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ECONOMICS OF GASOLINE PRODUCTION FROM UNDERGROUND COAL GASIFICATION VIA MOBIL-M PROCESS*

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ABSTRACT

A conceptual process design and cost estimate is presented for a facility producing approximately 100 m^3/h (15,000 barrels per day) of M-gasoline via methanol from synthesis gas generated by gasification of coal in situ. The design was based on experimental data obtained at the Laramie Energy Technology Center on the linked vertical well in situ coal gasification process. In-place coal consumption is 756 Mg/h (20,000 tons/day), based on a subbituminous Wyoming coal. The capital investment was estimated to be \$535 million in first quarter 1978 dollars. M-gasoline product price was calculated as a function of debt/equity ratio, annual earning rates on debt and equity, in-place coal cost, and plant factor (onstream efficiency). Using a debt/equity ratio of 70/30, an interest rate on debt of 9%, an after-tax earning rate on equity of 15%, an in-place coal cost of \$5.50/Mg (\$5/ton), an LPG (propane) by-product credit of \$3.80/GJ (\$4/10^6 Btu), and a plant factor of 90%, the product price of M-gasoline (including mixed butane LPG) is about \$240/m^3 (\$0.90/gal) at the plant gate. Calculated overall thermal efficiency for the facility was 22%, based on in-place coal.

Introduction

A promising method of utilizing coal which is currently inaccessible or uneconomical to mine is the "in situ" (underground) coal gasification process. Department of Energy (DOE) funded experiments with air injection gasifying Wyoming subbituminous coal underground have demonstrated technical feasibility on a small scale. These tests have been performed by Laramie Energy Technology Center and Lawrence Livermore Laboratory (LLL). If the injection of oxygen rather than air to gasify the coal underground proves successful on a continuous basis, a higher-Btu gas can be produced because large amounts of nitrogen diluent would not be present. Previous oxygen injection field tests in the Soviet Union and Poland and the successful injection of oxygen for 2 days during a recent, longer air injection experiment by LLL lend support to this concept. The higher-Btu gas available from oxygen injection can be treated and processed to yield high grade fuels or chemicals.

Oak Ridge National Laboratory (ORNL) has completed several evaluations of the potential for producing valuable end products from conceptual commercial "in situ" ventures.[1,2] This work was done for the Division of Systems Engineering of DOE/Fossil Energy. Today, I will be discussing our most recent evaluation, the production of gasoline from an in situ derived synthesis gas via the Mobil Methanol-to-Gasoline process.[2] This evaluation presumes the successful development of large-scale underground coal gasification with steam/oxygen injection.

Linked Vertical Well Process

There are several ways or modes in which the linked vertical well (LVW) process can be operated for large-scale field development and gas production.

These different operational modes arise primarily from variations in the well sequencing patterns used and the direction in which the coal seam is gasified relative to the direction of injection gas and product gas flow. The system assumed here is termed the direct-flow or forward system in the Russian literature[3] because the direction of gasification of the coal seam is the same as the direction in which the injection gas and product gas travel. The well sequencing pattern that develops is such that each borehole is used successively for linking, production, and injection.

In practice, a number of parallel lines of wells can be operated simultaneously to exploit large areas of coal seams. Such an arrangement is shown in Fig. 1.

The LVW process described is the one suggested by researchers at the Laramie Energy Technology Center (LETC) to be used for development of the field areas of the conceptual plant design evaluated in this report. It should be pointed out that large-scale operation of this system has not yet been demonstrated at LETC, although it was used by the Russians at the Podmoskovnaya and Shatskaya underground coal gasification stations. In the LETC operations to date, reverse combustion linking has been followed by air injection for forward gasification through the same well used for the linking air injection.

The reasons for choosing this procedure are: (1) it starts with a well sequencing pattern which is maintained through the life of the field, and (2) it simplifies the initial installation and subsequent moving of well header piping systems.

Synthesis Gas Production

The design basis for the linked vertical well in situ coal gasification process used in this evaluation was developed from information

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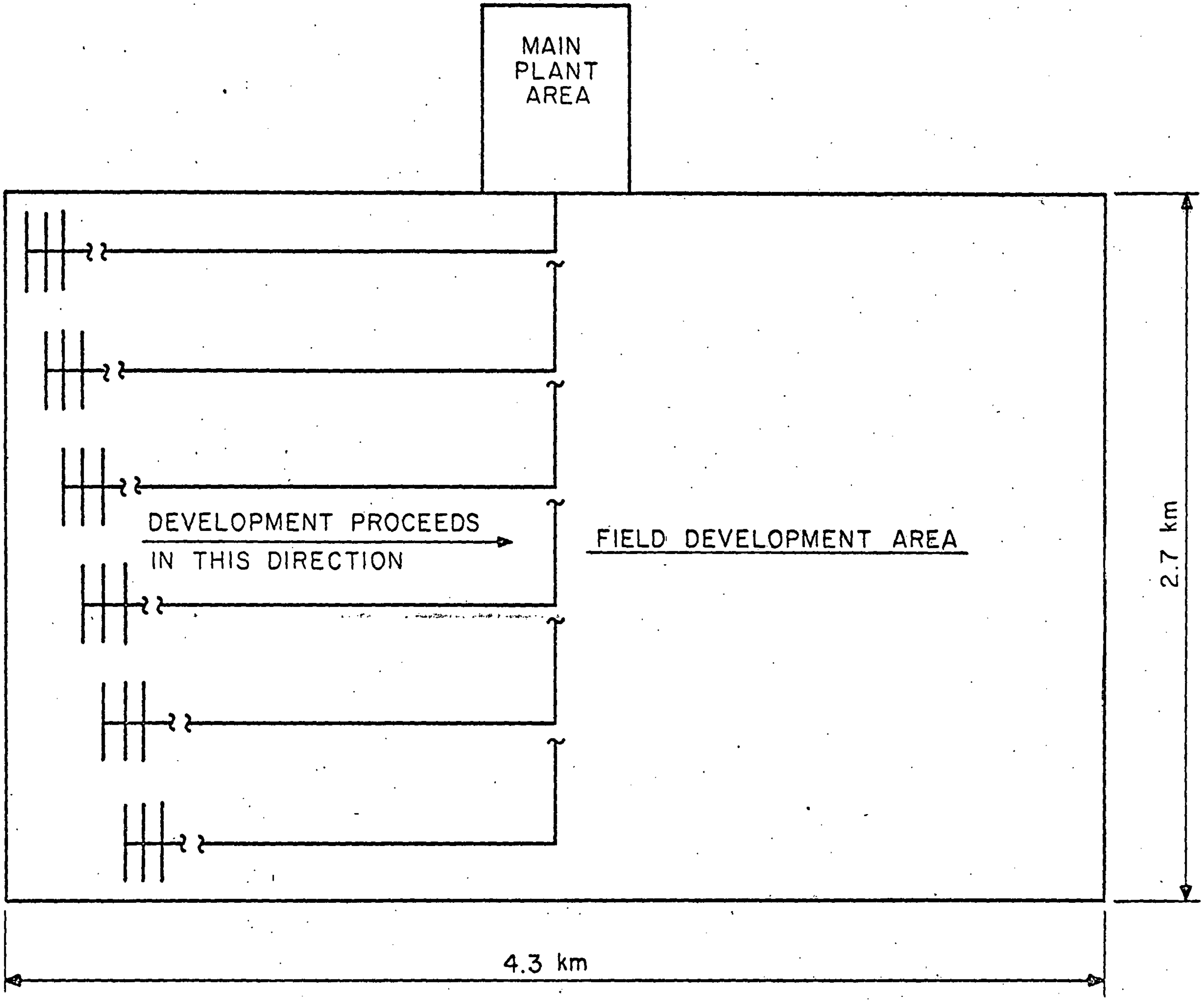

