MASTER

COAL LIQUEFACTION PROCESS RESEARCH MOBIL M-GASOLINE PROCESS

R & D INTERIM REPORT NO. 1

work performed by
THE RALPH M. PARSONS COMPANY
Pasadena, California 91124
Subcontract No. 7186

for the
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CONTROL AND SUPPORT
U.S. DEPARTMENT OF ENERGY
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June 1978

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ABBREVIATIONS

BFW	boiler feedwater
BPD	barrels per day
BPH	barrels per hour
BPSD	barrels per stream day
DCF	discounted cash flow
DME	dimethyl ether
FCI	fixed capital investment
gpd	gallons per day
gpm	gallons per minute
HHV	higher heating value
LPG	liquified petroleum gas
MM	million
RPSP	required product selling price
SNG	substitute natural gas
TPD	tons per day

SECTION 1
INTRODUCTION

SECTION 1

INTRODUCTION

This report summarizes the projected economic potential for the methanol-to-gasoline (M-Gasoline) process under development by Mobil Research and Development Corporation. The potential was reviewed by developing a preliminary economic estimation of the design, construction, and operation of a commercial-sized facility to produce 50,000 barrels per day (BPD) of gasoline.

The review included a preliminary definition of facilities and projected economics for the production of methanol utilizing synthesis gas produced by the gasification of coal. The limited time and effort allotted for completion of the study permitted moderate, but not exhaustive, optimization of the conceptual industrial complex. A summary of more recent data was received following completion of the study. Preliminary analysis of this more recent information indicated that it would not significantly alter the preliminary economic conclusions. At the direction of the Department of Energy, therefore, the conceptual design and economic evaluations were not revised to incorporate this later information.

The work was performed under the direction of the Fossil Energy Office of Program Planning and the analysis of the Energy Research and Development Administration (ERDA), now a part of the Department of Energy.

The primary objective of the study was to project the economic potential of the process. A secondary objective was to develop recommendations for improvements.

The review was based on information contained in a report on methanol production entitled "Economic Feasibility Study, Fuel Grade Methanol from Coal"¹ prepared for ERDA under Contract No. 76-18 and on Mobil Research and Development Corporation's report entitled "Development Studies on Conversion of Methanol and Related Oxygenates to Gasoline"² prepared for ERDA under Contract E(49-18)-1773.

Preliminary engineering was performed as required to define the basis for plant design and economic estimates.

SECTION 2
SUMMARY

SECTION 2

SUMMARY

A preliminary review was made of the Mobil M-Gasoline process for the conversion of methanol to gasoline using crude methanol derived from coal. A plant to produce 50,000 barrels of product gasoline was selected as being representative of present commercial plant sizes. Key process steps included (1) coal gasification to produce synthesis gas, (2) conversion of synthesis gas to crude methanol, and (3) conversion of crude methanol to gasoline.

The projected required gasoline selling price is about 82¢ per gallon based on the use of purchased coal at \$20 per ton, a 12% discounted cash flow (DCF) rate of return with a 65/35 debt-to-equity ratio, 9% interest rate, and a fixed capital investment based on mid-1977 dollars. The estimated fixed capital investment is about \$1.6 billion. All economics in this report are based on mid-1977 dollars.

Preliminary engineering was performed as required to obtain the fixed capital investment estimates.

About 76% (62.4¢ per gallon) of the product cost is attributable to synthesis gas generation, including coal costs. The methanol synthesis and methanol conversion to gasoline steps each contribute about 12% to the product cost.

The coal cost at \$20 per ton accounts for about 30% of the required gasoline selling price, which amounts to approximately 25¢ per gallon of gasoline.

Sensitivity analyses are presented that define the effects of various financing methods, DCF rates, cost of coal, and the fixed capital investment on required product selling or transfer price; these were performed for each of the three major plant sections and for the overall plant. The results indicate that:

- The required product selling price is very sensitive to the fixed capital investment (FCI). An increase of 50% in the FCI increases the gasoline cost about 32% at the base case of a 65/35 debt-to-equity ratio and a 12% DCF rate of return.
- The method of financing can significantly affect the required selling price of the gasoline. A facility that is 100% equity financed and that requires a 16% rate of return has a gasoline selling price about 46% higher than a project with 65/35 debt-to-equity financing.
- A 50% reduction in the price of feed coal (from \$20.00 to \$10.00 per ton) reduces the required gasoline selling price by approximately 15%.

The basic process design and investment estimates for the production of crude methanol from coal are based on information contained in a report prepared for ERDA by E.I. DuPont de Nemours.¹

The methanol-to-gasoline conversion design is based on information presented in a report prepared for ERDA by Mobil Research and Development Corporation.²

Recommendations for process improvements with potential for a reduction of capital investments and operating cost are presented. The prime target for significant cost improvement lies in the reduction in cost for production of synthesis gas. The potential improvement areas in synthesis gas production include lower oxygen consumption, higher coal-to-gas thermal efficiencies, and larger coal gasification reactors using advanced coal gasification processes.

SECTION 3
DESIGN AND EVALUATION BASIS

SECTION 3

DESIGN AND EVALUATION BASIS

The criteria used to prepare the design for review of the production of gasoline using Mobil's methanol-to-gasoline conversion process and the production of methanol from coal are discussed in this section.

3.1 PLANT CAPACITY

The production of gasoline is to be representative of the output of a medium- to large-sized oil refinery. A production rate of 50,000 BPD of gasoline was selected. This gasoline production rate is equivalent to a crude oil feed rate of 75,000 to 100,000 BPD and is considered representative of many refineries presently in operation in the United States.

3.2 PLANT PARAMETERS

The plant is to be a grassroots facility using coal, water, and air as raw materials. All required utilities, such as steam and power, are to be generated within the plant complex.

All coal is to be purchased and delivered to the plant by railroad car. No coal mining or preparation is included in the study.

Required makeup water is assumed to be available from a nearby river.

3.3 PRODUCT SLATE

The primary product is to be gasoline. Liquefied petroleum gas (LPG), which consists of propanes and butanes, will be produced as a byproduct. The gasoline will be produced using Mobil's M-Gasoline Process combined with alkylation of the unsaturated gases.

3.4 PROCESS STATUS

The production of methanol from coal is to be based on processes in either advanced development or commercially proven status. Facilities include presently available plant and equipment sizes. In this regard, the Texaco partial oxidation process is considered a candidate for commercial coal gasification.

The conversion of methanol to gasoline is based on Mobil's preliminary design for a unit producing 10,000 BPD of gasoline using Mobil's fixed bed catalytic process. This process is briefly described in reference 2.

3.5 ENVIRONMENTAL CONSIDERATIONS

All plant effluents, such as gaseous, liquid, and solid waste streams, are to meet local and Federal regulations. Facilities are to be provided to ensure compliance with these regulations. Solid wastes, such as coal ash, are to be delivered by railroad car to a selected site for disposal.

3.6 METHANOL SYNTHESIS

A review was made of available information on the subject of production of methanol utilizing synthesis gas derived by coal gasification. The material presented in an ERDA report¹ was selected as the basis for the methanol production process.

This selection was based on the following factors:

- Sufficient information was available or could be derived to scale-up the process to produce the required methanol quantity.
- Commercial-sized units were utilized in the study.
- The study was based on the production of a nominal 5,500 tons per day (TPD) of fuel grade (95%) methanol. The methanol requirements for the production of 50,000 BPD of gasoline require the equivalent of approximately 17,250 TPD of fuel grade methanol (19,670 TPD of crude methanol). Thus, six trains will be utilized in the large plant, rather than the two trains utilized in the reference study, with an individual train scale-up of only 5% (2,875 tons per train compared to 2,750 tons per train).

3.7 CHANGES TO METHANOL STUDY

The design used in the methanol from coal study¹ was kept intact to the maximum possible extent for this evaluation. Certain relatively minor changes were made to adapt their design to the needs of this study. These changes include the following:

- The methanol fractionation facilities were deleted. The Mobil methanol-to-gasoline process uses crude (83 to 85%) methanol. The surplus steam resulting from deletion of the fractionation unit was used in the Mobil M-Gasoline unit and for driving the cooling water circulation pumps.
- The methanol storage and loading facilities were deleted. Intermediate 10-day crude methanol surge storage was provided.

