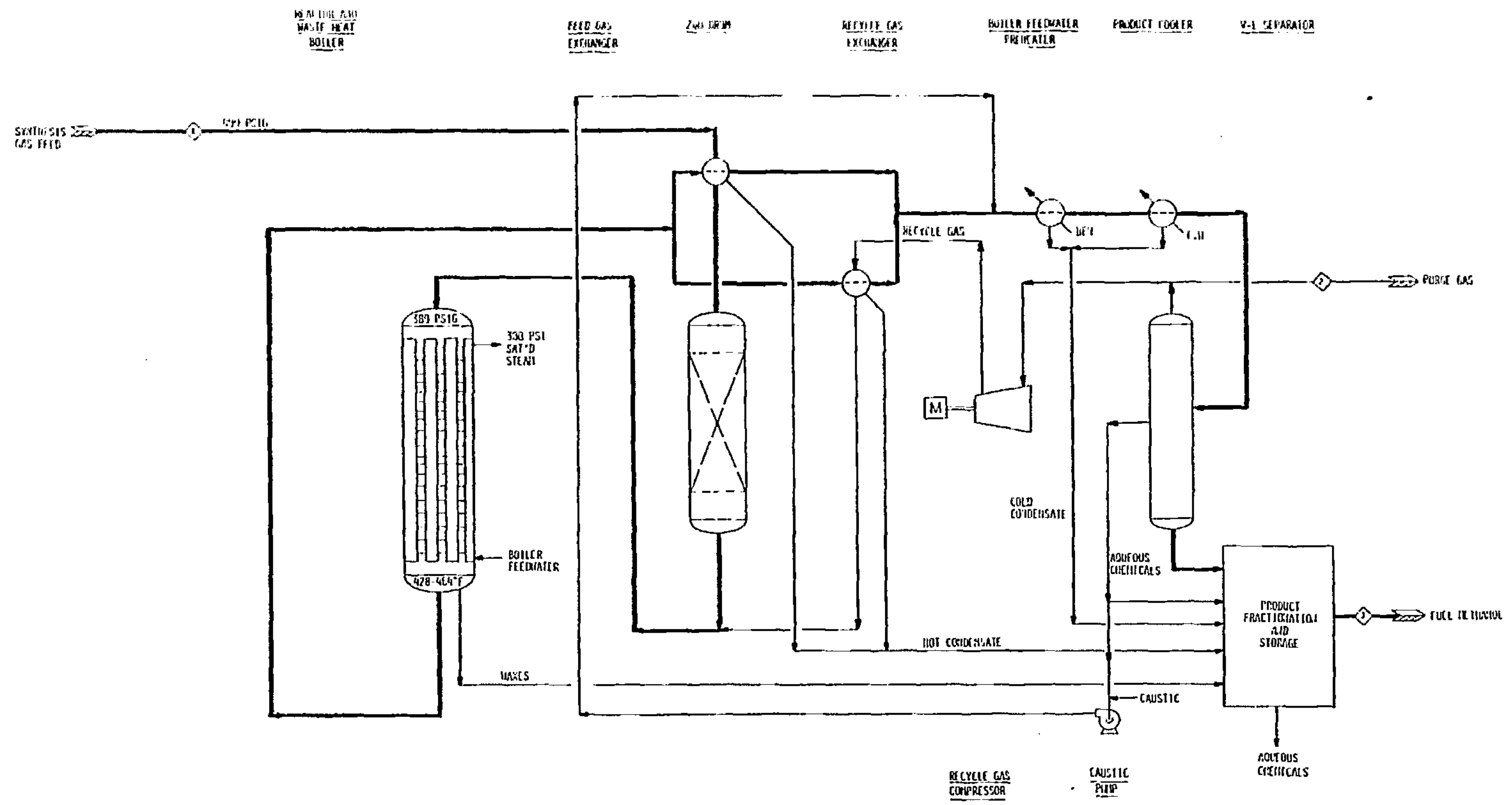

	① SYNTHESIS GAS	② PURGE
MOL %		
H_2	45.83	29.24
CO	52.95	31.30
CO_2	0.13	1.50
CH_4	0.62	23.79
C_2H_6	-	3.13
C_2H_4	-	0.94
C_3H_8		1.88
C_3H_6	-	3.48
N_2+A	0.47	4.35
TOTAL (DRY)	100.00	100.00
MMSCFD (DRY)	1207	136

	③ TOTAL LIQUID
WT %	
C_4	5.93
C_5-C_{11}	29.92
C_{12}-C_{23}	17.73
C_{24}	43.63
ALC., ETC.	2.77
TOTAL (DRY)	100.00
TPD	5573
TOTAL °API	59.7
BPD	42850