Fig. 1. In situ coal gasification facility layout.

obtained primarily at the Laramie Energy Technology Center.

To date, LETC has conducted four LVW field tests (Hanna I through IV), all with air injection. However, for generating a synthesis gas capable of being converted to methanol it is preferable to inject oxygen, and possibly steam, to gasify the coal. To date, very little work involving steam-oxygen injection for in situ coal gasification has been done in the United States. The Bureau of Mines conducted steam-oxygen injection experiments at Gorgas, Alabama in 1947 and 1952.[4] These data were not usable in our evaluation for the following reasons: the experiments were conducted in a thin (1 m) bituminous coal seam; oxygen injection runs were of short duration (e.g., 3 to 36 hours); product gas exhibited cyclic fluctuations in heating value; and inadequate experimental details precluded extrapolation of results. In November 1977, Lawrence Livermore Laboratory performed a successful two-day steam-oxygen injection experiment as part of a longer LVW air injection field test (Hoe Creek 2). Experimental data from this test were not available in time for use in this evaluation. Steam-oxygen gasification techniques have been extensively field tested in the Soviet Union and in Poland. Little information is available in the literature on the actual operation of these tests, however; usually only product gas quality is given.[5]

Because of the lack of experimental data for the steam-oxygen injection process, the basis for operating conditions and yields for this mode of gasification was a linear permeation mathematical model of forward gasification which was developed at LETC.[6-8] This model of the LVW process predicts gas compositions and heating values, gas production rates, temperatures, and thermal efficiencies.

The design basis parameters derived from the mathematical model and from experimental results of the Hanna II, Phase II field tests are shown in Table 1. Table 2 shows the composition of the raw gas produced with air injection (Hanna II, Phase II test data) and steam-oxygen injection (mathematical model prediction).

Based on Hanna II, Phase II results, LETC concluded that no underground gas leakage occurred. However, Russian experience indicates 8-10% gas loss and LLL estimated 10-30% leakage in Hoe Creek 2. Therefore we assumed a 10% raw gas loss due to leakage for this evaluation.

Field Development and In Situ Coal Gasification (Plant 1)

Operation of the facility at maximum design throughput of 120 m^3/h (18,000 BPD) of M-gasoline and LPG requires that 60 producing wells be on line. These 60 producing wells are arranged in six parallel trains of 10 wells each. Each train also requires 10 injection wells and 10 linking wells, so that a train consists of a total of 30 wells. The arrangement of trains in the field development area was shown in Fig. 1. The field development area is large enough to support six parallel trains for 20 years.

Initial production starts with only one train of wells in operation. The remaining five trains are brought on-line at intervals of roughly two weeks. A well has a producing lifetime of about 73 days. As each row of wells is exhausted, the train is moved to the next adjacent row. For a given train, these moves occur at 12-week intervals. Since there are six trains,

Table 1. LVW steam-oxygen gasification process design parameters

Type of coal	Subbituminous (Hanna No. 1 seam)
Seam thickness	9 m (30 ft)
Depth of seam	90 m (300 ft)
Well pattern and spacing	Square; 46x46 m (150x150 ft)
Gasification reaction zone advance rate	0.6 m/d (2 ft/day)
Process sweep efficiency	80%
Process thermal efficiency	80%
Overall process efficiency	64%
Steam/oxygen injection gas composition	60/40 mole %
Steam + oxygen injection rate requirement	30.66 mol/kg maf coal (23,270 scf/ton)
Steam-oxygen injection gas temperature	440 K (340°F)
Steam-oxygen injection gas pressure	0.62 MPa (75 psig)
Dry raw gas produced/ steam + oxygen injected	1.73 mol/mol[a]
Single well production rate	210 mol/s (15 MMscfd)
Raw gas wellhead temperature	610 K (640°F)
Raw gas wellhead pressure	0.47 MPa (54 psig)
Linking air injection rate	130,000 mol/m of link (33,000 scf/ft)
Linking air injection pressure	23 kPa/m of depth (1 psi/ft)
Reverse combustion linking rate	1.5 m/d (5 ft/day)

[a] Assumes 10% loss of raw gas due to underground leakage.

Table 2. Raw gas compositions (vol %)

Constituent	Injection gas: Air[a]	Injection gas: Steam/oxygen[b]
H_2	15.27	27.24
CO	13.58	23.59
CO_2	8.69	14.54
CH_4	4.24	5.16
N_2	42.00	0.62
O_2	0.00	
Ar	0.51	
H_2S	0.06	0.10
C_2^+	0.50	0.85
H_2O	15.00	27.63
Oil	0.15	0.27
Total	100.00	100.00
HHV, kJ/mol	143 (162 $\frac{\text{Btu}}{\text{scf}}$)	230 (261 $\frac{\text{Btu}}{\text{scf}}$)
Mol. wt.	23.84	26.58

[a]Hanna II, Phase II field experiment.

[b]Mathematical model prediction.

a move takes place every two weeks. Shortly after the sixth train is brought on stream, the first train is shut down. During the ensuing 14 days, the field equipment and piping used by the first train are disconnected, moved, and reconnected to the next row of wells, and production from this train is resumed. Each of the six trains follows this same cyclic pattern of relocation.

