EXECUTIVE SUMMARY

A study of the economics of liquid fuels production by Fischer-Tropsch (F-T) processing was initiated for two purposes. The first was to provide assistance in assessing the near to intermediate term potential for commercial production of liquid fuels by indirect liquefaction based on relatively cheap remote natural gas. The second was to examine the economics of slurry F-T processing versus the economics of two existing processes, entrained bed Synthol and fixed bed ARGE.

The principal results of this study are:

(1) None of the cases appears profitable for a nominal 40,000 BBL/D facility coming onstream in 1990, regardless of the value assigned to natural gas.

(2) The slurry case appears to have an advantage over Synthol and ARGE. This advantage is primarily rooted in the relative capital costs of the F-T synthesis areas.

These results are discussed further below.

Synthesis Gas Production

The initial step in this evaluation was to choose the appropriate synthesis gas production process for each F-T process. The plant size for this commercial production facility was chosen as 875 MM SCFD of synthesis gas for each case. This choice was based on some recently published studies of commercial F-T plants. Both steam methane reforming followed by a Prism membrane and partial oxidation of the natural gas were investigated. The results indicated that for a natural gas price of \$1.00/MMBTU, the slurry and ARGE synthesis gas compositions could best be supplied by partial oxidation of the gas while the Synthol synthesis gas composition could best be supplied by steam methane reforming. The synthesis gas production economics are tabulated below for plants coming onstream in 1990.

	Slurry	Synthol	ARGE
Molar H_2/CO Ratio Required	0.9	2.0	1.7
Preferred Process	POX	SMR	POX
Required Investment (MM \$)	\$871	\$453	\$772
Synthesis Gas Transfer Price (\$/MSCF)	\$1.44	\$1.15	\$1.22

Additional synthesis gas production cases based on coal gasification were also developed as a sensitivity. These cases were based on a midwestern coal valued at \$1.50/MMBTU. The synthesis gas from the gasification facilities costs nearly \$5.00/MSCF. The synthesis gas for the slurry case from coal was somewhat cheaper relative to the other cases since the low H_2/CO ratio requires less CO shifting.

F-T Synthesis and Product Upgrading

The next step in the evaluation was to develop economics for the F-T synthesis loop and the product upgrading sections. Computer simulations were performed for the synthesis loop for the Slurry, Synthol and ARGE cases. This enabled some preliminary equipment sizing to be done for this area.

While costs were developed for major pieces of equipment in the synthesis loop itself, costs were developed on a macro basis (i.e., total flow to a unit) for the product upgrading units. Since it was desirable to produce an all liquid product slate, C_1's and C_2's from F-T synthesis were sent to an autothermal reformer where they were processed to synthesis gas which would be sent to F-T synthesis for additional liquids production.

The results of the economic analysis are tabulated below. The table gives capital costs and annual revenues required for F-T synthesis and product upgrading for the three cases based on a nominal 40,000 BBL/D facility coming onstream in 1990. The required revenues provide for covering the costs plus a 7.5% net profit after tax on investment. Also tabulated are projected 1990 revenues and the revenue shortfall for each case. The projected revenues are based on product prices obtained from Corporate Energy.

	40,000 BBL/D 1990 Onstream		
	Slurry	Synthol	ARGE
Capital Investment (MM $)	1,270	1,955	1,810
Required Annual Revenues (MM $)	852	979	964
Projected Revenues (MM $)	495	481	495
Revenue Shortfall (MM $)	357	498	469
Revenue Shortfall ($/BBL Output)	25	36	34

Although the slurry case appears to be the most attractive of the three cases, the above tabulation predicts that none would be profitable. In fact, none would be profitable even if natural gas was assigned a value of zero. The projected 1990 revenues for each case are about $35/BBL. The revenues need to be about $60/BBL for the slurry case and about $70/BBL for the Synthol and ARGE cases. The advantage for the slurry case is based on a distinct capital cost advantage in the F-T synthesis loop.

The sensitivity cases based on coal gasification proved to be prohibitively expensive. For each of these cases a revenue shortfall in excess of one billion dollars annually was predicted for a 1990 onstream plant.

Total Plant Investment

The synthesis gas production and F-T synthesis and product upgrading sections have been treated separately in an effort to better illustrate the differences among the cases. The approximate total capital investment for the gas based facilities are as follows for 1990 onstream:

Slurry	$2.1 billion
Synthol	$2.4 billion
ARGE	$2.6 billion

Comparison with Mobil MTG and LNG

A comparison was also made involving the Mobil MTG process and the F-T processes. A revenue shortfall approximately equivalent to the slurry case was predicted for the MTG process on a dollars per barrel of output basis. Since it was not possible to ensure that the economic data for the F-T cases and the Mobil MTG case were on a totally consistent basis, some reservations must be expressed concerning the conclusion of economic equivalence between MTG and F-T slurry.

The distinction between indirect liquefaction and LNG appears to be more clear-cut. A recent SRI International (9) study indicated that $1.00/MMBTU natural gas could be liquefied, delivered and regassified for as little as $3.00/MMBTU in 1983 dollars, depending on the shipping distance. Escalation to a 1990 basis, with natural gas remaining at $1.00/MMBTU, would put the LNG costs in the $3.50-4.00/MMBTU range. These LNG values were corroborated by other information obtained from sources within APCI. By contrast, the cost of liquid product from the slurry case in 1990 is more than $11.00/MMBTU, excluding delivery. Transportation fuels should have a higher form value than gas, but this difference would not be enough to significantly close the above gap.

Recommendations for Future Work

Further study should involve the confirmation of the projected capital cost advantage of the slurry process. In addition, even though the product yield structure for the slurry case does not appear to be a critical factor in the relative economics, these yields should be confirmed by further experimental work since the current data base at the conditions chosen for this study is somewhat limited. If initiated, this work would have to be aimed at achieving a long term payout under a scenario which shows large increases in energy values in general and diesel prices in particular over today's levels.

INTRODUCTION

The purpose of this economic analysis is twofold. The first is to assist in the assessment of the near to intermediate term potential for commercial production of liquid fuels by indirect liquefaction based on relatively cheap remote natural gas. The second is to examine the economics of the slurry phase Fischer-Tropsch (F-T) process and catalyst versus the economics of two existing commercial F-T processes, entrained bed Synthol and fixed bed ARGE. As a sensitivity, the three F-T processes are also compared for the case where the source of the synthesis gas is gasification of midwestern coal. In addition, the economics of these three F-T processes are compared to the economics of liquid fuels production from the Mobil Methanol-to-Gasoline process and also to the economics of LNG.