3.8 MOBIL M-GASOLINE PROCESS

The information presented in reference 2 was used for the conversion of methanol to gasoline. After completion of the report, additional information³ was received from Mobil. Preliminary analysis of this more recent information

indicated that it would not significantly alter the preliminary economic conclusions at the direction of the Department of Energy; therefore, the conceptual design and economic evaluation were not revised to incorporate this later information.

SECTION 4
PROCESS DESCRIPTION

SECTION 4

PROCESS DESCRIPTION

The descriptions of the synthesis gas production and methanol synthesis sections of the plant are based primarily on material presented in reference 1. Modifications were made as necessary to adapt from the production of 5,500 TPD of fuel grade (95 wt%) methanol to the 19,670 TPD of crude (84 wt%) methanol required to produce 50,000 gallons per day (GPD) of gasoline.

The methanol-to-gasoline section is based on information contained in reference 2.

The process plant is divided into three sections. The first process section is the production of clean, sulfur-free synthesis gas. This gas has the proper H_2 to CO ratio to be fed to the methanol synthesis section. The process description is given in subsection 4.1, and the process is depicted in drawing R-01-FS-1. Drawings are located at the end of this section.

The second process section is the conversion of synthesis gas to crude methanol, which includes the compression of synthesis gas to the required conversion pressures. The process description is given in subsection 4.2, and the process steps are shown in drawing R-01-FS-1.

The final process section is the conversion of crude methanol to gasoline and liquified petroleum gas (LPG) product. This unit includes the methanol-to-gasoline reactors; the gas plant, and the alkylation plant. The process description is given in subsection 4.3, and the process steps are shown in drawing R-01-FS-2.

The required utilities, such as steam and electrical power, are provided as part of each process section.

4.1 SYNTHESIS GAS PRODUCTION

The synthesis gas production section of the plant consists of:

- Coal receiving, storage, and grinding.
- Coal gasification using the Texaco partial oxidation process.
- Water gas shift reaction and gas cooling.
- Acid gas removal to remove CO_2 and H_2S , along with sulfur recovery plants.
- Oxygen plants.

- Steam and power generation including flue gas cleanup.
- Cooling water and industrial water systems.
- Water reclamation.
- Ash disposal.

4.1.1 COAL RECEIVING, STORAGE, AND GRINDING

Coal is received at the plant from unit trains, each with a capacity of about 10,000 tons of 8-in. x 0 coal. The trains are unloaded into receiving track hoppers, sampled, and transferred to large storage piles by belt conveyors.

The coal is reclaimed from the large storage piles and delivered to either of two trains of primary crushers where it is reduced in size to 3/4-in. x 0. A portion of the 3/4-in. x 0 coal is withdrawn as feed to the boilers.

The remaining coal is fed to secondary crushers where it is wet ground to 14 mesh x 0. The coal is delivered by the mills as a 50 to 54 wt% solids slurry. This slurry is pumped to ball mills for final size reduction to 80% minus 200 mesh. The slurry is then pumped from the ball mills to day storage tanks as feed to the gasifiers.

4.1.2 COAL GASIFICATION

Coal slurry is pumped from the day storage tanks to six identical coal gasification trains.

Coal slurry is received in the gasification unit slurry surge tank where it is heated. The slurry is then pumped to a Texaco partial oxidation gasifier where the preheated slurry is mixed with oxygen and reacted in a non-catalytic combustion chamber. Partial oxidation occurs at 800 psig and 2,300°F and results in a mixture of hydrogen, carbon monoxide, carbon dioxide, water vapor, small quantities of methane, and residual char. The majority of the sulfur is converted to hydrogen sulfide and the remainder to carbonyl sulfide. The nitrogen in the coal is converted to free nitrogen with some traces of ammonia and hydrogen cyanide. The ash is melted by the high temperature of the reaction to yield a free-flowing molten slag.

The molten slag drops into a quench chamber where it solidifies upon contacting water. The slag is then removed from the bottom of the gasifier by cyclic discharge.

About 5% of the carbon in the feed is carried away with the synthesis gas as an unburned char containing 40 to 60% ash.

4.1.3 SYNTHESIS GAS COOLING AND DUST REMOVAL

The partially quenched gas from the slag separator enters a waste heat exchanger and is cooled to 633°F by generating 1,270-psig, 576°F steam. The 633°F exit temperature is sufficient to maintain a gas composition of 50% water vapor at the exit of the gas cleaning system. This requirement is set by the design of the shift section.

The gas then flows to a three-stage dust removal section in which it is adiabatic saturated. The moisture content of the 633°F gas increases to maintain 50% volume water vapor in the gas. The bulk of the dust is removed in the saturation process, and the residual dust is collected in a high-energy venturi scrubber followed by a final cleanup using a wash tray. The makeup water is added to the scrubbing cycle in the wash tray section. The char slurry is withdrawn from the quench scrubber with a maximum of 10% solids slurry. The liquid level in the quench section is maintained by bleeding a side stream from the venturi cycle that contains about 1% solids in the slurry. The liquid level in the venturi cycle is controlled by the makeup of preheated sour water coming from shift conversion.

4.1.4 SHIFT AND SYNTHESIS GAS COOLING

Sour synthesis gas, after ash removal and final quench, has a hydrogen to carbon monoxide molar ratio of 0.87. A molar ratio of approximately 3.7 is required for the methanol synthesis. The carbon monoxide level is adjusted by the water gas shift reaction to provide the proper hydrogen to carbon monoxide molar ratio and sufficient hydrogen for methanol synthesis. The reaction is:

A cobalt-molybdenum catalyst is used to promote this reaction in the presence of sulfur.

Synthesis gas leaves the gas cleaning section at 430°F and 770 psig and has a steam to dry gas ratio of 1:1. About one-sixth of this gas is bypassed around the shift reactor for control purposes, and the remaining gas is preheated to 50°F above the reaction initiation temperature.

Sour water, condensed as a result of cooldown and heat recovery, is separated from the gas and recycled for synthesis gas quench and scrubbing. Excess condensate is recycled on level control to the wet grinding mills.

4.1.5 ACID GAS REMOVAL

There are several competitive physical solvent processes available to treat synthesis gas. Allied Chemical Corporation's Selexol™ process was chosen for the study cited herein as reference 1.

Synthesis gas treating is accomplished in two stages. In the first stage, H_2S is selectively removed with respect to CO_2 and COS . The treated gas leaving the H_2S absorber has a total sulfur content of 210 ppm, which results in the feed stream to the sulfur plant having an H_2S concentration of 23 mol % when coal at the design sulfur content is used. With a lower sulfur content coal, the sulfur plant feed gas contains less sulfur. In the second stage, residual COS and a major portion of the inlet CO_2 are removed yielding a treated methanol synthesis gas containing 10 mol % CO_2 and less than 1 ppm total sulfur.

A. H_2S Removal Unit

Synthesis gas leaving the shift conversion unit at 715 psig and 130°F enters the H_2S absorber where it is contacted countercurrently with a lean solution presaturated with CO_2 from the CO_2 absorber. Because of the higher solubility of H_2S relative to CO_2 in the solvent, only about 20% of the inlet CO_2 is absorbed along with almost 100% removal of the H_2S . To minimize the amount of CO , CO_2 , and H_2 losses from the H_2S stripper overhead, rich solution is flashed in three successive stages, and the flashed vapors are combined, recompressed, and recycled back to the absorber after cooling.

Flashed rich solution from the third stage is pumped to the top of the H_2S stripper where stripping is accomplished using 100-psig steam. Stripper reboiler duty is minimized by heating the rich solution to 275°F against the lean solution from the stripper bottom.

B. CO_2 Removal Unit

Desulfurized gas is cooled before being contacted with chilled lean solution at 22°F in the CO_2 absorber. Lean solution from the H_2S stripper is chilled to 20°F using propane refrigerant before being combined with lean solution from the CO_2 stripper bottoms. The lean solution circulation is controlled to give a very close equilibrium at the absorber bottom that allows the CO_2 concentration in the residue gas to be held at the 10 mol % level desired for the low-pressure methanol synthesis process. Rich solution from the absorber bottoms is let down to 275 psia through a hydraulic turbine that provides part of the energy required for driving the lean solution pump. Flashed vapors are compressed and combined with the inlet gas, thus minimizing CO and H_2 losses from the CO_2 stripper overhead. Flashed rich solution is further let down to 16 psia and acts as a refrigerant before being stripped using dry nitrogen from the air separation plant. The stripping rate is set to yield a COS content below 1 ppm in the final product gas.