Therefore, at any one time, only five of the six trains are normally operating while the sixth train is being moved. Thus, the main plant normally receives raw gas from only five of the six trains of field wells. However, it is necessary to design the facility to be capable of handling the entire six train output, equivalent to the production of 120 m^3/h gasoline and LPG, because there will be brief periods when all six trains of wells are expected to be producing. This maximum production rate is the maximum design rate for the facility. To provide a conservative estimate of the potential of M-gasoline from in situ coal gasification, the base case economics are predicated on a five train production rate, equivalent to 100 m^3/h gasoline and LPG; this is the normal rate expected when the facility is operating.

To investigate the savings resulting from more effective field operation (longer well life, shorter well moving time, quicker well startup, etc.), the product price is also presented assuming that an average production rate of 110 m^3/h gasoline and LPG is achieved, representing an average of 5.5 trains producing. It is likely that this rate could be achieved through improvement of field procedures.

Alternatively, a seventh train of wells could be added to the field development area, ensuring that the maximum design rate could be consistently maintained when the facility operates. Product price for this alternative is also given.

Block Flow Diagram

The block flow diagram for the M-gasoline facility is shown in Fig. 2. Raw gas from the wells is transferred to plant 3 where it is cooled and scrubbed. Part of the scrubbed gas from plant 3 is sent to fuel gas treating (plant 9) to provide energy to operate the facility.

The remaining process gas from plant 3 is compressed in plant 4 and fed to CO shift reactors (plant 5) to adjust the H_2/CO ratio for the methanol synthesis reaction. For methanol production, the desired ratio of H_2 to CO is 2.17. Therefore, 32% of the CO in the feed gas must be reacted with steam to produce additional hydrogen (and CO_2). Sulfur-resistant cobalt molybdate catalysts have sufficient activity at high-temperature shift conditions to permit 32% conversion of CO in a single stage. The advantage of using such catalysts is that H_2S does not have to be removed prior to shifting.

After shifting, and prior to methanol synthesis, the shifted gas requires treatment to remove essentially all H_2S and most CO_2. A number of processes are available for removing these acid gases.

Direct use of the Stretford process, which absorbs and reduces H_2S to sulfur in an aqueous solution of sodium carbonate and various additives (vanadate salts, organic acids, etc.), was precluded by the very high CO_2 partial pressure.[9] However, the Stretford process is applicable to the treatment of regenerated acid gases (at much lower CO_2 partial pressure) evolved from other acid gas removal processes. This function was included in the fuel gas treating circuit in plant 9 by adding a parallel absorber to that required to remove H_2S from the fuel gas.

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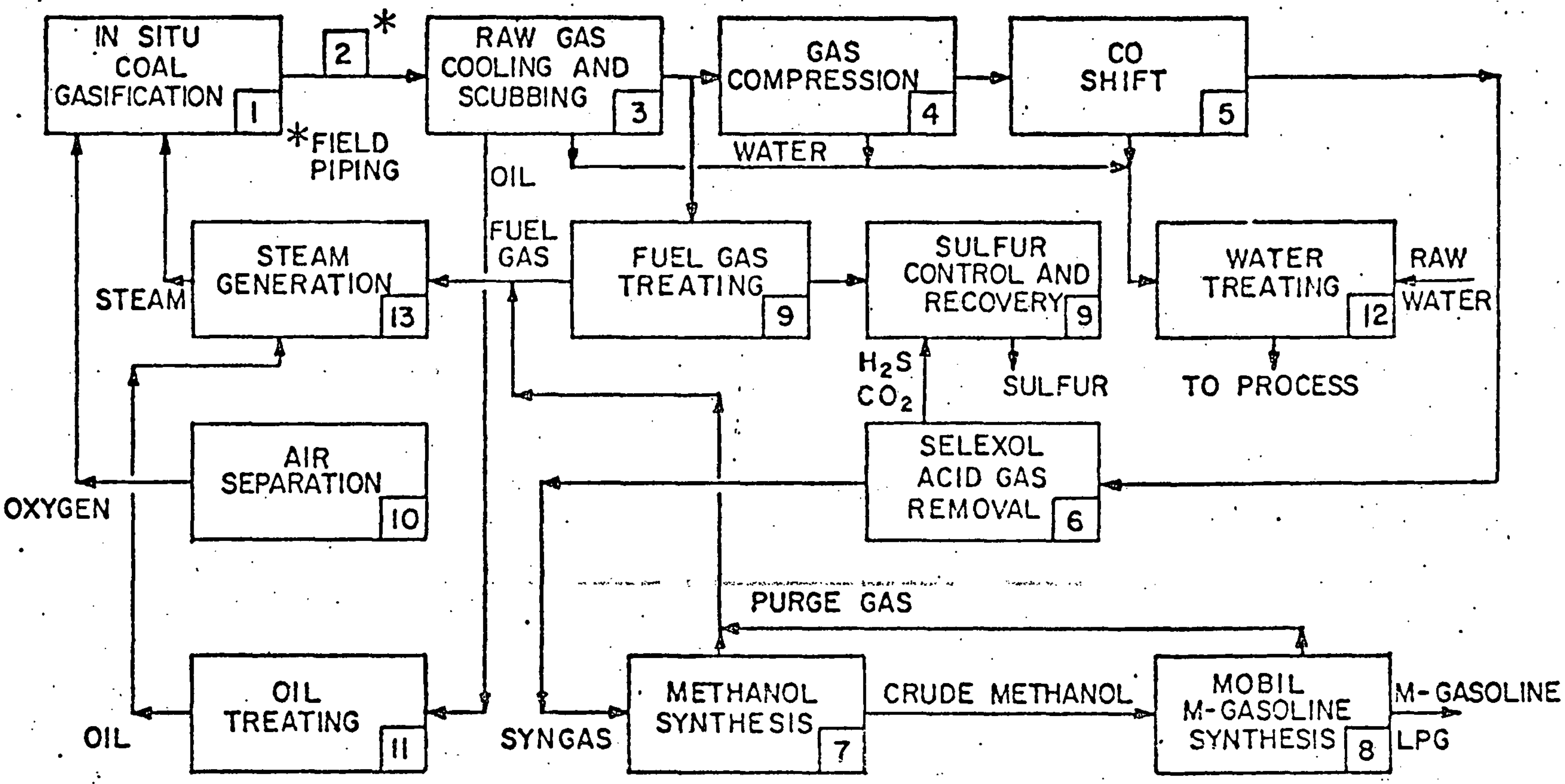

Fig. 2. Block flow diagram for M-Gasoline facility.