The size of the commercial facilities to be evaluated was chosen based on the flow of synthesis gas ($CO + H_2$) to the F-T synthesis units. This was set at 96,000 lb mole/hr (875 MM SCFD) and essentially matches the plant size chosen for an earlier F-T economic study performed for the Department of Energy by the Mitre Corporation$^{(1)}$. The actual parameter fixed in the Mitre study was the coal fed to the plant, however. Since the light gases (C_1's and C_2's) formed in the F-T synthesis reaction are also reformed to synthesis gas and sent to F-T reactors, the total amount of synthesis gas fed to F-T reactors is somewhat different for each case. This difference is established by the relative reaction yield structures with respect to C_1 and C_2 make. The increment in liquid production resulting from the decision to reform and recycle the light gases is 15% for the slurry case, 16% for the ARGE case and 23% for the Synthol case. In terms of liquid product output, each F-T case can be considered to be nominally 40,000 BBL/D in size.

SYNTHESIS GAS PRODUCTION

Approximately 875 MM SCFD of synthesis gas ($H_2 + CO$) is required for the three Fischer-Tropsch cases which were examined. The base cases used natural gas as a feedstock. Both steam methane reforming and partial oxidation were evaluated as synthesis gas production methods. As a sensitivity, coal based synthesis gas production cases were also developed. These cases used the Shell pressurized gasifier with midwestern coal as the synthesis gas source. Summary economics for these cases appears in Table 1.

Steam Methane Reforming Versus Partial Oxidation

Initially, it was decided that steam methane reforming of natural gas would be used. The required H_2/CO molar ratios are 2.0 for Synthol, 1.7 for ARGE and 0.92 for the slurry case. PSG Economic Evaluation provided economics for these three cases. The processing scheme involved a reformer which is fed natural gas and some recycle CO_2. This is followed by a Benfield acid gas removal system. At this point in the process the H_2/CO molar ratio is about 3.2. A Prism membrane is used to purge sufficient hydrogen such that the required product compositions are met. It is envisioned that this hydrogen rich purge stream would be used as reformer fuel. The fuel heating value of the purge streams does not exceed the reformer fuel demand in the Synthol and

ARGE cases, but for the slurry case, the purge stream is so large that it cannot all be consumed as reformer fuel. It is uncertain that an alternate use for this hydrogen could be found in a remote location - thus, some of the fuel credits taken in the slurry case may prove to be invalid. Even with the full fuel credits, the slurry synthesis gas cost is significantly higher than the costs for the other two cases since the front end of the plant is also larger.

For these reasons, it was later decided to re-examine the route to synthesis gas production, looking instead at partial oxidation of natural gas with CO_2 recycle and perhaps CO_2 import. Available data indicated that partial oxidation with no CO_2 recycle would yield an H_2/CO ratio of about 1.8. Thus, shifting a small portion of the CO production to H_2 would be necessary to meet the composition requirements of the Synthol case (2.0 H_2/CO) while recycling CO_2 at a rate of about .05 mole CO_2 per mole of hydrocarbon feed satisfied the ARGE composition requirement (1.7 H_2/CO). The slurry case composition (.92 H_2/CO) can be reached by feeding imported CO_2 to the unit such that the molar CO_2 per mole of hydrocarbon fed is about 0.6. Sufficient CO_2 will be available since it is a product of the Fischer-Tropsch synthesis as well as the partial oxidation reaction.

PSG Economic Evaluation also provided synthesis gas economics via partial oxidation of natural gas for the three molar ratios of interest. As is evident from the table below, partial oxidation appears to be the economic choice for facilities of this size for the slurry and ARGE cases, but the steam methane reforming option is more attractive for the Synthol case.

Molar H_2/CO Ratio	1990 Onstream Synthesis Gas Price ($/MSCF)(1) SMR/Prism	POX
0.92	$1.55	$1.44
1.7	$1.24	$1.22
2.0	$1.15	$1.22

(1)Natural Gas Cost = $1.00/MMBTU

The data for partial oxidation shown in Table 1 and summarized above may be slightly confusing since the same synthesis gas price is projected for both the 1.7 and 2.0 molar ratios. A plot of synthesis gas price versus H_2/CO ratio shown in Figure 2 shows that the minimum cost will occur at a H_2/CO ratio of about 1.8. This minimum occurs where neither CO shifting nor CO_2 recycle is required to produce the desired molar ratio.

The choice between synthesis gas production methods is quite sensitive to the natural gas price. Higher natural gas prices would favor partial oxidation since less gas is consumed per MM SCF of synthesis gas produced relative to reforming. At the chosen natural gas price of $1.00/MMBTU, it appears that a 1.7 H_2/CO molar ratio is close to the point of indifference between choosing partial oxidation or steam methane reforming. At higher H_2/CO ratios, reforming would become increasingly attractive since less purging of hydrogen would be required and more CO shift would be necessary with partial oxidation.

Coal Gasification

A coal based facility for synthesis gas production was investigated as a sensitivity to the base cases. The coal cases are not intended to be directly compared to the base cases since a midwestern location is envisioned for the coal based plant versus a remote location for the gas based plants. In fact, the coal price assumed for this sensitivity study ($1.50/MM BTU) is 50% more than the base price ($1.00/MM BTU) assumed for the remote gas.

It is nevertheless interesting to compare the relative synthesis gas prices for the three F-T cases resulting from coal. As the table shows, the least expensive synthesis gas is produced for the slurry case.

Molar H_2/CO Ratio	1990 Onstream Coal Based Synthesis Gas Price ($/MSCF)
0.92	$4.57
1.7	4.85
2.0	4.91

The coal cases are based on Shell pressurized gasifiers which produce synthesis gas at about 400 psi. The advantage for the slurry case stems from the 0.5 H_2/CO ratio which is present at the gasifier outlet. Some shifting of CO is necessary in all cases to attain the appropriate H_2/CO ratio, but the slurry case requires the lowest amount. The differences in relative costs shown above are all due to differences in capital and operating costs occurring in the shift unit and the subsequent acid gas removal. The gasification areas are identical for all three cases.

FISCHER-TROPSCH PROCESS DESCRIPTION

Reaction Area

Synthesis gas with the appropriate H_2/CO ratio is fed to the F-T reactors. A slurry reactor is used for the slurry process while the Synthol process uses an entrained bed and the ARGE process employs a fixed bed.

The slurry reactor essentially acts as a bubble column. The liquid phase is molten wax. Preheated feed gas is sparged into the reactor at the bottom of the column. The coprecipitated Fe/Cu/K catalyst in the liquid phase is suspended by the motion of the gas bubbles. The heat generated is removed by steam generating heat exchangers located in the slurry. The slurry process was designed to operate at 500°F and 315 psia. The catalyst is separated from the molten wax by sequential settling and filtration operations. The recycle gas to feed gas ratio is about 0.6.

The Synthol process uses entrained bed reactors. The feed gases and catalyst mixture enters at the bottom and are entrained overhead following the reaction. Heat exchangers using recirculating oil for cooling are installed in the reaction zone. Following the reaction zone is a catalyst settling hopper. Within this hopper are cyclones which serve to separate the reaction gases from the catalyst. To contain the circulating catalyst, a standpipe is also necessary. This process uses a fused iron catalyst. The catalyst recirculation rate is about 27,500 tons/hr. The synthesis reaction is carried out at temperatures of about 600°F. The reaction pressure is about 315 psia. The recycle gas to feed gas ratio is about 2. Both these conditions suppress the formation of higher liquid hydrocarbons which would prevent stable operation of the reactors.