CO_2 -rich gas is unsuitable from an environmental standpoint for venting to the atmosphere because of its sulfur content. Therefore, it is compressed and sent to strip dusty liquor (subsection 4.1.11) before being routed to the boiler for incineration of residual sulfur compounds.

4.1.6 SULFUR PLANT

Hydrogen sulfide from the acid gas removal unit (subsection 4.1.5) is converted to sulfur in a Claus-type, three-stage sulfur recovery plant.

Acid gas from the treating unit leaves the H₂S stripper reflux accumulator and enters the inlet scrubber where any entrained sour water is separated.

The major reactions in the sulfur plant are:

- Combustion

- Conversion

- Overall

In the usual Claus plant with rich acid gas containing over 50% H₂S, about one-third of the acid gas is oxidized with a stoichiometric quantity of air, and reaction (1) in the combustion chamber of the Claus plant reaction furnace and reaction (3) also occur, yielding about 60% of the sulfur. The hot combustion products are cooled in a waste heat boiler, generating saturated, 105-psig steam, and sulfur is then condensed by generating low-pressure steam.

However, in the present case, a straight-through Claus plant cannot be used. When the gasifiers are fed with the design, high-sulfur coal, an acid feed gas of 23% H₂S content is expected. The maximum combustion temperature for such a gas, while using the contained H₂S to SO₂, is marginal at about 2,100°F. Such an operation is undesirable because it causes formation of SO₃ and consequent catalyst poisoning. If a lower-sulfur coal is fed, the H₂S content is reduced even further, leading to serious uncertainty about the operability of a bypass-type Claus plant (a straight-through Claus plant is clearly not feasible). The problems noted here are aggravated further by the requirement to process SO₂ containing gas from the flue gas cleanup unit. Consequently, a sulfur-burning Claus plant is required such as that described in reference 4.

Most of the sulfur formed in the converter is condensed by cooling the reactor effluent gas. The reaction gas is again heated up to 425°F before it enters the second converter. There, more conversion occurs due to reaction (3), and the sulfur formed is condensed by cooling the reaction gas to 340°F.

Reaction gas goes through one more stage of reheat, conversion, and sulfur condensation, which brings the total sulfur recovery to 95.0% for the three stages. All the sulfur produced is routed through steam-traced lines to the sulfur pit, which is designed for one-day storage. Sulfur is maintained

in the liquid form by steam coils in the pit. Facilities are provided for solidification and storage in the solid form of 30 days' production.

The tail gas leaving the sulfur plant is routed to the coal-fired boilers for incineration and subsequent processing in the flue gas cleanup unit.

4.1.7 OXYGEN PLANTS

The oxygen plants are conventional plants that are available from commercial air separation plant suppliers.

The oxygen production facility is based on 10 plants of about 2,100-TPD capacity each, totaling approximately 21,000 TPD. Each plant will include the cold box, distillation columns, heat exchange equipment, air and oxygen compressors, and auxiliary equipment. The product will be gaseous oxygen of 99.5% purity, compressed to 935 psig by centrifugal compressors.

4.1.8 STEAM AND POWER GENERATION

A coal-fired steam plant is provided to produce the amount of required high-pressure, saturated steam (1,200 psig) in excess of that produced by waste heat boilers in the gasifier and shift conversion areas. In addition, all saturated, high-pressure steam is sent to the steam plant for superheating to produce high-pressure steam header conditions of 1,200 psig and 950°F. Facilities are also provided for treated makeup water storage, deaeration of plant condensate, and electrical power generation.

Steam at pressures of 500, 100, and 20 psig is also produced in the gasifier, shift, and synthesis areas to supply heating requirements for various users and power for smaller drivers.

In addition to firing coal, the steam plant boilers also fire spent char from the gasifier area and purge gas from the methanol synthesis section. Air for the boilers is preheated to 220°F by 20-psig steam. Then, CO₂-rich gas at 220°F from the stripping section in the gasification area is added to the airstream.

A flue gas cleanup system is provided to collect the fly ash and unburned carbon from the flue gas of the coal-fired boilers and to recover SO₂ from the gas by a regenerative process. This flue gas contains approximately 770 TPD of particulates and 180 TPD of sulfur as SO₂ and SO₃. Over 99% of the solids and 97% of the sulfur is removed from the flue gas. A high-energy wet scrubbing system, using cooling tower blowdown as makeup, collects essentially all the fly ash and unburned carbon in the boiler flue gas.

The Wellman-Lord wet scrubbing (single alkali) regenerative process was used for this study.¹ The dust-free gas enters the three-stage SO₂ absorber at about 136°F. Then, SO₂ is absorbed into a 28 wt% sodium sulfite solution where the following reaction takes place:

The desulfurized flue gas that contains less than 110 ppm SO₂ is compressed by an induced draft fan, reheated to 220°F by using 20-psig steam, and vented to the atmosphere. (The flue gas is reheated to reduce fogging around the plant and to protect the stack against corrosion.) The top of the stack is 140 ft above the ground.

SO₂ is recovered from the bisulfite-rich solution leaving the absorber by heating the solution in a double-effect evaporator/crystallizer, which reverses the absorption reaction. The SO₂-rich gas stream is routed to the sulfur plant for subsequent processing to elemental sulfur.

About 2% of the SO₂ entering the absorber is converted to sodium sulfate by the following reaction:

The Na₂SO₄ is recovered by selective crystallization (15 TPD) as a dry and relatively clean salt.¹

Dusty liquor purged from the wet gas scrubbing system containing some dissolved SO₂ and SO₃ (pH of 2) flows to the ash sluice system of the boilers to remove bottom ash and is then neutralized with lime and flows to ash disposal.¹

4.1.9 COOLING WATER SYSTEM

Induced draft, evaporative-type cooling towers are used to provide 583,400 gallons per minute (gpm) of 80°F cooling water. The cooling towers have a total capacity of 8,751 MM Btu/hr at the design wet bulb condition.

Cooling water from the cooling tower is pumped to a cooling water header at 80°F. The water is used to cool various process streams and is, in turn, heated to 84°F. The water is then used in the surface condensers to condense turbine exhaust steam and is heated to 110°F. This water is returned to the cooling tower for reuse in the system. Sulfuric acid is added to the cooling tower pump to control scaling. Cooling tower blowdown is pumped to the flue gas cleanup area. A cooling water chlorinator has been provided to control biological growth, and an inhibitor addition system is used to control corrosion.

4.1.10 INDUSTRIAL WATER SYSTEMS

Water in the amount of 21,140 gpm (31,000 acre-ft/yr) will be pumped from a river source. A summary of the overall water balance is shown in Figure 4-1.

Over 90% of the raw water is consumed in utilities and cooling tower losses. The balance of the water is consumed in the process, lost in evaporation, or used in the ash slurry.

Steam system makeup water is pumped to an ion exchange unit where the water is purified for use as boiler feed water. This treated water is then pumped to the treated water storage tank.

The firewater system consists of firewater pumps (two diesel-engine driven and one motor driven), firewater pressure pumps, and the water distribution system. The raw water tank is used as the source for firewater.

4.1.11 WATER RECOVERY AND ASH DISPOSAL

The CO₂-rich gas from the acid gas treating plant (subsection 4.1.6) is used to strip dissolved gases containing H₂S from the dusty liquors rejected from both the gasifier slag quench section and the synthesis gas scrubber bottom effluent. After stripping, the gas is reheated before joining the combustion air to the coal-fired boilers.

Slag slurry from the gasifier (2,888-gpm, 14.5 wt% solids) is pumped through a receiving tank and splitter to horizontal, double-deck vibrating screens dressed to make 3/4-in. and 28-mesh separations. From these screens the +3/4-in. ash goes to the ash conveyor, the 3/4-in. x 28-mesh ash goes to centrifuges, and the 28-mesh x 0 ash with most of the water goes to thickeners.

The centrifuges receiving 3/4-in. x 28 mesh from the screens discharge solids to the ash conveyor and effluent to the thickeners.