A preliminary assessment indicated that two acid gas removal processes would be suitable: The Benfield process (licensed by the Benfield Corporation), and the Selexol* Solvent process (licensed by Allied Chemical Corporation). However, the Benfield process requires considerably more reboiler steam, a characteristic of all chemical absorption systems. The Selexol process relies on physical absorption, which is enhanced at the high operating pressure, and regenerates much of the dissolved gas by simple pressure reduction. For the purposes of this study the Selexol process appeared more suitable.

The resulting sweet gas is converted to methanol in plant 7. Methanol synthesis proceeds by the reaction of hydrogen and carbon oxides over a catalyst according to the following stoichiometry:

$$CO + 2H_2 \rightleftarrows CH_3OH$$

$$CO_2 + 3H_2 \rightleftarrows CH_3OH + H_2O$$

Present reactors can operate at 5-10 MPa (750-1500 psig) and below 535K (500°F). The feed gas must, however, be essentially sulfur-free to prevent rapid deactivation of the copper catalyst.

Since the introduction of low pressure methanol synthesis in 1966, three commercial processes have received attention. Between 1968 and November 1977, the Imperial Chemical Industries process has accounted for about 87% of ordered and/or installed low pressure world methanol plant capacity.[10] The Lurgi process accounts for 10.5%, and Mitsubishi Gas Chemical Corporation is responsible for the remaining 2.5%.

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*Selexol is registered trademark of the Allied Chemical Corporation.

The principal USA licensees of the ICI process are Davy Powergas, Pullman-Kellogg, and Foster Wheeler Energy Corporations. Davy Powergas was contracted to perform an engineering study for ORNL of a conceptual low pressure ICI methanol plant. Under terms of the contract, Davy Powergas provided the plant material balance, utility balance, capital costs, and operating charges. Because the ICI process is proprietary, no internal details of the process were disclosed.

Crude methanol is transferred to M-gasoline synthesis (plant 8) where it is converted to gasoline and by-product LPG. Mobil Research and Development Corporation (under a Department of Energy contract) has been developing technology for the conversion of methanol to highly branched hydrocarbons suitable for use as gasoline. The technology is referred to as the M-Gasoline process, an abbreviation of methanol-to-gasoline. Similarly, the end product of this process is generally termed M-gasoline. The reaction can be simply represented as:

$$xCH_3OH \rightarrow (CH_2)_x + H_2O \; . \qquad (1)$$

Mobil Oil has been producing zeolites as cracking catalysts for the past 20 years. One of these was discovered to catalyze the reaction in Eq. (1). In contrast to the zeolite (named faujasite) used as a petroleum cracking catalyst, which has channels about 1 nm in diameter, the new zeolite has channels about 0.5 nm in diameter and is very selective for linear paraffins. Using this catalyst, Mobil was able to convert methanol to products in the gasoline range. Because this conversion is highly exothermic the conversion is done in two steps to aid in heat removal. First, methanol is dehydrated to dimethyl ether in a DME reactor. Then the dimethyl ether is further dehydrated to olefins that in turn are converted to aromatics and paraffins in the M-gasoline reactor.

Mobil supplied ORNL with basic M-Gasoline process design information that was developed under their DOE contract. The information included an overall material balance, flowsheet, approximate equipment sizing, and HF alkylation unit duties and utility requirements.[11] ORNL modified the flowsheet, equipment specifications, and material balance to reflect the differences in the crude methanol fed to the M-Gasoline plant. The resulting process design was then developed in sufficient detail to permit reasonable utility and cost estimates to be made. For example, the M-gasoline product cooling train was carefully reviewed and modified to ensure confidence in its operability.

Treated fuel gas and purge gases from plants 7 and 8 are burned in the steam boilers in plant 13 to provide steam for injection in the field, generation of electricity, and other uses such as steam-turbine driven gas compressors.

Air separation in plant 10 provides oxygen for injection in the field plus inert gas (N_2) for blanketing storage tanks and other uses.

Oil removed in plant 3 reports to oil treating (plant 11). After treating, it is used to supplement fuel gas supplied to plant 13. Water treatment in plant 12 is also included in the facility. Here raw water is prepared for use in the facility and waste water is treated for reuse.

General Facility Requirements

Products

M-gasoline is produced as a blend of all the stabilized gasoline, all the alkylate, and a small amount (30%) of the mixed butanes made in the facility. The blend, 93.4 wt % stabilized gasoline, 3.4 wt % alkylate, and 3.2 wt %

mixed butanes, has research, motor, and average octane numbers of 92.9, 82.9, and 87.9, respectively.[12] A durene content of 4.8 wt % is expected in the product. It is assumed that the M-gasoline product is sent by pipeline to refinery and/or distribution centers, suitable for sale or additional blending.

The remaining 70% of mixed butanes produced is suitable for marketing as butane LPG. The butane LPG is mostly isobutane (74.5 vol %) and normal butane (25.1 vol %) with a trace of isopentane and pentene. The high isobutane content should permit sale as a potential alkylation feedstock as well as a commercial butane fuel. Transportation by rail and/or truck tanker is assumed.

The propane LPG released from the HF alkylation unit is also assumed to be shipped by rail and/or truck tanker. The product composition (97.8 vol % propane, 1.6 vol % isobutane, and only 0.6 vol % ethane) meets current commercial requirements for propane LPG.

Sulfur can be shipped by tank car in liquid form. Provision is made for storing it in solid form if market demand is weak. Sulfur, produced by the Stretford process, is expected to be of high purity (99.5 wt %).

Installed spares

Installed spare pumps and motors are provided, generally to the extent of at least 50%. Davy Powergas, Inc. indicated that $1.5 million was required for assuring adequate equipment sparing in the methanol synthesis plant (plant 7). This was added to the Davy Powergas capital cost estimate.

There are no installed spares for other types of equipment; however, overcapacity is provided in critical areas. Spare rotors are provided for centrifugal compressors.

Utility systems

The major utility systems for the facility include steam, electricity, fuel gas, fuel oil, and cooling water. Other systems are provided for boiler feed water, steam condensate, and process water. Utilities generation and consumption are summarized in Table 3 at maximum design capacity.

Fresh water (raw water) is assumed to be purchased. All other utilities required by the facility are generated on site.

Large centrifugal process gas compressors are turbine driven using steam produced in steam boilers in plant 13. During startups when fuel gas will not be available, oil will be used to raise the required steam.