The ARGE reactor design for this evaluation is similar to that of the existing ARGE reactors at Sasol I except that each reactor in the current design contains two tube bundles in series compared to one bundle for the Sasol I reactors. Each bundle contains about 2800 tubes of 2 inches diameter and 40 feet length. The tubes are filled with promoted iron catalyst. The reaction temperature varies from about 390°F to 480°F and is slowly increased during operation in order to maintain a constant degree of synthesis gas conversion. Reaction pressure is about 355 psia. The heat generated is removed by steam generating heat exchangers in the reaction zone. Gas is recycled such that the recycle to feed gas ratio is about 2.5. Some of the main products of the fixed bed are waxes (C_{19}+ material) which are collected immediately after the reactor.

Key design parameters for each process are given in Table 2, and the reaction yield structures are shown in Table 3. Major utilities for each process are summarized in Table 4. The overall processing scheme is shown in Figure 1.

Reaction Product Processing

The reaction products are sent through a series of separation steps where the gaseous and liquid products are separated. The steam recovered from the highly exothermic reaction can be used to satisfy plant energy requirements, especially those of the air separation units which provide oxygen for the partial oxidation units and the autothermal reformers.

The liquid hydrocarbon products are sent to fractionation and product upgrading units. Gaseous reaction products and unconverted synthesis gas are fed to a CO_2 removal unit. Part of the clean gas is combined with makeup synthesis gas, compressed, and recycled to the F-T reactors. The remaining C_2 and lighter products are sent to an autothermal reformer where further synthesis gas is made. The synthesis gas is then sent to the F-T reactors to increase the overall yield of liquid products. The autothermal reformer was selected instead of partial oxidation due to the presence of olefinic components in the F-T reaction products. Import CO_2 is used in the autothermal reforming step to produce the required H_2/CO ratio. Sufficient CO_2 is formed in the F-T reaction to satisfy any import needs of the partial oxidation and autothermal reformer units. This processing scheme produces no salable products lighter than C_3.

In the slurry and ARGE processes, substantial amounts of wax (C_{19}+) are produced. The waxes are fed to a hydrocracker, producing about 80% diesel and 15% gasoline with the remainder being gases. Since much less waxy material is made by the Synthol process, it was decided to sell the C_{19}+ material made by this process directly without further processing.

Product Upgrading Process Description

The product upgrading scheme used in this evaluation has been modelled after the product upgrading flowsheet used in the previously mentioned Mitre study(1). This processing scheme is shown in Figure 3. The raw product is split into three streams in the F-T product fractionator. The light ends are sent to be autothermally reformed to produce additional synthesis gas.

Another stream, containing the C_3 and C_4 olefins, is catalytically polymerized to dimers and trimers and blended into gasoline. Isobutane and remaining C_3/C_4 olefins from catalytic polymerization are fed to an HF alkylation unit. The product alkylate is also blended with the gasoline fractions. The light polygas from catalytic polymerization is added directly to the gasoline pool while the heavy polygas from the same unit is first hydrogenated before it is blended into gasoline.

The third major stream from the F-T product fractionator, the liquids, is sent to a hydrotreater. The hydrotreated product is then fractionated into three streams. The lightest fraction, composed of C_5 and C_6 components, is fed to a catalyzed isomerization unit where pentanes and hexanes are isomerized to enhance the antiknock quality of the gasoline. The C_7-C_{11} fraction goes to a catalytic reformer, whose product is sent to gasoline blending. The heavy fraction is diesel fuel.

Alcohols are recovered in an alcohol recovery unit. Gases from catalytic and C_5/C_6 isomerization are sent to a hydrogen purification unit where hydrogen is separated from other light ends. The light ends from catalytic reforming, C_5/C_6 isomerization, and the hydrogen purification unit are fed to a light ends recovery unit where a gasoline blend stream is separated from LPG.

The upgraded products for the three cases are given in Table 5. Total products are essentially the same, both on a volume basis and on a weight basis. As expected, the Synthol case has a large advantage in gasoline production while the slurry and ARGE cases provide more diesel fuel.

ECONOMIC ANALYSIS AND DISCUSSION OF F-T SYNTHESIS AND PRODUCT UPGRADING

Process Simulation Unit Sizing and Cost Estimation

Process simulations of the F-T synthesis areas were developed for the three cases(11). From these simulations it was possible to estimate preliminary major equipment sizes for the F-T synthesis areas. Capital costs were then developed by roughly estimating prices for the major equipment and factoring this major equipment estimate to obtain a total installed cost.

Capital costs for areas outside of the synthesis loop were prepared by scaling available data for similar units based on total flows to the unit. Process units handled in this manner include wax hydrocracking, autothermal reforming and product upgrading. Utilities and offsites were estimated as a percentage of onsites capital cost although it was decided not to attempt to differentiate among the three cases in this area.

Discussion of Capital Costs

The capital cost estimates for the three cases are given in Table 6. Capital costs are shown for a plant coming onstream in 1990 and exclude the cost of the synthesis gas plant. An investment of about $1.3 billion is estimated for the slurry plant with the ARGE and Synthol cases requiring about 50% greater capital. Each plant will produce about 40,000 BBL/D liquid fuels.

The primary area of projected savings for the slurry case is in the reaction area itself. Synthol is estimated to be more expensive mainly due to the tall reactors, large catalyst settling hopper and standpipe, and the extensive structure needed to support them. Higher investment for the entrained bed relative to slurry is also supported by a UOP study performed for DOE(3). Another factor in the higher investment for Synthol is the higher recycle to fresh feed ratio relative to the slurry case. The higher recycle to fresh feed ratio is also a factor in the estimated higher cost of the ARGE process relative to the slurry case. The main factor in the higher ARGE cost, though, is the requirement for twice as many reactors as the slurry case and four times as many as the Synthol case. Equipment parameters for the F-T synthesis section are given in Table A.1 of the Appendix for each case. This table illustrates the differences in number and size of reactors, heat exchanger surface areas, and recycle compressor sizes which are major factors in establishing the projected capital cost differences among the F-T synthesis areas.

The Synthol case product upgrading and autothermal reforming sections cost more than these sections for the other processes since it produces relatively greater amounts of light ends. Conversely, the slurry and ARGE cases require wax hydrocracking because they make relatively large amounts of $C_{19}+$.

A specific remote location for the F-T comparison has not been selected. It has been suggested, though, that if the location were Alaska, considerable investment and operating costs could be saved by shipping the raw product in the Alaska pipeline and upgrading it at an existing refinery. Although no attempt has been made to evaluate this option, it is assumed that much of the savings would be eroded by having to accept lower prices for the raw products.