ELECTRIC POWER RESEARCH INSTITUTE
PROCESS FLOW DIAGRAM
FISCHER-TROPSCH SYNTHESIS

THE RALPH M. PARSONS COMPANY
PASADENA, CALIFORNIA

JOB NO. 5604 1
DWG NO. 5604-FS-5-B
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SECTION 11

COMPARATIVE ECONOMICS

The comparative feed and product quantities and thermal efficiencies are as follows:

	Case 1	Case 2	Case 3	Case 4	Case 5
	Foster-Wheeler Gasification/ Chem Systems Methanol	British Gas Council/Lurgi Slagger/Chem Systems Methanol	Koppers-Totzek Gasification/Chem Systems Methanol	Texaco Gasification/ Chem Systems Methanol	British Ga Council/ Lurgi Slagger/ Fischer-Tropsch Synthesis
Coal					
TPD-gasified	24,566	22,918	20,702	20,282	22,918
TPD-utility	-	-	3,872	1,818	-
Coal TPD-total	24,566	22,918	24,574	22,100	22,918
Thermal Output (HHV) MMBtu/day -					
liquid	315,000	315,000	315,000	315,000	223,545
fuel gas	18,444	4,500	-	-	90,634
MMBtu/day - total	333,444	319,500	315,000	315,000	314,179
Efficiency (HHV)	55.5	57.0	52.4	58.2	56.0
(LHV)	51.0	52.3	48.1	53.5	54.0

The following should be noted:

1. Koppers-Totzek's design is based on 95% carbon conversion in the gasifier whereas the other gasification processes were assumed to have close to 100% carbon conversion.

2. Although Foster-Wheeler and British Gas Council-Lurgi Slagger show higher gasification efficiencies (mainly because of high CH_4 content in gas), the advantage is lost in the overall efficiency because of the necessity of removing this CH_4 before synthesis. This involves an additional energy requirement (compression) for separation and results in the CH_4 being used for most of the plant fuel as opposed to mainly coal firing as in Cases 3 and 4. The high pressure, low inert Texaco gas is best suited for synthesis.

3. The Foster-Wheeler raw gas analysis is considered "conceptual" in that no hydrocarbons higher than C_2 are shown at 1700°F.

4. The British Gas Council-Lurgi Slagger design assumes total recycle to extinction (in the gasifier) of tars, oils, naphthas and phenols at relatively low oxygen consumption.

The comparative capital estimates are as follows, all in millions of dollars:

	Case 1	Case 2	Case 3	Case 4	Case 5
	Foster-Wheeler Gasification/ Chem Systems Methanol	British Gas Council/Lurgi Slagger/Chem Systems Methanol	Koppers-Totzek Gasification/ Chem Systems Methanol	Texaco Gasification/ Chem Systems Methanol	British Gas Council/Lurgi Slagger/ Fischer-Tropsch Synthesis
Coal Preparation & Grinding	60	28	60	50	25
Gasification	134	101	390	265	101
Tar & Phenol Recovery	-	69	-	-	69
Acid Gas Removal	114	21	54	70	21
CO Shift	30	39	35	35	39
CO_2 Removal	88	147	129	163	147
Oxygen	250	185	258	256	185
Methanol Synthesis	122	122	122	122	-
Fischer-Tropsch Synthesis	-	-	-	-	233
Final Desulfurization, Zinc Oxide	10	10	10	10	10
Raw Gas & Tail Gas Compression	51	55	86	-	45
Sulfur Recovery	20	20	20	20	20
Cryogenic Cleanup & Separation	21	21	-	-	21
Utility Boilers, Super heaters, Power Generators & Stack Gas Cleanup	49	56	135	75	49
	949	869	1,300	1,066	971
Utilities & General Offsites at 15%	142	130	195	160	146
TOTAL Constructed Including Engineering	1,091	1,000	1,495	1,226	1,117
Contingency	164	150	224	184	168
TOTAL Plant Investment	1,255	1,150	1,719	1,410	1,285
Interest on Construction Loan	279	255	382	313	285
Royalty Allowance	6	6	9	7	7
Startup & Working Capital	183	169	232	195	183
TOTAL Capital Requirement	1,723	1,580	2,342	1,925	1,760

The following should be noted:

1. Important

   All the above capital estimates are "curve-type" estimates. The breakdown is intended to illustrate relative costs and trends and not meant for detailed comparison.

2. Relative accuracy of cost estimates with consideration to source of data and status of development of the gasification processes are judged to be as follows, in order with greater accuracy first - Cases 3, 4, 2, 1. Additionally, Case 2 - methanol is considered to be of better accuracy than Case 5 - Fischer-Tropsch.