Small compressors, air cooler fans, and most pumps are electric motor driven. Electricity is provided by turbine generators in plant 13 using steam from the 6.2 MPa (900 psia) steam generator units. Some additional electricity is generated in plants 7 and 10 by power recovery turbines.

Process cooling is provided both by air and cooling tower water. Forced-draft wet cooling towers are used. Air coolers are equipped with fans driven by electric motors. The use of air cooling was maximized because of the expense and scarcity of makeup water for the cooling tower circuit.

Thermal efficiency

The overall thermal efficiency calculated for the M-gasoline facility evaluated in this study is 22%. This efficiency was calculated as the heating value of the hydrocarbon products (gasoline and LPG) divided by the heating value of the in-place coal. Higher heating values were used. No credit was taken for by-product sulfur in this calculation.

The overall efficiency of 22% includes an assumed efficiency of 64% for the underground gasification process (see Table 1 in the process discussion).

Table 3. Plant utility balance. (Values rounded to nearest 5 units).

Utility		Plant no. 1	2	3	4	5	6	7	8
Elec. (kW)	produced							6,570	
	consumed	2,000		2,940	5,425	1,085	24,355	3,675	3,220
Steam (kg/h)									
6.2 MPa (900 psia)	produced								
	consumed				426,665	189,900		141,045	102,810
0.8 MPa (120 psia)	produced					49,050			
	consumed	197,435			770				125
0.3 MPa (50 psia)	produced					91,225			
	consumed						35,510		2,120
Water (m^3/h)									
Cooling	produced								
	consumed				410				385
Process	produced			175	35	200		390	115
	consumed						15	390	
HP BFW	produced							625	
	consumed					60		625	
LP BFW	produced								
	consumed					140			
HP Cond.	produced				425			140	1,345
	consumed								1,240
LP Cond.	produced						35		5
	consumed								
Raw	produced								
	consumed								
Fuel (GJ/h)									
Gas	produced							2,445	130
	consumed								10
Oil	produced	50		1,015					
	consumed								

Table 3. (continued)

Utility		Plant no. 9	10	11	12	13	14	Total
<u>Elec. (kW)</u>	produced		1,620			61,475		69,665
	consumed	2,475	745	105	1,500	7,300	14,840	69,665
<u>Steam (kg/h)</u>								
6.2 MPa	produced					1,483,990		1,483,990
	consumed		389,660			233,910		1,483,990
0.8 MPa	produced			4,080		233,775		286,905
	consumed							198,330
0.3 MPa	produced							91,225
	consumed	1,060			141,110			179,800
<u>Water (m^3/h)</u>								
Cooling	produced						36,365	36,365
	consumed		35,570					36,365
Process	produced	20			855			1,790
	consumed	20			530		835	1,790
HP BFW	produced				1,545			2,170
	consumed					1,485		2,170
LP BFW	produced				380			380
	consumed			5		235		380
HP Cond.	produced		390			235		2,535
	consumed				1,295			2,535
LP Cond.	produced				140			180
	consumed				180			180
Raw	produced							0
	consumed				770			770
<u>Fuel (GJ/h)</u>								
Gas	produced	1,135						3,710
	consumed					3,700		3,710
Oil	produced			1,060				2,125
	consumed			1,065		1,060		2,125

This is made up of two components: an assumed sweep efficiency of 80% and an assumed process thermal efficiency of 80%. That is, it was assumed that 80% of the in-place coal was gasified, and 80% of the thermal content of the gasified coal was recovered (neglecting gas leakage). In addition, the loss of 10% of the raw gas caused by leakage is included in the overall efficiency. Deletion of these effects permits the determination of surface plant thermal efficiency as 38%.

Fig. 3 presents graphically the thermal losses associated with major steps in the M-gasoline facility, referred to the thermal content of in-place coal. Much of the loss of the in-place coal thermal content occurs before the raw gas reaches the production well. The high thermal loss associated with cleaning the raw gas and converting it to a compressed synthesis gas results not only from the low delivery pressure, 0.39 MPa (56 psia), of the raw gas to the main plant but also from the inclusion of all utility requirements for the facility within this segment. Because of the many purge and fuel streams, it would be difficult to properly apportion the utilities otherwise.

Capital investments

Table 4 summarizes the depreciable capital investment requirement for the M-gasoline facility. The total includes capital investments for various plant sections, engineering, construction overhead, contingencies, contractor's fees, and special charges (royalties, taxes, etc.).

The capital investment shown here does not include the cost of the coal (or land and mineral rights) required for the facility. Coal is charged as a raw material as part of the operating costs. The cost is treated as a variable in the economic calculations.