Product Prices and Projected Annual Revenues

Product prices and projected annual revenues for the year 1990 are given in Table 7. The price projections are based on information provided by APCI Corporate Energy. The total projected revenues are nearly the same for all cases with Synthol lagging the other two by about 3%. Because they are so close, both in quantity and in projected value, the relative product slates do not significantly affect the relative economic attractiveness of the cases. A change in output for one of the cases by a few percentage points would not significantly affect its standing among the three cases. Note also that according to these price projections, gasoline and diesel are essentially valued the same on a volume basis. This is an advantage for the slurry and ARGE cases and is detrimental to Synthol.

Production Costs and Required Annual Revenues

Table 8 illustrates the production costs and required annual revenues for the F-T cases. The required revenues are set such that operating costs are covered plus a 7.5% after tax return on investment is realized. The required revenues are then compared to the projected revenues. Although the facility is envisioned as one large plant including the synthesis gas production, for illustrative purposes the synthesis gas production costs are treated separately. Synthesis gas is then purchased by the F-T plants at the appropriate transfer price shown in Table 1.

Table 8 shows that the largest operating cost for the F-T plants is the purchase of synthesis gas. Other operating costs include a balance of plant power requirements and steam credits, fuel gas, catalysts and chemicals, operating labor, and maintenance material and labor. Details of these costs are presented in Table A.2 of the Appendix.

The net result of these calculations is that each of the cases shows a shortfall of revenues on a 1990 onstream basis. As Table 8 shows, on a dollars per barrel of product basis, the shortfall is $25/BBL for the slurry case, $36/BBL for Synthol and $34/BBL for ARGE. The advantage to the slurry case stems from the projected capital savings in the F-T synthesis area which significantly outweighs the extra cost of synthesis gas for this case.

The results essentially say that for natural gas valued at $1.00/MMBTU, products whose projected 1990 value is around $35/BBL would need to sell at about $60/BBL in the slurry case and about $70/BBL in the Synthol and ARGE cases. It should also be noted that assuming a value of zero for the natural gas would not nearly eliminate the revenue shortfall. Natural gas cost comprises about $9.00/BBL of the shortfall for the slurry and ARGE cases and about $13.00 for the Synthol case.

As would be expected from the results presented above, the decision to further process the light gases formed in the F-T reaction to produce an all liquid product slate is not economic in the sense that the additional capital and operating costs required exceed the additional revenues achieved by upgrading the low value gas. However, it should be pointed out that the per barrel revenue shortfall for the incremental liquids production for each case is less than the average per barrel shortfall for the total liquid production in each case. In other words, given the low value assigned to gas, an all-liquid producing plant is more attractive economically than one producing a mixture of SNG and liquids, even though neither case is attractive on an absolute basis. In any event, the decision to produce an all-liquid product slate did not affect the ranking of the three cases economically.

In the recent past, many proposed synthetic fuel projects used energy price projections which significantly outpaced general inflation to justify their existence. For the sake of discussion, this type of analysis can be applied here and the slurry case was chosen as an example. If the 1990 price projections are accepted as valid, and general inflation remains at 5% per annum, the product values must increase by 12% per annum over a plant operating life of 16 years in order for the 7.5% after tax return on investment to be realized. This differential in product value growth and cost growth may not seem large, but at these escalation rates over the life of the project, revenues would increase by a factor of nearly 5.5 while costs would slightly more than double. In the above analysis the remote natural gas price was also assumed to increase at only 5% per annum which could be an optimistic assumption.

Figure 4 demonstrates that such rapid growth in energy values has occurred over the last fifteen years as it tracks the pricing of crude oil, gasoline and diesel. Also included for comparison purposes in Figure 4 is the Producer Price Index for all commodities. This index has grown by about a factor of three while crude oil, gasoline and diesel fuel have increased eight or nine fold. The key question facing synthetic fuels development work is if and when we will see a renewal of this explosive growth.

Sensitivity to Coal Based Synthesis Gas Production*

Table 8 also illustrates the sensitivity of switching liquid fuel production by F-T synthesis to a coal based plant with coal valued at $1.50/MMBTU. The revenue shortfall becomes much worse for all cases, ranging from about $90/BBL in the slurry case to about $120/BBL for Synthol. The slurry case is actually helped relative to other cases by switching to coal since less shifting of CO is necessary to reach the appropriate H_2/CO ratio in the feed gas.

Excess Steam Credits

The F-T reaction is so highly exothermic that more than enough steam is available to make the plants (including synthesis gas production) self sufficient in power. Based on the assumption that no user of excess steam or power from steam would be near the remote location, the current cases were credited with only enough steam to exactly balance each plant's steam and power requirements. In reality, by superheating more of the 425 psia saturated steam which is produced in the slurry and Synthol cases and condensing the 250 psia saturated steam instead of letting it down to only 30 psia in the ARGE case, the following additional power credits could be realized.

Slurry	161,000 kw
Synthol	132,000 kw
ARGE	104,000 kw

The gross value of these credits at $.03/kwh ranges from $25 MM - $40 MM annually. If these credits were included in the economics, these gross values wouldn't be fully realized since additional capital and fuel gas would be needed to produce this power or power equivalent.

*A revised Product Cost Summary is located in Appendix 2.

Comparison of F-T Processes with LNG Production

Obviously, another use for the remote gas is its conversion to LNG so that a comparison of this option to indirect liquefaction should prove to be useful. A recent SRI International study (9) indicated that natural gas valued at \$1.00/MMBTU could be liquefied, delivered, and regassified for as low as \$3.00/MMBTU in 1983 dollars, depending on the shipping distance. Escalation to a 1990 basis, with natural gas remaining at \$1.00/MMBTU, would put the LNG costs in the \$3.50-4.00/MMBTU range. These LNG values were corroborated by other information obtained from sources within APCI. In contrast, the required 1990 selling price of liquid product from the slurry case is more than \$11.00/MMBTU, excluding delivery, so the indirect liquefaction option is not competitive with LNG on an absolute basis.

The preceding discussion does not take into account the form value of the fuel. Transportation fuels should be more highly valued than natural gas to be used as boiler fuel. The point, though, is that this enhanced valued projected for 1990 is not sufficient to make F-T attractive.

Part of what is being illustrated here is the inherent inefficiency of indirect liquefaction. The inefficiency of synthesis gas manufacture is coupled with the inefficiency of F-T synthesis such that only about one half, or even less, of the original energy is recovered as liquid products, depending on the process scheme used.

ECONOMIC COMPARISON OF THE MOBIL MTG PROCESS AND THE FISCHER-TROPSCH PROCESSES

Process Description

Figure 5 shows the overall Mobil MTG natural gas to gasoline process. Natural gas, with steam in a volume ratio of 1:3, is steam reformed to produce synthesis gas at the rate of about 875 MMSCFD. The H_2/CO ratio of the synthesis gas is 4.92:1 and requires no CO_2 recycle. The syn gas is cooled and compressed to 470°F and 1500 psi and enters the methanol reactor where it is converted to methanol using a copper based catalyst.