Bottom ash slurry from the coal-fired boilers (1,540 gpm, 3.0 wt%, and 200 mesh) has a pH of 2.0 to 2.5. The mix is neutralized with milk of lime (Ca[OH]₂) and is then routed to the thickeners.

Underflow from the thickeners is filtered, and the solids are discharged to the ash conveyor. Overflow from the thickeners, containing up to 0.5 wt% solids, and filtrate, is recycled to the coal slurry preparation area. A 3-1/2-acre reservoir is provided for water surge and upset conditions.

Quantities of 4,010 TPD of ash and 2,560 TPD of contained water are returned to loading and ultimate refilling at the mines.

Dusty liquor withdrawn from the synthesis gas section is cooled by heat exchange, depressured, and stripped with CO₂-rich gas. The stripped liquor goes to thickeners where evaporation reduces its temperature to 120°F. Underflow from the thickeners is filtered, and the wet char is discharged and conveyed to the steam plant as fuel. Thickener overflow and filtrate are combined, reheated to 410°F, and reinjected to the wash tray of the scrubbing section. A fine-solids purge stream is withdrawn to the gasifier slag disposal system, thereby removing excess sour water.

The condensate from the M-Gasoline plant (2,070 gpm) has a pH of 3.5. The condensate is stripped with CO₂-rich gas and is then neutralized with milk of lime. This recovered water provides makeup requirements to the gasifier section and wet grinding circuits. Excess condensate is sent as makeup to the cooling water system.

FIGURE 4-1
OVERALL WATER BALANCE
COAL TO M-GASOLINE PROCESS

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All blowdown from the steam system is sent as makeup to the cooling water system. Blowdown from the cooling water system is used as makeup to the flue gas wet scrubbing system and to sluice ash from the coal-fired boilers. Blowdown from the raw water treatment and boiler feedwater treatment facilities, as well as water recovered from the oily water sewer, is added to the gasifier slag disposal system.

4.2 CRUDE METHANOL SYNTHESIS

Imperial Chemical Industry's low-pressure methanol synthesis was used for this study.¹ Six methanol plant trains having a total capacity of about 19,670 TPD crude methanol (84 wt% methanol) are provided.

Synthesis gas at 680 psig and 85°F from the treating unit is passed through desulfurizer drums. Each drum contains two packed beds of activated carbon granules that act as a sulfur guard for the methanol synthesis catalyst. The drums are designed to operate on onstream for 3 days. At the end of this period, one drum is taken offstream while a fresh bed is put online. The spent carbon is regenerated with steam and is then returned to service.

From the desulfurizer drums, the gas enters the synthesis makeup compressor and is compressed to 1,542 psig. The discharge gas from the compressor is combined with synthesis recycle gas and is cooled in an air cooler to 130°F.

The cool gas then enters the synthesis loop, which consists of a synthesis converter, heat exchange train, and recycle compressor. In the loop the gas is divided, with a portion of the gas sent as quench for the converter and the remainder sent as feed to the converter. To achieve the temperature necessary for reaction to methanol, the feed gas is first passed through the converter feed preheater where the gas is heated by hot reactor effluent.

The preheated feed gas enters the top of the converter vessel and flows downward through several catalyst beds to the converter outlet. At the exit of each catalyst bed, cold quench gas is injected to control the inlet temperature of the next catalyst bed.

Reaction to methanol proceeds by the following:

The outlet gas contains about 5.5 mol % methanol together with large amounts of unreacted CO, CO₂, and H₂; inerts such as N₂, Ar, and CH₄; and some quantities of byproducts such as water, dimethyl ether, and higher alcohols.

To separate the unreacted gases from the product methanol, the converter outlet gas is cooled in a heat exchanger train. The resulting two-phase mixture from the cooldown train is separated, and the unreacted gases are sent to the recycle compressor. The condensed liquid containing about 73 mol % methanol and 25 mol % water corresponding to about 84 wt% methanol is sent to the methanol-to-gasoline conversion plant (M-Gasoline plant).

The purge gas from the synthesis loops is sent to the steam and power generation (subsection 4.1.8) in the synthesis gas generation unit.

The utilities required for this section are consolidated with those of the synthesis gas production and are described in subsections 4.1.8 through 4.1.10.

4.3 METHANOL-TO-GASOLINE CONVERSION

This section of the plant provides the conversion of methanol to gasoline and other hydrocarbons, a gas plant to separate the various hydrocarbon streams, an alkylation plant to produce additional gasoline from unsaturated hydrocarbons, product storage, and blending, and the necessary utilities to support the plant operation.

4.3.1 METHANOL-TO-GASOLINE

The conversion of methanol to gasoline is based on the M-Gasoline process utilizing fixed-bed, catalytic reactors.² The hydrocarbon product from this process is primarily gasoline; it consists of highly branched paraffins, highly branched olefins, naphthenes, and aromatics.

Five M-Gasoline plant trains are provided; each train processes about 27,000 BPD of crude methanol.

The overall flow scheme for the M-Gasoline plant is shown in block flow diagram R-01-FS-2, and the conversion of methanol to hydrocarbons for one train is shown in process flow diagram R-01-FS-3.

Crude methanol flows from the methanol plants into Surge Drum 1203. Flashed gases are returned to the methanol plant. Crude methanol is fed under flow control to Reactor Effluent/Methanol Exchanger 1301 where it is vaporized and superheated. Methanol vapor enters DME Reactors 2501-01 or -02 at 600°F and 300 psig, where it is catalytically converted to an equilibrium mixture of dimethyl ether, methanol, and water by the following reaction:

This reaction is exothermic and, as a result, the effluent leaves the DME reactor at about 770°F. The mixture then flows to five parallel M-Gasoline Reactors, 2502-01, -02, -03, -04, and -05, where catalytic conversion to M-Gasoline occurs. The second DME Reactor, 2501-02, and the fifth M-Gasoline Reactor, 2502-05, are shown being regenerated.

To adjust the M-Gasoline reactor inlet temperature to 650°F and to limit the temperature rise to about 200°F, a recycle gas stream is introduced along with DME Reactor 2501 effluent into the M-Gasoline Reactors.

The M-Gasoline reactors use a shape-selective zeolite catalyst to complete the conversion of methanol to hydrocarbons and water. A conversion of +99% is achieved in a single pass. Essentially no hydrocarbons are produced above C₁₁.

This process can be represented by the following overall reaction:

The M-Gasoline reaction products are cooled by exchange with recycle gas in Reactor Effluent/Recycle Gas Exchanger 1302 and with crude methanol in Exchanger 1301. The reactor products are cooled further by preheating boiler feedwater (on only two of the M-Gasoline plant trains) and by air and water cooling. The products are separated into a hydrocarbon vapor phase, and liquid hydrocarbon and water phase in Product Separator 1201.

The vapors from Product Separator 1201 are split into a recycle stream that flows to Gas Recycle Compressor 1801, and an offgas stream that is sent to the gas plant (subsection 4.3.3). Hydrocarbon liquids are also sent to the gas plant.

The water, which has a pH of 3.5, is recycled to the gasifier and other facilities after neutralization.

Durene, which freezes at 176°F, is the most abundant heavy aromatic produced during the methanol conversion. To ensure that sufficient liquid exists to dissolve any durene that may condense, hydrocarbon liquid from Product Separator 1201 is injected upstream of BFW Heater 1304 and Effluent Air Cooler 1303.

The DME reactors and five M-Gasoline reactors are provided in each train to allow for continuous operation at 27,000 barrels per stream day (BPSD) throughput during regeneration of the catalyst. These reactors are valved to allow a rapid change of service from process to regeneration. The regeneration of the M-Gasoline reactors will take place about once every 5 to 6 days and the regeneration of the DME reactors about every 2 weeks or longer. Nitrogen is added to air to form a dilute oxygen mixture that is used to burn off the coke that forms on the catalyst during normal operation. Compressor 1802 is used to boost air to Regeneration Recycle Compressor 1803. The dilute oxygen stream is heated to 650°F at 300 psig in Regeneration Furnace 1401.

During the major burning period, the reactor outlet oxygen concentration is held at less than 1 vol% by controlling the rate of makeup air or by adjusting the inlet temperature. The inlet temperature is gradually raised to 900°F, and after burning has ceased, the oxygen concentration is allowed to rise to 20 vol%. The reactor is then cooled, and the unit is purged of residual oxygen.