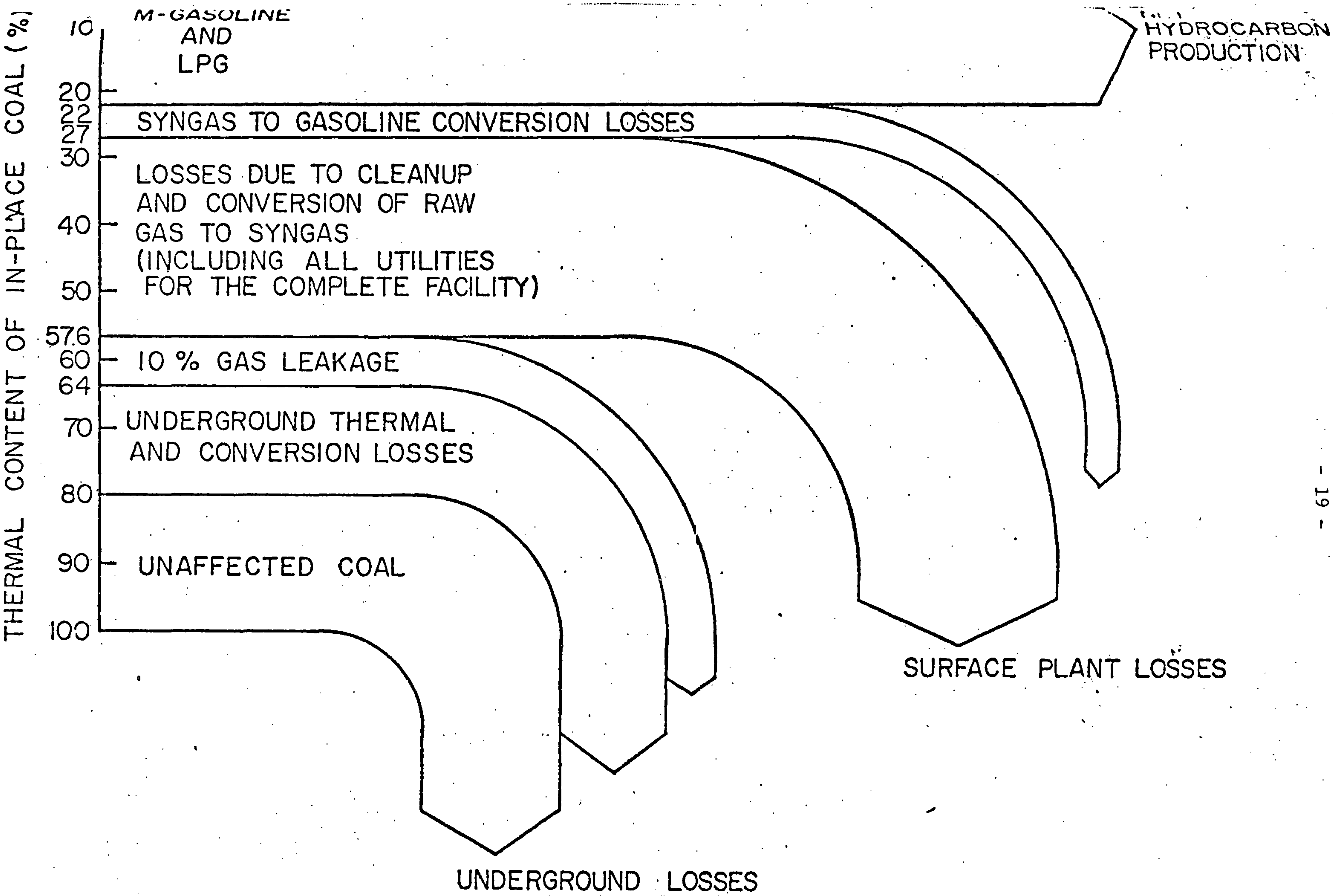

Fig. 3. Thermal losses in underground coal gasification to M-Gasoline process.

Table 4. Depreciable capital investment summary for 100 m^3/h M-gasoline facility.

	Capital investment $\$10^6$
Site development	4.397
Field gas treating plant	16.749
Field piping system	28.278
Raw gas treating plant	13.705
Gas compression plant	33.604
CO shift plant	6.887
Selexol plant	22.421
Methanol plant	55.034
M-gasoline plant	20.908
Stretford plant	5.354
Air separation plant	83.303
Oil treating plant	1.302
Water treating plant	6.199
Steam and power plant	35.095
Offsites	60.551
Capitalized drilling costs	2.970
Engineering	6.889
Construction overhead	28.018
Contingencies	63.420
Contractor's fee	12.685
Special charges	27.006
Total capital investment	534.775

All costs given here are based on first quarter 1978 prices and are expressed in first quarter 1978 dollars. No escalation beyond this time is included.

Cost estimates for various parts of the facility were obtained from several sources. Package plant costs (sulfur control, air separation, HF alkylation) were obtained from the literature. A capital cost estimate for the methanol synthesis plant, including initial catalyst charge, was obtained through a contract with Davy Powergas, Inc.[13] The ICARUS CØ$T[R] computer program[14] was used to generate cost estimates for the gas cooling and scrubbing plant, the gas compression and CO shift plants, the M-gasoline plant, and other sections of the facility not covered elsewhere, such as oil and water treating, utilities generation, offsite facilities, tankage, yard piping, electrical substation, fire protection, etc.

Initial well drilling and preparation work which occurs during the plant construction period is included in plant capital costs. After the plant is started up, this cost is included as an operating charge. It was decided to use \$164/m (\$50/ft) as the estimated total cost per foot for well drilling including casing. The estimated sensitivity of M-gasoline product price to increasing drilling costs to \$246/m (\$75/ft) is discussed in the economic analysis.

A contingency allowance was provided for the facility at 15% of bare plant cost. The bare plant cost includes the capital required for the various plants, engineering, and construction overhead. The contractor's fee was assumed to be 3% of the bare plant cost. The special charges include freight, state sales tax (at 3%), and process plant royalties and licensing fees.

All of the above charges represent depreciable capital costs. In addition, nondepreciable charges for the facility include working capital and initial catalyst and chemical charges. Working capital was estimated at $32 million, representing 6% of total depreciable capital. This allows for about one month's funds for raw materials, operating supplies and chemicals, payroll and payroll burden, and product inventory. Spare parts and other warehouse stock are also included. The effect on product price of increasing working capital to 9% of total depreciable capital is discussed in the economic analysis.

The initial catalysts and chemicals are estimated to cost $11.2 million. Costs for commercial quantities of the M-gasoline synthesis plant catalysts are not available because they have been manufactured in only limited amounts. Therefore a cost of $11/kg ($5/lb) was assumed for the base case. The effect on product price of M-gasoline catalyst at $22/kg is also examined in the economic analysis.

The cost for land rights was not estimated as a capital charge. Instead, land and mineral rights were included in the cost of coal as an operating charge. This approach is discussed under operating costs.

<u>Operating costs</u>

The operating costs described here do not include depreciation (recovery of capital), interest on debt, return on investment, or taxes. These costs are accounted for internally by the overall economics program, as discussed in the economic analysis.

Utilities are not included here because the facility generates its own utilities, except for water. Raw water cost was assumed to be $0.26/m^3$ ($1.00/1000 gal) delivered at the plant gate. In the economic analysis, product price sensitivity to variations in raw water cost are examined.

To calculate operating labor and supervision costs, the total operating manpower requirement of the facility was estimated. The "home office" component of overhead has been omitted from the overall product price calculation. The product price thus corresponds to a "plant gate" price for M-gasoline. Marketing, distribution, and home office administrative costs would have to be added to arrive at a final sales price, but no attempt was made to calculate this.

The cost of coal used by the facility was charged as an operating expense. The coal cost was varied parametrically from 0 to $11/Mg. This cost refers to coal in place, not net coal gasified, and is assumed to include all charges for the coal. This approach was adopted because of the present state of uncertainty in land and lease right costs, royalties, state severance taxes, etc. To calculate the coal cost in $/yr, the cost was multiplied by the amount of coal in place in the field area exploited in one year.

Field equipment moving expenses are based on the estimate that every train of wells is moved once per quarter. Every 90 days there is a moving cost based on a percentage of material and labor costs for the initial installation. An additional quarterly cost for labor and equipment used in moving field systems is also applied.