The crude methanol-water solution is vaporized and enters the first stage of the MTG reactor at 600°F and 315 psi. This reactor contains a fixed bed of activated alumina catalyst. The methanol vapor is partially dehydrated to form dimethyl ether.

The methanol-dimethyl ether stream at 650°F enters the second stage reactor which contains a fixed bed of ZSM-5 catalyst. Here, the methanol and dimethyl ether are completely dehydrated to light olefins and then undergo further condensation and rearrangement to form primarily gasoline-cut hydrocarbons. The conversion of methanol is highly exothermic, approximately 740 BTU/lb methanol. This heat is recovered by producing steam or by heat exchange with cooler streams.

The hydrocarbons enter a separation train. In the first stripping column, C_1-C_2 hydrocarbons are removed. The C_1-C_2 are recycled to the steam reformer. Next, the C_3-C_4 hydrocarbons are separated from the C_5+ cut and sent to an alkylation unit. The raw C_5+ gasoline stream is blended with alkylate and butanes to obtain the desired composition.

Economics

The economics shown in Table 9 for the MOBIL MTG process were based on those presented in the SRI PEP Review No. 81-2-3. "Gasoline from Natural Gas Via Methanol by the MOBIL Fixed Bed MTG Process." Due to the limited data presented in the PEP review, it was impossible to accurately separate the syngas cost from the gasoline synthesis cost. For that reason, economics for the three Fischer-Tropsch processes and Mobil MTG process are presented in Table 9 starting with natural gas.

The basis for the economics in Table 9 is a synthesis gas production rate of 96,160 mole/hr (~875 mmscfd). In Table 5, we see that the three F-T processes generate approximately the same output, 40,000 BBL/day. The MTG process however, produces significantly lower output, 29,000 BBL/day from the same quantity of syngas. (Refer to Table 10). This results from the significantly higher H_2/CO ratio being used for MTG and from different process efficiencies.

The synthesis gas cost for the Mobil MTG process H_2/CO ratio of 4.92 is significantly cheaper than for the ratios required for F-T synthesis and is on the order of $0.74/mscf. This is due to the absence of a Prism unit and CO_2 recycle and the lower natural gas requirements. However, the MTG process would require approximately 35% more synthesis gas to make the same output as the Fischer-Tropsch processes.

The Mobil MTG process shows a significant shortfall of revenues between what is required and what is projected. Operation of this process under current assumptions would not be economically viable. Since the Mobil MTG process makes less product, comparison with the Fischer-Tropsch processes shows the MTG process to also have a lower absolute shortfall. However, when revenue shortfalls are compared on a per barrel of product basis or equal output basis, the slurry Fischer-Tropsch process and the MTG process are approximately the same, $25/BBL and $28/BBL respectively, and are significantly lower than those of either ARGE or Synthol. Slurry Fischer-Tropsch and Mobil MTG do not currently appear to be economically viable processes, but they would be preferred over Synthol and ARGE. Although we feel relatively confident in the results of this study, we cannot be completely assured of the consistency between estimating methods for the Fischer-Tropsch processes and the Mobil processes, due to the lack of details provided in the PEP report.

CONCLUSIONS AND RECOMMENDATIONS

(1) Based on 1990 fuel price projections, none of the liquefaction technologies studied is economically attractive.

All of the Fischer-Tropsch cases and the Mobil MTG case exhibit substantial revenue shortfalls when required revenues for profitability and projected 1990 revenues are compared. Process improvements would not provide a means to achieve profitability. What would be needed is a return to the situation where energy price increases significantly outpace general inflation or a quantum jump in prices. For example, a 12% per annum increase in liquid product values beginning in 1990 coupled with a 5% per annum escalation in costs, including natural gas, would yield a 7.5% NPAT in the slurry case. No natural gas price would make any of the cases profitable at the projected 1990 fuel prices.

(2) The slurry case appears attractive economically relative to the ARGE and Synthol cases.

The advantage for the slurry case is based almost entirely on the estimate of significantly lower capital costs for this synthesis loop.

(3) A corollary to the previous conclusion is that the relative product slates of the three cases are so close in both quantity and projected value that they have little impact on the relative economics. Other things being equal, the liquid yield from the slurry case would have to be decreased by about 25% to eliminate its current advantage.

(4) Introducing coal gasification as a synthesis gas source for the Fischer-Tropsch processes provides an economic advantage for the slurry case in a relative sense, but on an absolute basis all of the cases become much worse economically.

The relative advantage to slurry for a coal based facility stems from decreased CO shift requirements needed to reach the proper H_2/CO feed ratio. However, the high capital investment necessary for the coal gasification facility pushes the cost of liquid fuels from any of the Fischer-Tropsch processes to prohibitive levels.

(5) Mobil MTG appears to be roughly equivalent in economic attractiveness to the slurry F-T case.

This conclusion is based on the similar revenue shortfall per barrel of product for MTG relative to slurry. Although the MTG information was acquired from a respected source (SRI International), some reservation concerning this conclusion must be expressed since it was difficult to ensure that the economics used were developed on a totally consistent basis. MTG would likely provide stiff competition for slurry F-T anyhow since it is a more proven process.

(6) LNG appears to be a considerably more attractive market for remote natural gas than producing liquid fuels would be.

> This conclusion is based on an SRI International study which shows that natural gas at $1.00/MMBTU can be liquefied, delivered and regassified for as low as $3.00/MMBTU in 1983 dollars, depending on the shipping distance. Escalation to a 1990 basis, with natural gas remaining at $1.00/MMBTU, would put the LNG costs in the $3.50-4.00/MMBTU range. In contrast, the cost of producing liquid fuels by slurry F-T in 1990 is in excess of $11.00/MMBTU, excluding delivery. Transportation fuels should have a higher form value than gas, but this difference would not be enough to significantly close the above gap, which is solely based on production costs.

Further study could be undertaken to confirm the economic advantage of the slurry versus other established F-T processes. Such work should probably initially involve confirmation of the currently projected capital cost advantage. Also, even though the relative product yield structures were not a critical factor in determining the relative economics, the projected yields for the slurry process should be confirmed by further experimental work since the current data base at the conditions chosen for this study is somewhat limited. If initiated, this work would have to be aimed toward achieving a long term payout. In addition, it should proceed only under some scenario which shows that the demand for diesel fuels is likely to increase preferentially to other fuels. Otherwise, other process options may possess an advantage over F-T even in a market where liquid fuels are more highly valued.

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11. Memorandum - C. Ebihara/C. L. Yeh to A. R. Kuhns, 5 November 1985.
12. Memorandum - N. Cilen to A. R. Kuhns, 6 November 1985.