The DME reactors and M-Gasoline reactors may be independently regenerated. Total regeneration time for a complete cycle is 1 day.

Regeneration offgases are vented to the power plant boilers.

4.3.2 GAS PLANT

A single-train oil absorption gas plant has been provided to fractionate the total hydrocarbon product from the five M-Gasoline plant trains into a fuel gas stream, a C₃ to C₅ olefinic stream, and a C₅ gasoline blending stock.

The process flow scheme for the gas plant is shown in process flow diagram No. R-01-FS-4. The flow scheme is typical of the lean oil absorption plants used in many refineries.

The vapor and liquid hydrocarbon streams from the M-Gasoline plants are combined with the overhead vapor from Stripper 1102 and the absorption oil from the bottom of Absorber 1101 and flow through Low Pressure Separator Feed Cooler 1301 to Low Pressure Separator 1201.

Light hydrocarbons in the gas stream are absorbed into the higher molecular weight liquid hydrocarbons. The heat of absorption, approximately equivalent to the heat of condensation, is removed at a temperature higher than that which would have been required by straight condensation.

Vapor separated in Product Separator 1201 enters Absorber 1201 where it is contacted with lean oil at ambient temperatures. Vapor from the top of the absorber goes to plant fuel.

Hydrocarbon liquid separated in Product Separator 1201 is preheated in Depentanizer Bottoms/Stripper Feed Exchanger 1302 and then flows to Reboiled Stripper 1102 where it is deethanized.

Stripper bottoms flow to Depentanizer 1103 where a C₃ to C₅ olefinic stream is fractionated overhead as feed to the alkylation plant (subsection 4.3.4). Depentanizer bottoms, after heat exchange and cooling, are recycled to the absorber. A portion of the depentanizer bottoms is withdrawn on level control to gasoline blending stock storage.

4.3.3 ALKYLATION

A single-train alkylation plant is provided to alkylate the butenes and pentenes recovered by the gas plant (subsection 4.3.3) into additional gasoline.

Sulfuric acid and hydrogen fluoride (HF) are the preferred catalysts for the alkylation of C₃ and C₄ olefins by isobutane (and isopentane). This study is based on an HF alkylation process as offered commercially.

Some 8,750 BPD of alkylate are produced.

The olefinic feed stream from the gas plant is depropanized prior to the alkylation because of the small quantity of propylene available for alkylation. The dry depropanizer bottoms, containing an excess of isobutane, are mixed with additional isobutane as recycle and are then brought into intimate contact with the HF catalyst in a water-cooled reactor. The olefins react with isobutane to yield the corresponding alkylates:

alkylate

The reactor effluent is settled, and the acid phase, HF, is recycled to the reactor; occasionally, a small portion is regenerated. The hydrocarbon phase is fed to a main fractionator where the alkylate is freed of recycle isobutane and lighter products. Stabilization of the alkylate product by removal of normal butane is accomplished by a main fractionator vapor side stream. Small amounts of propane that enter the plant are removed from the main fractionator overhead. This purge stream is freed of acid in an HF stripper and is then recycled to the depropanizer.

Alkylate from the bottom of the fractionator passes through a caustic treater for removal of trace acid fluorides before flowing to storage.

4.3.4 STORAGE - BLENDING - LOADING

Intermediate storage for methanol is provided by four 355,000-barrel, nitrogen-blanketed tanks. Component storage tanks for gasoline blending stocks are provided as follows:

- (1) Butanes for blending are stored in three 20,000-barrel spheres.
- (2) Alkylate is stored in two 70,000-barrel floating roof tanks.
- (3) Raw gasoline is stored in two 145,000-barrel floating roof tanks.

Transfer pumps are provided for blending the gasoline stocks to finished gasoline at the rate of 6,000 barrels per hour (BPH). Finished gasoline storage is provided by eight 200,000-barrel floating roof tanks.

LPG product is stored in three 20,000-barrel spheres. All LPG will be dispatched by road tanker. About 30% of the finished gasoline will be dispatched by road tanker, with the remaining 70% dispatched by rail tank car and unit train.

Steel silos are provided for receiving ash delivered by belt conveyor from the gasifier slag disposal system. The ash is returned by unit train to the coal mine or other suitable site.

4.3.5 STEAM GENERATION AND UTILITIES

An auxiliary steam generation plant has also been provided to supply 500-psig steam to the M-Gasoline plant area, and for superheating excess saturated, 100-psig steam for use in turbine drivers. The boilers are normally

fired by fuel gas from the gas plant. The boilers can also be fired by fuel oil, LPG, or methanol to facilitate startup of the complex.

All blowdown from the steam system is sent as makeup to the cooling water system.

The cooling water requirements are discussed in subsection 4.1.9.

4.4 PRODUCT CHARACTERISTICS

4.4.1 LIQUIFIED PETROLEUM GAS

The LPG product is a mixture of propane and propylene. Butanes are not normally exported from the plant.

Properties of LPG are:

<u>Component</u>	<u>Property</u>	Value
Propane, vol%	87	
Propylene, vol%	13	
Total	100	
Vapor pressure at 100°F, psig	205	
Butane and heavier, % max	2.5	
Sulfur, grains/100 ft ³	Nil	
Corrosion, copper strip, max	No. 1	

4.4.2 FINISHED GASOLINE PRODUCT

The finished gasoline product has a boiling range typical of premium gasolines (C₄ to C₁₀).

Gasoline octanes exceed requirements for unleaded regular (ASTM Grade 2) or leaded premium.

Product composition is similar to that of petroleum-derived gasolines. The gasoline consists of highly branched paraffins, highly branched olefins, naphthenes, and aromatics. The finished gasoline differs from conventional gasolines in that durene (C₁₀ alkylbenzene) can be present in relatively large amounts (3 to 6 wt%). Durene levels of 5 wt% have been found to cause unsatisfactory carburation in some cars under certain test conditions because of crystallizing out of durene, which freezes at 175°F. However, effects are minimal at the 4% level. Gasoline with a high durene content can be improved by blending with conventional gasoline.

Additives are required to inhibit gum formation (5-lb metal deactivator plus 15-lb antioxidant per 1,000 barrels of gasoline). High gum-forming gasolines can be improved by blending with conventional petroleum products that form lower potential gums.

The projected properties of the finished gasoline are:

<u>Component</u>	<u>Property</u>
C ₅ 's and lighter, vol%	14.9
C ₆ + synthesized gasoline, vol%	67.6
Alkylate, vol%	<u>17.5</u>
Total	100.0
Research octane - clear	92
Reid vapor pressure, psig	9.5
Oxygenates, wt%	nil
Nitrogen, wt%	nil
Sulfur, wt%	nil
Corrosion, copper strip (ASTM D 130)	1A

- NOTES:
1. THIS FLOWSHEET REPRESENTS ONE OF FIVE TRAINS REQUIRED TO PRODUCE 50,000 BARRELS PER DAY OF FINISHED GASOLINE.
 2. BFW HEATERS TO BE SUPPLIED ON 2 OF 5 TRAINS.
 3. STREAM DATA HAS BEEN PRORATED FROM 100 BPSD PILOT PLANT FLOW SHEETS. THIS STREAM DATA HAS BEEN USED FOR EQUIPMENT SIZING ONLY.