General economic assumptions used in preparing the operating cost estimate for the facility include the following:

- Plant operating lifetime: 20 years
- Construction period (pre-operational period): 3 years
- All costs are based on first quarter 1978 prices and are expressed in first quarter 1978 dollars. No escalation beyond this time is included.

- Maintenance is estimated at 4% of depreciable capital per year.
- Plant factor (operating factor) is 90% for the base case.
- Direct labor rate is $8.75/hr.
- Labor burden is 35% of direct labor.
- Supervision is 15% of labor plus labor burden.
- Operating supplies are 30% of direct operating labor.
- Overhead is 135% of direct labor plus supervision.
- Federal income tax rate is 48%.
- No state income tax.
- Local taxes and insurance are 3% of capital per year.

Table 5 gives the operating costs for the facility in two general categories. The first, proportional expenses, are those which vary according to product throughput, e.g., raw water, chemicals, etc. The second category includes expenses such as operating labor, overhead, etc., which are assumed to remain constant.

Economic analysis

Product costs were based on the cost of production per barrel of M-gasoline and include coal costs, operating costs, recovery of investment, return on investment, and taxes.

Annual after-tax rate of return on equity was treated as a parameter using rates of return of 10, 12, 15, and 17%. The annual interest rate on debt was assumed to be 9%. The base case calculations were done for a capital structure of 70% debt, 30% equity. The effect of various changes in base case parameters on product price were examined.

Table 5. Operating costs

	$\$10^6$/year
Proportional Expenses[a]	
Catalysts and chemicals	5.381
Raw water	1.489
Drilling	3.960
Field move	6.867
Miscellaneous	1.000
	~~21.657~~ 18.697
Constant Expenses	
Operating labor	4.004
Labor burden	1.401
Supervision	0.811
Operating supplies	1.201
General, admin. overhead	6.500
Maintenance material and labor	21.400
	35.317

[a]Based on 100% onstream factor.

The method used for the calculations of product costs was based on the principle that the income from product sales must be sufficient to recover invested capital, pay all operating expenses and taxes, pay the interest on debt capital, and earn the required rate of return on equity. A computer program[15] was used to perform these calculations. The method used is mathematically equivalent to the discounted cash flow procedure.

Input data for the base case described above are summarized in Table 6 .

Depreciation for tax purposes was calculated by the sum of the years' digits method. A depreciable life of 16 years was assumed, starting at the end of the construction period.

The product prices calculated for the base case parameters listed in Table 6 are shown in Fig. 4 . In this figure, the debt/equity ratio, plant factor, and tax structure are fixed, and the product price is plotted as a function of the cost of coal, with the annual after-tax rate of return on equity as a parameter.

Table 7 gives a summary of the results of the sensitivity analyses for variations in a number of base case parameters. The base case product price is provided for comparison. Product prices for all cases are shown for coal cost of \$5.50/Mg (\$5/ton) and after-tax rate of return on equity of 15%/year. The parameter changes studied are also indicated in this table. In performing the sensitivity analyses, only one parameter at a time was varied from its base case value to determine the effect on product price. For example, in case 11, the cost of raw water was reduced 50%. With other parameters remaining at their base case values, it is seen that the product price of M-gasoline decreased to \$239.60/m^3 from \$240.60/m^3 at a coal cost of \$5.50/Mg

Table 6. Base case parameters for in situ M-gasoline facility evaluation

Tax losses credited to parent corporation	
Investment tax credit =	10%
Debt/equity ratio =	70/30
Annual interest rate on debt =	9%
Annual rate of return on equity =	10,12,15, and 17% (after tax)
Constant operating costs =	$35.317 x 10^6/year
Proportional operating costs =	$21.657 x 10^6/year
Mobil-M catalyst =	$11/kg ($5/lb)
Drilling, casing, and logging =	$164/m ($50/ft)
Raw water =	$0.26/m^3 ($1/1000 gal)
Coal cost =	$0-$11/Mg (in place)
Depreciable capital investment =	$535 x 10^6
Working capital =	6% of depreciable capital
Nominal plant capacity =	100m^3/d M-gasoline and LPG
Construction period =	3 years
Plant lifetime =	20 years
C_3 LPG product price =	$3.80/GJ ($4/10^6 Btu)
C_4 LPG product price =	M-gasoline price
Sufur credit =	$59/Mg ($60/LT)
Plant factor =	90% for entire 20-year lifetime

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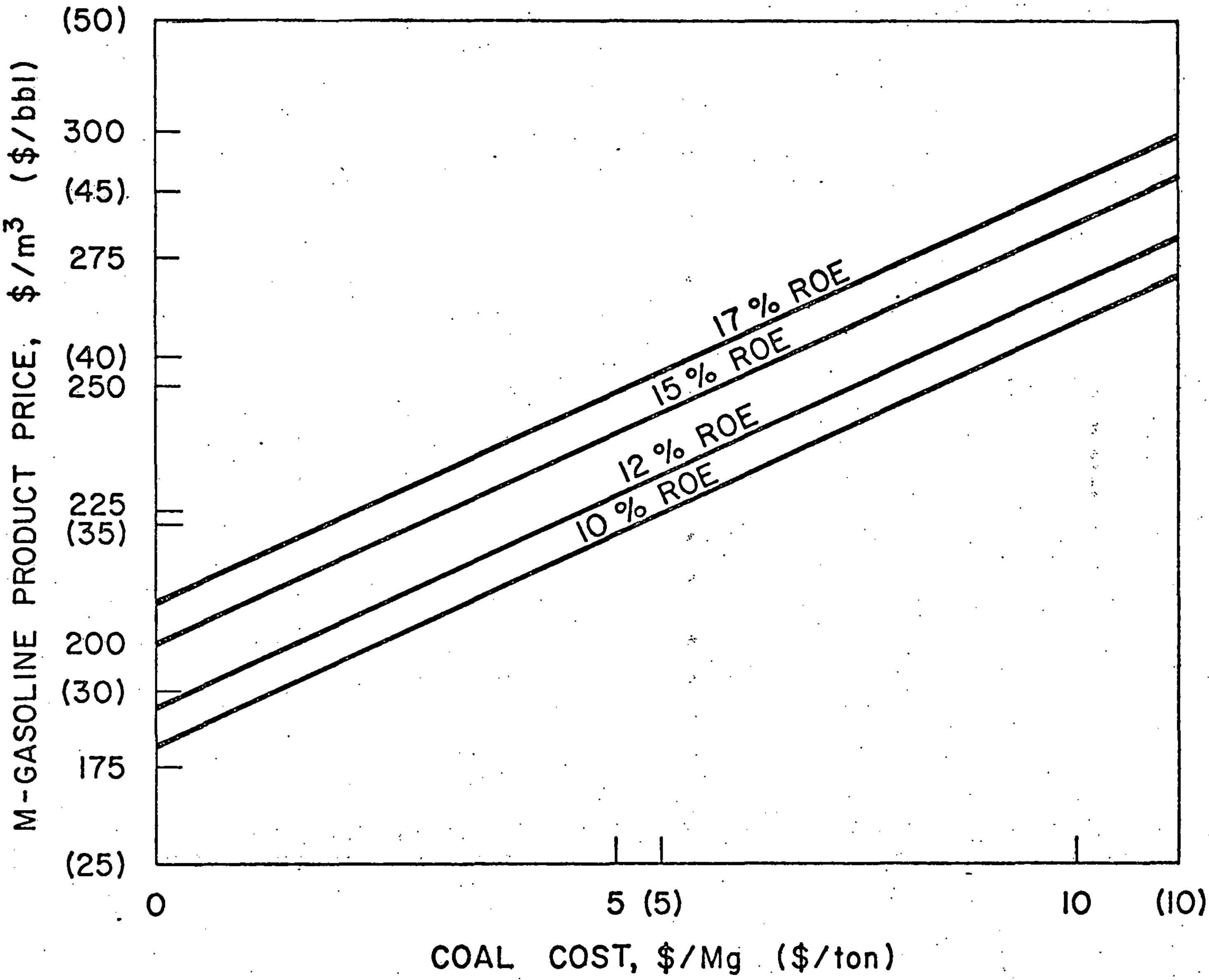