TABLE 1

ECONOMICS OF SYNTHESIS GAS PRODUCTION
875 MM SCFD
(1990 ONSTREAM)

Feedstock	Steam Methane Reforming Remote Natural Gas			Partial Oxidation Remote Natural Gas			Gasification Coal		
H_2/CO Molar Ratio	0.92	1.7	2.0	0.92	1.7	2.0	0.92	1.7	2.0
Feedstock Cost ($/MMBTU)	1.00	1.00	1.00	1.00	1.00	1.00	1.50	1.50	1.50
Capital Investment (MM$)	656	503	453	871	772	781	3,650	3,910	3,960
Annual Costs (MM$)									
Feedstock	195	168	161	121	115	114	164	164	164
Other Operating	81	59	53	68	36	36	231	251	256
Capital Related	171	131	118	227	201	203	920	980	995
TOTAL	447	358	332	416	352	353	1,315	1,395	1,415
SYN GAS COST ($/MSCF)	1.55	1.24	1.15	1.44	1.22	1.22	4.57	4.05	4.91

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TABLE 2

COMPARISON OF F-T PROCESS DESIGN PARAMETERS

	Slurry Reactor	Entrained Fluidized Bed (Synthol)	Fixed Bed (ARGE)
Operating Pressure, psia	315	315	355
Operating Temperature, °F	500	600	450
Feed H_2/CO Ratio	0.9	2.0	1.7
Recycle/Feed Gas Ratio	0.6	2.0	2.5
Catalyst Loading, SCF Feed/ft. cat. hr	2000*	700	350
Feed Conversion, %	95	95	95
Reference	(12)	(4)	(4)

*25 wt. % catalyst concentration

TABLE 3

SYNTHESIS GAS COMPOSITION AND PRODUCT YIELD STRUCTURES
(Wt.%)

FEED	SLURRY	SYNTHOL	ARGE
H_2	6.2	12.9	10.9
CO	93.8	87.1	89.1
	100.0	100.0	100.0
PRODUCTS			
CO_2	54.9	3.6	15.0
H_2O	13.4	52.0	43.3
CH_4	1.3	4.4	2.5
C_2	1.4	3.6	1.2
C_3	1.7	6.2	2.2
C_4	1.4	4.9	1.7
C_{5-11}	6.6	17.7	6.8
C_{12-18}	4.2	3.1	4.9
C_{19+}	13.9	1.8	19.7
Oxygenates	1.2	2.7	2.7
	100.0	100.0	100.0

TABLE 4
UTILITY SUMMARY

	Power (kw)	Process Steam Import (150 psia or less) (M LB/HR)	Export Steam (M LB/HR)	Fuel Gas (MM BTU/HR)
Slurry Case				
Partial Oxidation	256,300	-	1,020(2)	-
F-T Synthesis	31,200	-	2,890(3)	173
Product Upgrading	10,350	61	-	550
Autothermal Reforming	18,450	101	130(4)	135
Wax Hydrocracking	1,570	7	-	42
Utilities and Offsites	10,000	-	-	100
TOTAL	327,870	169	4,840	1,000
Synthol Case				
Steam Methane Reforming(1)	110,700	-	435(2)	-
F-T Synthesis	66,300	-	3,146(5)	260
Product Upgrading	11,600	51	-	760
Autothermal Reforming	23,700	153	197(4)	204
Utilities and Offsites	10,000		-	100
TOTAL	222,300	204	3,778	1,324
ARGE Case				
Partial Oxidation	192,700	-	2,360(2)	-
F-T Synthesis	46,400	-	3,020(6)	-
Product Upgrading	9,770	58	-	510
Autothermal Reforming	16,970	107	130(4)	143
Wax Hydrocracking	1,660	7	-	44
Utilities and Offsites	10,000	-	-	100
TOTAL	277,500	172	5,510	797

(1) Reformer fuel gas is included with process natural gas
(2) 1500 psia, 850°F
(3) 425 psia (SAT)
(4) 600 psia (SAT)
(5) 2,786,000 lb/hr 425 psia (SAT) and 360,000 lb/hr 150 psia (SAT)
(6) 250 psia (SAT)

TABLE 5

COMPARISON OF FISCHER-TROPSCH PRODUCT SLATES
875 MM SCFD SYNTHESIS GAS

	SLURRY		SYNTHOL		ARGE	
	lb/hr	BBL/D	lb/hr	BBL/D	lb/hr	BBL/D
C_3 LPG	7,080	960	16,420	2,220	7,740	1,050
C_4	1,510	180	2,570	300	1,180	140
Gasoline	170,100	16,400	319,700	30,800	149,250	14,390
Diesel	249,400	22,670	37,780	3,430	254,630	23,150
Heavy Oil	-	-	21,590	1,700	-	-
Alcohols	17,000	-	32,950	-	32,580	-
TOTAL	445,090	40,210	431,010	38,450	445,380	38,730

TABLE 6

CAPITAL INVESTMENT SUMMARY FOR LIQUID FUELS SYNTHESIS
(1990 START-UP)
MM$

	SLURRY	SYNTHOL	ARGE
Fischer-Tropsch Synthesis	$ 255	$ 615	$550
Product Upgrading	150	180	150
Autothermal Reforming	110	135	115
Wax Hydrocracking	30	-	35
Utilities and Offsites	165	165	165
Contingency	180	275	255
Subtotal (1985 $)	890	1370	1270
Work In Progress Escalation	130	200	185
Interest During Construction	250	385	355
TOTAL	$1,270	$1,955	$1,810

(1) Excludes investment for synthesis gas production. Purchase of synthesis gas is treated as an operating cost in Table 8.

TABLE 7

PRODUCT PRICE PROJECTIONS
(1990 DOLLARS)

C_3 LPG	\$17.40/BBL
C_4	\$21.00/BBL
Gasoline	\$36.60/BBL
Diesel	\$36.12/BBL
Fuel Oil	\$25.00/BBL
Alcohols	\$ .15/lb

ANNUAL REVENUE PROJECTIONS
(1990 - MILLIONS)
330 DAY/YR

	SLURRY	SYNTHOL	ARGE
C_3	\$ 5.5	\$ 12.7	\$ 6.0
C_4	1.2	2.1	1.0
Gasoline	198.1	372.0	173.8
Diesel	270.2	40.9	275.9
Fuel Oil	-	14.0	-
Alcohols	20.2	39.1	38.7
TOTAL	\$495.2	\$480.8	\$495.4

TABLE 8

ANNUAL COSTS FOR FISCHER-TROPSCH SYNTHESIS *
NOMINAL PLANT CAPACITY 40,000 BBL/D
(1990 ONSTREAM)
MM$

	SLURRY	SYNTHOL	ARGE
Synthesis Gas(1)	416	332	352
Other Operating	106	139	141
Capital Related(2)	330	508	471
Total 1990 Required Revenue	852	979	964
Projected 1990 Revenues	495	481	495
Revenue Shortfall	357	498	469
Revenue Shortfall ($/BBL Product)	25	36	34

Sensitivity to Coal Based
on Synthesis Gas Production

	SLURRY	SYNTHOL	ARGE
Total 1990 Required Revenues	1,747	2,057	2,012
Revenue Shortfall	1,252	1,576	1,517
Revenue Shortfall ($/BBL Product)	93	121	116

NATURAL GAS PRICE	$1.00/MM BTU
COAL PRICE	$1.50/MM BTU

*A revised Product Cost Summary is located in Appendix 2.