STREAM NO.	1	2	3	4	5	6	7	STREAM NO.	8	9	21	22	23	24	25	26	27
STREAM NAME	CRUDE METHANOL	DME REACTOR EFFLUENT	RECYCLE GAS	M-GASOLINE REACTOR EFFLUENT	SEPARATOR VAPOR	SEPARATOR LIQUID	SEPARATOR WATER	STREAM NAME	SEPARATOR OFF-GAS	HYDROCARBON QUENCH	REGEN. AIR	REGEN. REACTOR GAS	REGEN. REACTOR INLET	REGEN. REACTOR OUTLET	REGEN. SEPARATOR VAPOR	REGEN. SEPARATOR WATER	REGEN. OFF-GAS
FLOWS, LB MPM	-	-	10,838	10,868	10,863	5	-	FLOWS, LB MPM	-	-	25	21	-	-	-	-	-
H ₂	-	-	-	-	-	-	-	H ₂	-	-	138	3,245	3,245	3,245	3,245	-	138
N ₂	-	-	-	-	-	-	-	N ₂	-	-	37	37	37	37	-	-	-
O ₂	-	-	-	-	-	-	-	O ₂	-	-	206	-	-	-	-	-	-
CO	-	-	11,375	11,449	11,400	49	-	CO	-	-	25	-	-	-	-	-	-
CO ₂	-	-	700	703	703	3	-	CO ₂	-	-	1	13	-	-	-	-	-
C ₂ H ₆	-	-	371	375	372	3	-	C ₂ H ₆	-	-	5	21	-	-	-	-	-
C ₃ H ₈	-	-	262	267	262	5	-	C ₃ H ₈	-	-	5	21	-	-	-	-	-
C ₄ H ₁₀	-	-	3,033	3,090	3,038	52	-	C ₄ H ₁₀	-	-	-	736	736	769	769	-	33
C ₅ H ₁₂	-	-	633	653	633	30	-	C ₅ H ₁₂	-	-	-	-	-	-	-	-	-
C ₆ H ₁₄	-	-	4,436	4,673	4,444	229	-	C ₆ H ₁₄	-	-	-	-	-	-	-	-	-
IC ₄	-	-	1,489	1,674	1,492	182	-	IC ₄	-	-	-	-	-	-	-	-	-
IC ₅	-	-	485	559	488	71	-	IC ₅	-	-	-	-	-	-	-	-	-
IC ₆	-	-	534	625	537	87	-	IC ₆	-	-	3	365	-	-	-	-	-
C ₈ -C ₁₀	-	-	802	1,352	802	550	-	C ₈ -C ₁₀	-	-	-	2,310	-	-	-	-	-
BENZENE	-	-	5	15	5	16	-	BENZENE	-	-	-	134	-	-	-	-	-
TOLUENE	-	-	16	71	16	55	-	TOLUENE	-	-	-	231	-	-	-	-	-
C ₈ -C ₁₁ AROM.	-	-	21	315	21	294	-	C ₈ -C ₁₁ AROM.	-	-	-	1,235	-	-	-	-	-
METHANOL	8,599	1,988	-	-	-	TRACE	-	METHANOL	-	-	-	-	-	-	-	-	-
DIMETHYL ETHER	2,911	2,306	-	-	-	-	-	DIMETHYL ETHER	-	-	-	-	-	-	-	-	-
H ₂ O	-	6,216	140	11,630	140	3	11,496	H ₂ O	-	-	5	32	32	35	26	7	1
TOTAL	11,510	11,510	35,140	48,335	35,216	1,623	11,496	TOTAL	76	6,000	180	4,050	4,050	4,049	4,042	7	172
BPSD	27,295	-	-	-	-	12,226	14,214	BPSD	0.7	51.0	1.64	36.89	36.89	36.88	36.82	9	1.57
MMSCFD	-	104.8	320.0	440.2	320.7	-	-	MMSCFD	0.7	-	1.64	36.89	36.89	36.88	36.82	-	-
MM	28.5	-	22	22.8	22	73.7	18	MM	22	73.7	29	31	31	31	31	18	31

REV.	DATE	BY	CHKD.	DATE	BY	CHKD.	DATE	BY	CHKD.	DATE
0										
PRELIMINARY										
PROCESS FLOW DIAGRAM CRUDE METHANOL TO M-GASOLINE (1 OF 5 TRAINS)										
THE RALPH M. PARSONS COMPANY PARADISE, CALIFORNIA			JOB NO. 5658-01		DWG. NO. R-01-FS-3		REV. 0			

4-20

4-20

NOTE: MATERIAL BALANCE INDICATED IS APPROXIMATION USED FOR EQUIPMENT
SIZING AND COST ESTIMATING.

STREAM NO.	1	2	3	4	5	6	7	8	9
STREAM NAME	HC FROM M-GASOLINE UNIT	SHG TO PLANT FUEL GAS	DEPEND. OVHD.	C ₆ +	WATER	FLASH VAPOR	STRIPPER FEED	STRIPPER OVERHEAD	DEPEND. FEED
LB/HR	607,000	19,400	199,000	388,000	400	34,000	715,000	67,000	648,000
MM	-	23.43	-	-	-	29	-	50	-
°API	81	-	115	48	10	-	125	-	-
BPD	62,570	7.62	24,000	33,785	27	-	89,000	-	-
WISCFD	-	-	-	-	-	10.7	-	12.2	-
TEMP. (°F)	100	135	100	100	100	140	150	270	-
PRESS. (PSIA)	230	220	130	125	225	225	290	230	235

REV. NO.	CURT. APPR.	DATE	REVISION	BY	PROC. ENGR.	CHECK. ENGR.	CHIEF. ENGR.	CHIEF. OPER. ENGR.	PROC. DIR.
0		7/4/77	PRELIMINARY						
PROCESS FLOW DIAGRAM GAS PLANT									
THE RALPH M. PARSONS COMPANY PASADENA, CALIFORNIA				JOB NO. 5658-1	OWN. NO. R-01-FS-4	REV. 3			

SECTION 5
MATERIAL BALANCE

SECTION 5

MATERIAL BALANCE

The overall material balance is shown in Figure 5-1. Of the total feed input of about 180,000 TPD, approximately 92% is discharged as gas, 3.6% as ash slurry, 0.6% as sulfur, and 3.8% as LPG and gasoline.

The balances for the individual units are shown on their respective flowsheets.

Figure 5-1 - Overall Material Balance —
Coal to M-Gasoline Process

SECTION 6
ENERGY BALANCE

SECTION 6

ENERGY BALANCE

The overall energy balance is illustrated in Figure 6-1. Of the 26,383 MM Btu/hr higher heating value (HHV) energy input from coal, 14,268 MM Btu/hr (approximately 54%) is consumed within the complex, and 12,115 MM Btu/hr (or 46%) is available as product exports. Substitute natural gas (SNG) produced from the gas plant is used for plant fuel gas. Of this SNG, 135 MM Btu/hr is fed to the fired reboiler in the alkylation plant, 42 MM Btu/hr is used for catalyst regeneration in the M-Gasoline plant, and 230 MM Btu/hr is used to generate 500-psig, saturated steam.

Process coal charged to synthesis gas production is 91% of the total coal feed; the balance (or 9%) is burned in coal-fired boilers for steam and power generation. Char recovered from the gasification step (822 MM Btu/hr) and purge gas (873 MM Btu/hr) are also charged to the utility boilers. The total fuel burned in the utility boilers is 4,295 MM Btu/hr, or 14% of the coal energy input.

The thermal efficiency of the gasification section, based on process coal feed, is estimated to be about 70% based on HHVs of coal and synthesis gas.

The thermal efficiency for the synthesis gas to methanol step is projected to be about 85%.

Methanol-to-gasoline and LPG thermal efficiency is estimated to be about 87%. Including surplus SNG, the thermal efficiency is 89%.

The amount of energy as sulfur product represents 1.5% of the input coal energy.

	HIGHER HEATING VALUE		LOWER HEATING VALUE	
	BILLION BTU/DAY	% OF TOTAL COAL FEED	BILLION BTU/DAY	% OF TOTAL COAL FEED
COAL FEED				
TO METHANOL CONVERSION	576.3	91.0	556.2	91.0
TO UTILITIES & STEAM GENERATION	56.9	9.0	54.9	9.0
TOTAL	633.2	100.0	611.1	100.0
PRODUCTS				
LPG	22.9	3.6	20.7	3.4
GASOLINE	258.6	40.8	241.0	39.4
SULFUR	9.3	1.5	9.3	1.5
TOTAL	290.8	45.9	271.0	44.3

NOTE: ALL HEAT QUANTITIES ON DIAGRAM ARE HHV.

FIGURE 6-1
OVERALL ENERGY BALANCE
COAL TO M-GASOLINE PROCESS

SECTION 7
UTILITIES

SECTION 7

UTILITIES

The overall utility summary, shown in Table 7-1, illustrates the utility productions and consumptions by type and unit. The summary is for the entire coal to M-Gasoline complex.