Fig. 4. M-gasoline product price under base case conditions for varying coal costs and after-tax rates of return on equity (ROE).

Table 7. Summary of M-gasoline product price sensitivity analyses.

No.	Variations for sensitivity analyses	M-gasoline product price ($/gal)	($/m^3)
Base,	with coal cost = $5.50/Mg and 15% ROE	0.91	240.60
2	No parent corporation tax credit	0.98	258.80
3	No investment tax credit	0.96	253.50
4	100% equity	1.34	353.40
5	30/70 debt/equity ratio	1.14	300.30
6	10% annual interest rate on debt	0.93	245.30
7	8% annual interest rate on debt	0.89	235.90
8	+10% operating (constant and proportional) costs	0.94	248.00
9	Mobil-M catalyst = $22/kg ($10/lb)	0.92	243.00
10	Drilling, casing and logging = $246/m ($75/ft)	0.92	243.30
11	Raw water = $0.13/m^3 ($0.50/1000 gal)	0.91	239.60
12	+10% capital cost	0.97	256.20
13	Working capital = 9% of depreciable capital	0.92	243.90
14	C_3 LPG product price = $4.74/GJ ($5/10^6 Btu)	0.90	239.00
15	C_4 LPG product price = $5.69/GJ ($6/10^{6}Btu)	0.94	247.80
16	No sulfur credit	0.91	241.30
17	Plant factor = 80% for entire 20-year lifetime	0.99	262.70
18	Plant factor = 75% in first operating year, 90% for remaining 19 years	0.92	243.40
19	Increase to 7 field trains => 120 m^3/h	0.81	212.90
20	Increase actual production to 110 m^3/h	0.85	224.50

and an after-tax rate of return on equity of 15%/yr. The various cases are briefly described as follows:

Case 2: It is assumed that there is no parent corporation available to take advantage of tax losses attributable to the M-gasoline facility.

Case 3: No investment tax credit is taken.

Case 4: 100% equity funding is assumed.

Case 5: Financing is assumed to be 30% debt, 70% equity.

Case 6: The annual interest rate on debt is taken as 10%.

Case 7: The annual interest rate on debt is taken as 8%.

Case 8: Proportional and constant operating expenses are increased by 10% over the base case.

Case 9: The assumed cost of catalysts for the Mobil methanol-to-gasoline conversion reactors is increased to \$22/kg (\$10/lb).

Case 10: The cost for drilling, casing and logging the field wells is increased to \$246/m (\$75/ft).

Case 11: The cost of raw water delivered to the plant gate is reduced to \$0.13/m^3 (\$0.50 per 1000 gal).

Case 12: The depreciable capital cost is increased by 10% over the base case.

Case 13: Working capital is increased by 50%, from 6% to 9% of depreciable capital.

Case 14: The value of by-product propane LPG is increased by \$0.95/GJ to \$4.74/GJ (\$5/10^6 Btu).

Case 15: The value of exported butane LPG is fixed at \$5.79/GJ (\$6/10^6 Btu), rather than being priced the same as product gasoline.

Case 16: No credit is taken for the sale of by-product sulfur.

Case 17: An onstream operating efficiency of 80% is assumed for the entire 20 year plant lifetime.

Case 18: The onstream efficiency is reduced to 75% for the first year of operation. For the remaining 19-year plant lifetime, a 90% efficiency is assumed.

Case 19: A seventh train of field wells is added, permitting operation of the facility at maximum design capacity, 120 m^3/h (18,000 BPD) gasoline and LPG.

Case 20: Field production is increased to an average of 5.5 trains operating out of 6 trains. This increases plant production to 110 m^3/h (16,500 BPD) gasoline and LPG.

Remaining concerns

Several unresolved questions about "in situ" coal gasification remain to be answered before these large commercial facilities can be built. The major environmental concerns are ground subsidence and disruption or contamination of surface and subsurface water. Large scale experiments will be required to determine the extent and nature of these problems and, hopefully, will suggest solutions. Water contamination caused by leaching of toxic materials remaining in the gasified coal seam has received some attention and will require more to ensure that "in situ" coal gasification is compatible with the environmental goals of the nation.

A process-related concern is the ability of underground gasification to consistently and controllably produce gas at the desired rate and heating value over the 20 year lifetime of a commercial facility. This is complicated by the requirement for moving the wells used for air (or oxygen) injection and gas production every few weeks to access new segments of the coal seam as the coal gasification continues. DOE experiments have so far been limited to a single air injection well and single gas production well. Experiments on multiple wells demonstrating the ability to move from one set of wells to the next set further along the coal seam are needed.

In summary, "in situ" coal gasification offers the promise of producing valuable products such as M-gasoline from coal seams which are not economical to mine. Further experimentation is required, however, to answer the concerns remaining for large scale commercial application.

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