(1) Synthesis gas from partial oxidation for the slurry and ARGE cases and steam methane reforming for Synthol.

(2) Includes an allowance for 7.5% net profit after tax return on investment.

TABLE 9

ANNUAL COSTS FOR LIQUID FUEL PRODUCTS[A]
STARTING WITH NATURAL GAS
(1990 Onstream)

	Fischer-Tropsch			
(MM $)	Slurry	Synthol	ARGE	Mobil MTG
Natural Gas-Feedstock + Fuel	121	161	115	91
Other Operating	174	192	177	80
Capital Related[B]	557	626	672	417[C]
Total 1990 Required Revenues	852	979	964	588
Projected 1990 Revenues	495	481	495	323
Revenue Shortfall	357	498	469	265
Revenue Shortfall/BBL product	$25	$36	$34	$28

[A] All cases produce 96,160 mole/hr of syngas as an intermediate.

[B] Capital investment includes that for producing syn gas (steam reforming for Synthol and Mobil; partial oxidation for ARGE and Slurry) and that for producing liquid fuels (F-T or MTG).

[C] Total investment for Mobil MTG case is $1572 MM and is based on SRI PEP Review No. 81-2-3.

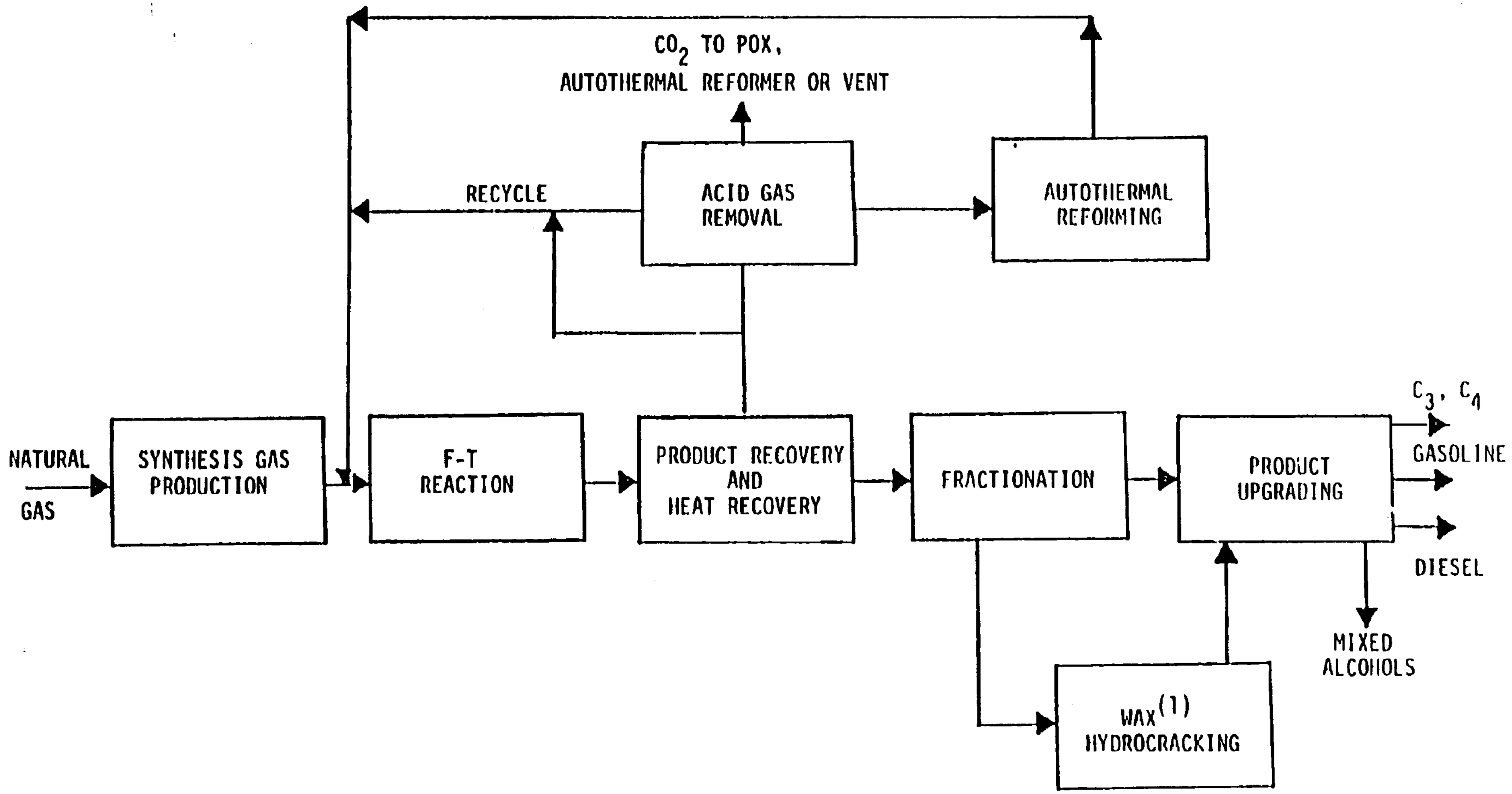
TABLE 10

MOBIL MTG PRODUCT SLATE
875 MM SCFD SYNTHESIS GAS

	lb/hr	BBL/D
Propane LPG	14,072	1,904
Butanes	22,960	2,697
Gasoline 96 RON	252,410	24,329
TOTAL	289,442	28,930