All utilities required for operation are generated within the complex. The complex represents the combining of two major plants, the Fuel Grade Methanol from Coal Plant¹ and the Methanol to M-Gasoline Plant.² The utility table shows the combined utilities for these two plants. The basic utility scheme for the Fuel Grade Methanol from Coal plant was modified slightly after scale-up as described below:

- The Methanol Fractionation section has been deleted resulting in availability for use of 100-psig, saturated steam and electricity elsewhere in the complex.
- Approximately 40% of the cooling water pumps in the scaled-up process (196,000 out of 507,600 gpm) have been changed from motor-drive to 100-psig/400°F steam drive. Also, turbine condensers have been added to condense the exhaust steam from these cooling water pump turbines.
- Electrical power previously used in the methanol fractionation section and for the cooling water pumps is now used for power requirements of the M-Gasoline from Methanol Plant. This amounts to 10,445 kW out of a total 103,852 kW generated.

Additional utility requirements for operation of the M-Gasoline from Methanol Plant are supplied by:

- Increasing the cooling water capacity of 507,600 gpm (reference 2) by 75,800 to 583,400 gpm total.
- Adding a 500-psig, saturated steam boiler to produce 185,000 lb/hr of process steam.
- Superheating 715,500 lb/hr of 100-psig, saturated steam from the deleted methanol fractionation section to 400°F for use as motive steam for:
 - The added cooling tower (75,800-gpm) pump motive steam.
 - The replaced motor-driven pumps in the cooling tower system (196,000 gpm) to steam-driven pumps.
 - The recycle compressors in the M-Gasoline unit (507,500 lb/hr).

Table 7-1 - Overall Utility Summary

UNIT DESCRIPTION	FUEL MMBTU/HR	1175 PSIG 925°F STEAM LB/HR	1275 PSIG SAT'D STEAM LB/HR	550 PSIG 750°F EXTRACTION STEAM LB/HR	500 PSIG SAT'D STEAM LB/HR	100 PSIG 400°F STEAM LB/HR	100 PSIG SAT'D STEAM LB/HR	20 PSIG SAT'D STEAM LB/HR	POWER KW-H	COOLING WATER GPM
COAL PREPARATION & ASH DISPOSAL	-	-	-	-	-	-	-	-	35,934	-
COAL GASIFICATION, SYN GAS COOLING & CHAR RECOVERY	-	-	(3,269,400)	-	-	-	(80,000)	147,600	9,773	2,600
AIR SEPARATION & OXYGEN COMPRESSION	-	3,096,100	-	-	-	-	-	-	610	253,500
SHIFT	-	-	(401,000)	-	-	-	(965,700)	(462,000)	1,386	-
GAS TREATING & REFRIGERATION	-	256,700	-	401,000	-	-	240,600	150,800	6,025	81,800
METHANOL SYNTHESIS	-	1,655,500	-	(702,600)	-	-	-	(352,900)	3,218	92,100
M-GASOLINE PLANT	42	-	-	-	-	507,500	-	-	7,085	55,300
GAS PLANT	-	-	-	-	100,000	-	46,000	-	320	1,200
DEPROPANIZER & ALKYLATION	135	-	-	-	85,000	-	49,500	-	1,200	6,100
SULFUR PLANT	-	-	-	-	-	-	(48,000)	(86,600)	1,078	-
STACK GAS CLEANUP	-	-	-	138,000	-	-	-	426,700	6,725	11,200
STEAM GENERATION	4,295	(5,916,300)	3,670,400	163,600	(185,000)	(715,500)	757,600	176,500	2,098	9,600
POWER GENERATION	-	408,000	-	-	-	-	-	-	-	56,800
COOLING WATER SYSTEM & OFFSITES	-	-	-	-	-	208,000	-	-	20,560	13,200
DISTRIBUTION LOSSES, LIGHTING, HVAC & INSTRUMENTS	-	-	-	-	-	-	-	-	7,840	-
TOTAL	4,472	0	0	0	0	0	0	0	103,852	583,400

NOTE: () DENOTES EXPORT

Total fuel fired to the steam generation section is 4,295 MM Btu/hr and includes 4,065 MM Btu/hr fired to the 1,175-psig, superheated steam generation facilities and 230 MM Btu/hr fired to the additional 500-psig steam boiler and 100-psig steam superheater. The fuel for the 1,175-psig, superheated steam consists of 2,371 MM Btu/hr of utility coal, 821 MM Btu/hr of char, and 873 MM Btu/hr of purge gas from Methanol Synthesis. The 500-psig steam boiler and 100-psig superheater is fired by SNG from the gas plant.

SECTION 8
ECONOMIC ANALYSIS

SECTION 8
ECONOMIC ANALYSIS

The economic analysis presented in this section is based on mid-1977 dollars. This analysis is based on a plant operation of 330 days/yr with an operating life of 20 yr. A 5-yr construction schedule was used in the analysis.

In addition to analysis of the total complex, an incremental analysis was performed by computing value added for each plant section - coal to syngas, syngas to methanol, and methanol to gasoline. In these analyses, LPG is credited at the same value per million Btu as gasoline.

The financial factors used for the base case analysis are:

- (1) Interest during construction is based on a 9% interest rate. The commitment fee is 0.75% of the uncommitted loan balance.
- (2) Working capital is based on the following:
 - 30-day coal inventory.
 - 30-day inventory of finished product.
 - 4% of major equipment for spare parts inventory.
 - 30-day accounts receivable.
 - 30-day budget for current expenses.
 - 30-day credit for accounts payable.
- (3) Consumption of coal increased from section to section - 25,915 for syngas, 26,640 through methanol, to a total of 27,920 TPD at \$20/ton for gasoline and LPG product.
- (4) Operating supplies are computed at 0.05% of fixed capital investment.
- (5) The operators per shift are:

<u>Unit</u>	<u>Operators per Shift</u>
Synthesis gas production	182
Methanol	35
Gasoline	<u>20</u>
Total	237

Supervision was added at 15% of the operators at an average hourly wage rate of \$7.50.

- (6) The payroll burden is 35% of total labor, and plant overhead is 60% of the operating and maintenance labor including the payroll burden.
- (7) General and administrative overhead is computed at 1.5% of manufacturing costs.
- (8) Maintenance costs are calculated at 4% of fixed capital investment; 40% is applied to labor and 60% is applied to materials.
- (9) Total property tax and insurance is 2.75% of fixed capital investment.
- (10) The double declining balance method of calculating depreciation was used. The useful life for depreciation purposes is 13 yr.
- (11) The salvage value of the plant is zero investment.
- (12) Where a 65/35 debt structure is used, 65% of the depreciable investment was borrowed. The interest on the loan is 9% per annum, repayable in 20 yr in equal payments of principal. Interest of 9% is charged on working capital.
- (13) The DCF rate of return is 12% after 52% combined Federal and state income taxes, and 9% investment tax credit.

8.1 REQUIRED PRODUCT PRICES

Calculations of required product selling prices (RPSP) are based on a 65/35 debt-to-equity structure at a 9% interest rate and 100% equity financing; RPSPs were developed for 0, 12, and 16% DCF rate of returns, respectively. The RPSP for the base case, 65/35 debt-to-equity ratio at a 9% interest rate and a 12% DCF rate of return, results in a gasoline price of 82¢ per gallon.

Using 100% equity financing, the required selling price for gasoline ranges from 55¢ per gallon at a 10% DCF rate of return to \$1.20 per gallon at a 16% DCF rate of return.

Parsons analysis of factors such as fixed capital investment and operating costs indicates that the fixed capital investment for the coal-to-syngas plant and coal prices have the most influence on required selling prices.

At the base case (65/35 debt-to-equity ratio and 12% DCF rate of return) annual revenues required for the entire complex amount to \$644 million.

8.2 FIXED CAPITAL INVESTMENT

The overall FCI is estimated at about \$1.6 billion based on mid-1977 costs. A summary of the fixed capital investment is given in Table 8-1. For purposes of the economic evaluation, the FCI for utilities, steam and

Table 8-1 - Summary Fixed Capital Investment

FACILITY	INVESTMENT (\$000,000)	PERCENT OF OF TOTAL
COAL PREPARATION	70	4.4
GASIFICATION	200	12.6
OXYGEN PLANTS	387	24.4
RAW GAS SHIFT	73	4.6
ACID GAS REMOVAL, SULFUR PLANT	172	10.8
METHANOL SYNTHESIS	145	9.1
METHANOL TO GASOLINE CONVERSION	87	5.5
GAS PLANT	4	0.3
HF ALKYLATION PLANT	12	0.7
STEAM AND POWER GENERATION	235	14.8
OFFSITES, STORAGE, AND OTHER UTILITIES	204	12.8
TOTAL	1,589	100.0
ROUNDED	1.6 BILLION	

power generation, storage, and offsites are allocated according to their usage in the three basic sections of the plant.