The comparative production costs* are as follows, all in $/MMBtu:

	Case 1		Case 2		Case 3		Case 4		Case 5	
	Foster-Wheeler Gasification/ Chem Systems Methanol		British Gas Council/Lurgi Slagger/Chem Systems Methanol		Koppers-Totzek Gasification/ Chem Systems Methanol		Texaco Gasification/ Chem Systems Methanol		British Gas Council/Lurgi Slagger/ Fischer-Tropsch Synthesis	
	First Year	Levelized	First Year	Levelized	First Year	Levelized	First Year	Levelized	First Year	Levelized
Coal	1.84	1.84	1.79	1.79	1.95	1.95	1.75	1.75	1.82	1.82
Water, Catalyst & Chemicals	0.07	0.07	0.07	0.07	0.08	0.08	0.07	0.07	0.07	0.07
Other Operating Costs	0.82	0.82	0.79	0.79	1.19	1.19	0.98	0.98	0.89	0.89
TOTAL Operating Cost	2.73	2.73	2.65	2.65	3.22	3.22	2.80	2.80	2.78	2.78
Capital Charges	3.19	2.45	3.05	2.35	4.59	3.53	3.77	2.90	3.46	2.66
TOTAL Production Cost -HHV	5.92	5.18	5.70	5.00	7.81	6.75	6.57	5.70	6.24	5.44
-LHV	6.73	5.88	6.49	5.69	8.89	7.69	7.48	6.49	6.76	5.89

*Calculated in accordance with EPRI's Economic Premises, Feb. 11, 1977 - utility financing.

SECTION 12

EFFECT OF PRESSURE ON METHANOL SYNTHESIS

In preliminary studies, it was concluded (based on data from previous studies for a 1300 TPD loop) that for synthesis gas produced at 500 psi, a stand-off in methanol production cost resulted when methanol synthesis is practiced at 500 psi or 1100 psi. This resulted because the savings in capital at the higher pressure were offset by the additional energy required to compress the syngas. Hence, 500 psi synthesis was used for Cases 1, 2, 3 and 4.

For this present study, in addition to the 500 psi design, Chem Systems provided a design for a 1000 psi case. This data was used to further check the above. The evaluation is as follows.

The higher pressure case showed somewhat more total energy recovered compared with the lower pressure:

	480 psia Synthesis %	1015 psia Synthesis %
Energy in Make-up Synthesis Gas	100	100
Methanol Product	80.4	81.5
Purge Gas	5.1	3.9
	85.5	85.3
Waste Heat Boiler	10.2	10.6
Boiler Feedwater	0.0	1.0
	10.2	11.6
Total Heat Recovered	95.7	96.9
Lost to Cooler	4.3	3.4
Total	100.1	100.3

The higher pressure case required less recycle gas and recycle fuel energy but more total energy when make-up gas compression from 480 to 1015 psia is included:

	480 psia Synthesis	1015 psia Synthesis
Pumps	13,600	5,200
Recycle Compressor	23,300	4,100
	36,900	9,300
Make-up Compressor	nil	63,400
Total	36,900 HP	72,700 HP

The capital cost of the higher pressures synthesis is less than the lower pressure.

	480 psia Synthesis	1015 psia Synthesis
Capital Cost of Synthesis	122	99
Capital Cost of Make-up Compressor	-	12
Total	122 Million $	111 Million $
Additional Boilers, Superheaters & Offsites	-	3
		114 Million $

For the high pressure case, extra coal must be burned in the boiler and superheater to raise additional steam to service the added compressor duty. However, additional steam is recovered in the loop. The net result is a higher coal cost of 1.3 million dollars per year.

The net savings in capital is 8 million dollars. This is considered to be a stand-off with the higher operating cost since, with capital charges at 15.6%, the savings is 1.3 million dollars per year. The effect of these differences on production cost is very small. It is concluded that with gasification at the 600 psi level, there is no significant different for Chem Systems methanol process if the synthesis is at 500 psi or 1000 psi.

Of course, if the gasification process is at a higher pressure, say 1200 psi, then there is advantage in having the synthesis at 1000-1100 psi. For instance, for Case 4 Texaco, the gasifier was at 650 psi and the synthesis at 500 psi. If the gasifier were at 1200 psi, synthesis could be 1050 psia. Although this case has not been worked in detail, the savings are still expected to be rather modest compared with Case 4. Some savings are expected in the gasifier section as well as the methanol synthesis. However, the oxygen plant is more expensive because of more O_2 compressor and overall efficiency is slightly less than Case 4 because of the extra oxygen power required. It is not anticipated that substantial savings would be made by going to pressures above the present 650 psi gasification although some improvement is expected. This improvement is expected to be more for ICI or Lurgi synthesis than for Chem Systems because of higher recycle gas requirements.

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