The bases for the estimates are described in the following subsections.

8.2.1 SYNTHESIS GAS AND METHANOL UNITS

The investment estimates contained in reference 1 were used as the main basis of the estimate.

Individual estimates of the plant sections were prepared using a plant capacity scale-up factor and correcting for escalation from mid-1975 to mid-1977. No corrections are included for plant site location. No contingency allowance has been included.

8.2.2 METHANOL TO GASOLINE

Basic equipment sizes and duties as reported in reference 2 for a 10,000-BPD plant were used as the design basis. The various equipment items - pumps, compressors, heat exchangers, and vessels - were detailed to the extent that costs could be obtained for each of the five 10,000-barrel trains. The gas plant was sized, equipment duties defined, and sufficient detailed engineering performed so that prices could be obtained.

The alkylation plant was estimated by Parsons based on experience for other similar facilities.

The intermediate and final product storage, and blending facilities were based on equipment sizing and preliminary engineering in sufficient detail to obtain preliminary cost estimates.

8.3 SENSITIVITY ANALYSIS

The sensitivity of the required selling price of gasoline is affected most by the fixed capital investment of the syngas-from-coal plant and the price of coal, while operating labor costs play an almost insignificant role in the final cost of making gasoline.

Table 8-2 presents values added in producing gasoline, and Figure 8-1 shows the effect of FCI and related charges on the required selling price of gasoline. For example, a 50% reduction in the fixed capital investment in the coal-to-syngas plant reduces the required selling price of gasoline approximately 18¢, whereas a comparable reduction in converting syngas to methanol or methanol to gasoline only reduces the price of gasoline 4¢ per gallon.

In the operating cost area, reducing coal costs (see Figure 8-2) also has a major effect on the required selling price of gasoline. For example, eliminating coal costs (now at \$20/ton) reduces the required selling price by 24¢ per gallon.

Figure 8-1 - Sensitivity: Required Product Selling Price to Fixed Capital Investments

Table 8-2 - Value Added^a
(cents/gallon)

PROCESS	OPERATING COSTS				CAPITAL BURDEN ^c	SULFUR CREDIT	TOTAL
	RAW MATERIAL COST	OTHER OPERATING COSTS	CAPITAL RELATED COSTS ^b	SUBTOTAL			
SYNGAS FROM COAL	22.7	7.0	9.4	39.1	26.0	(3.3)	62.4
METHANOL FROM SYNGAS	0.6	1.5	2.0	4.1	5.5		9.6
GASOLINE FROM METHANOL	1.1	1.6	1.9	4.6	5.3		9.9
TOTAL	24.4	10.1	13.3	47.8	37.4	(3.3)	81.9

^aBASED ON AN EQUIVALENT UNIFORM ANNUAL COST SERIES AT A 12% DCF RATE OF RETURN.

^bINCLUDES MAINTENANCE LABOR AND MATERIALS, PROPERTY TAXES, AND INSURANCE.

^cINCLUDES FIXED CAPITAL COSTS, 65/35 DEBT/EQUITY RATIO AT 9% LOAN INTEREST RATE, TAXES, AND PROFITS.

Figure 8-2 - Sensitivity: Required Product Selling Price to Cost of Coal

However, operating labor costs for the entire complex contribute only 2.4¢ per gallon of gasoline. This figure is based on 237 operators per shift for the overall complex.

These results clearly show the importance of selecting a low-cost, coal-to-gas conversion process and of negotiating the lowest possible coal price. Variations in the DCF rate of return as a function of the debt-to-equity ratio can contribute significantly to changes in the required selling price of gasoline (Figure 8-3 and Table 8-3).

Figure 8-3 - Sensitivity: Required Product Selling Price to DCF

**Table 8-3 - Required Gasoline Selling Price
(cents/gallon)**

RATE OF RETURN	DEBT/EQUITY RATIO	
	65/35 ^a	100% EQUITY
0	65	55
6	72	74
12	82	98
16	90	120
^a 9% LOAN INTEREST RATE.		

SECTION 9
RECOMMENDATIONS

SECTION 9

RECOMMENDATIONS

This report is a preliminary study of the potential economics of producing gasoline through three steps (1) coal gasification to produce synthesis gas, (2) conversion of synthesis gas to crude methanol, and (3) conversion of crude methanol to gasoline. This study does not necessarily represent optimum processing steps or conditions for any portion of the overall process. It is very likely that the economics can be improved by (1) better process optimization of existing processes and (2) the use of newer, improved processes and larger scale equipment.

The recommendations given below indicate areas in which improvements could lead to improved process economics.

9.1 PRODUCTION OF SYNTHESIS GAS

About 76% of the product cost is represented by the cost to produce the synthesis gas. The methanol synthesis and the methanol conversion steps share about equally in the remaining 24% of the product cost.

Also, approximately 70% of the fixed capital investment is in the preparation of the synthesis gas, with the remaining 30% divided approximately equally between methanol synthesis and methanol conversion.

The most promising area for product cost reduction lies in synthesis gas production.

A quick comparison with production of synthesis gas by a high-capacity, second-generation process currently under development by the Department of Energy indicates that the gasoline cost could be reduced from 82 to 59¢ per gallon. This reduction is caused by (1) higher thermal efficiency - 88.5% versus 69.7%, (2) larger gasifiers - two versus six used in this study, and (3) reducing the oxygen consumption to 0.65 lb per pound of coal.

9.2 METHANOL SYNTHESIS

The six methanol plants used in the study are in the same size range as the largest units in operation today. The technology of methanol production is well proven. Major breakthroughs in technology or costs are not to be anticipated. The one possible area of improvement is larger capacity plants of 5,000 to 10,000 TPD. Such plants could reduce the gasoline price by 3 to 5¢ per gallon.

9.3 GASOLINE PRODUCTION

The gasoline production portion of the plant contributes about 10¢ to the cost of the gasoline. There are areas where possible improvements are likely.

The first area is in the use of continuous, fluidized-bed methanol-to-gasoline reactors. This work is presently under development by Mobil. The continuous fluidized bed could result in more uniform product output (less cycling), better product yield, improved product quality (lower Durene content and better stability), and probably lower fixed capital investment. The savings here could be in the order of 2 to 5¢ per gallon.

A second possible area is to remove the water formed by the dimethyl ether (DME) reaction step immediately after methanol synthesis. The removal of this water would improve the ultimate catalyst life. The cost effect of this suggestion is expected to be less than 1¢ per gallon.

A third possible area is improved heat recovery in the methanol-to-gasoline conversion step. It is likely that additional high-pressure steam can be generated.

No significant savings are expected in the gas plant or alkylation plant since these are presently single-train units.

SECTION 10
LITERATURE CITED

SECTION 10

LITERATURE CITED

1. McGeorge, A., "Economic Feasibility Study, Fuel Grade Methanol from Coal," prepared by the Research and Development Division of the Industrial Chemicals Department, DuPont Company, Wilmington, Delaware. Prepared for ERDA under Contract No. 76-18, 1976.
2. Voltz, S.E., and J.J. Wise, "Development Studies on Conversion of Methanol and Related Oxygenates to Gasoline," Final Report, FE-1773-25, prepared by Mobil Research and Development Corporation, Paulsboro, New Jersey. Prepared for ERDA under Contract No. E(49-18)-1773, November, 1976.
3. Letter to The Ralph M. Parsons Company from Mobil Research and Development Corporation, dated June 17, 1977. This letter presents flow diagrams and tables listing the preliminary design conditions for the fixed bed unit.
4. Beavon, D.K., and J.E. Leeper, "Sulfur Recovery From Very Lean Hydrogen Sulfide." Paper presented at the 27th Canadian Chemical Engineering Conference, Calgary, Alberta, Canada, October 23 to 26, 1977.