FIGURE 1

OVERALL PROCESS FLOWSCHEME



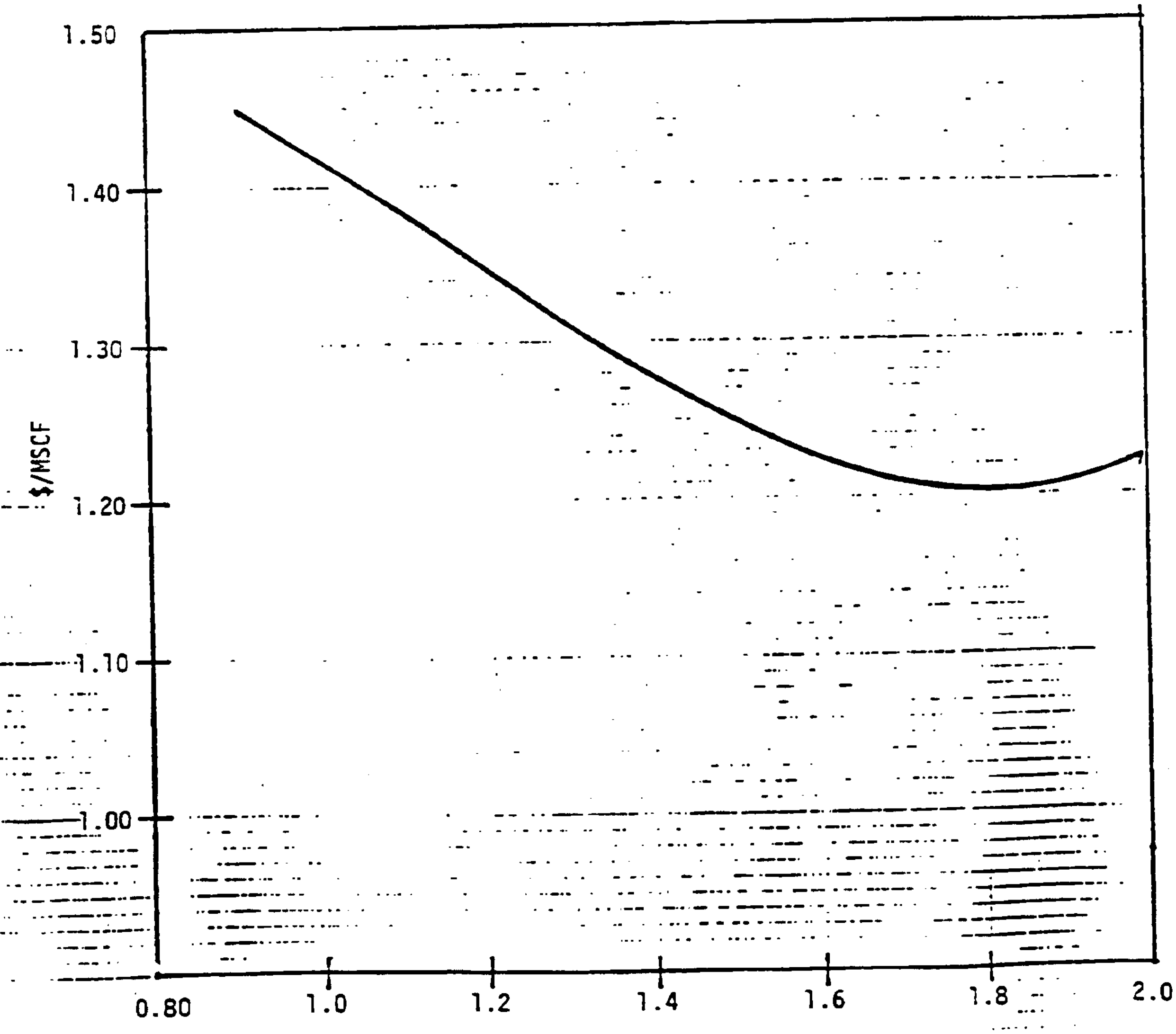
(1) NOT PRESENT IN SYNTHOL CASE

FIGURE 2

PARTIAL OXIDATION OF NATURAL GAS

SYNTHESIS GAS COST VERSUS H_2/CO RATIO

1990 ONSTREAM



NATURAL GAS PRICE $1.00/MMBTU

FIGURE 3

FISCHER TROPSCH PRODUCT UPGRADING FLOWSCHEME

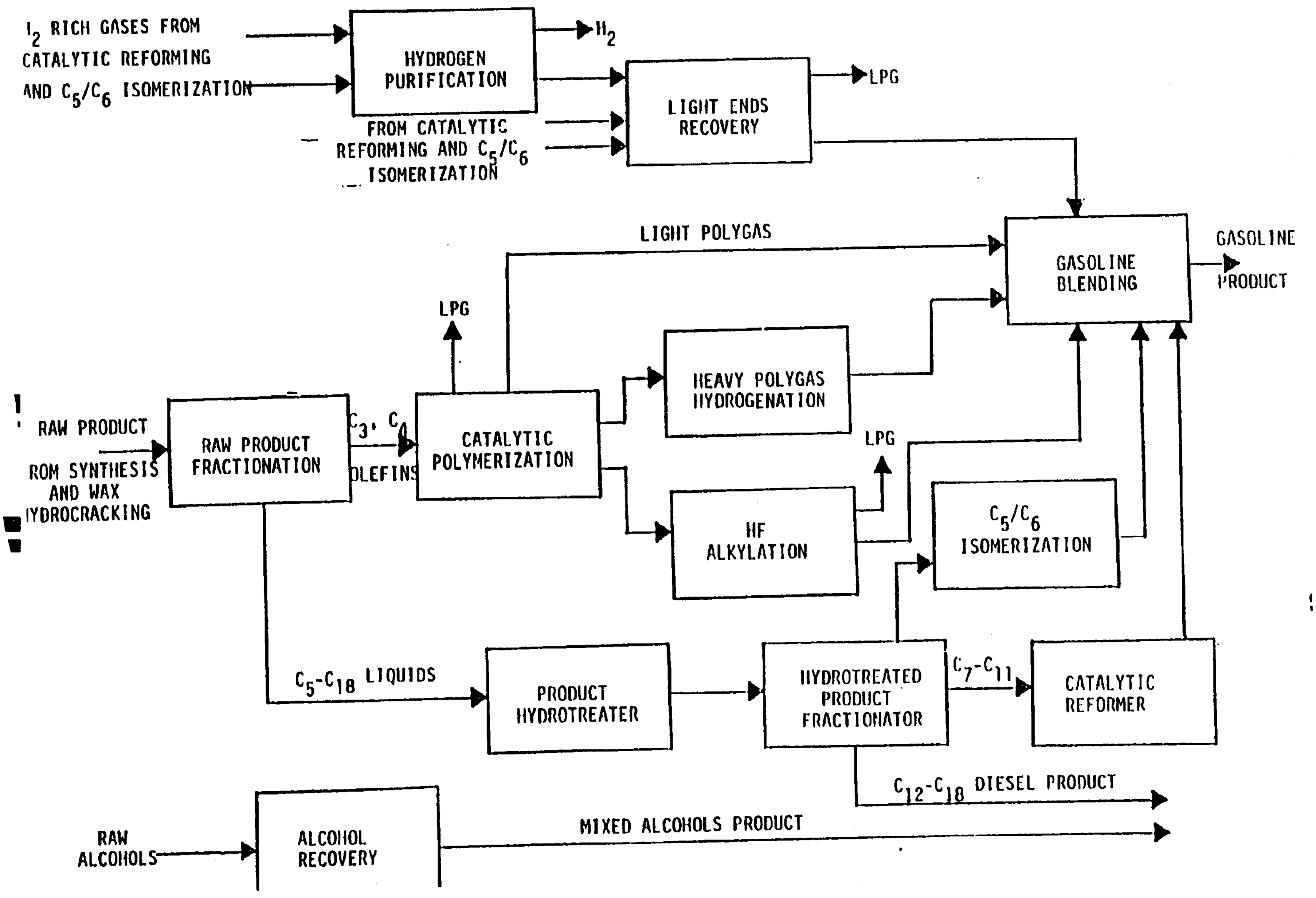
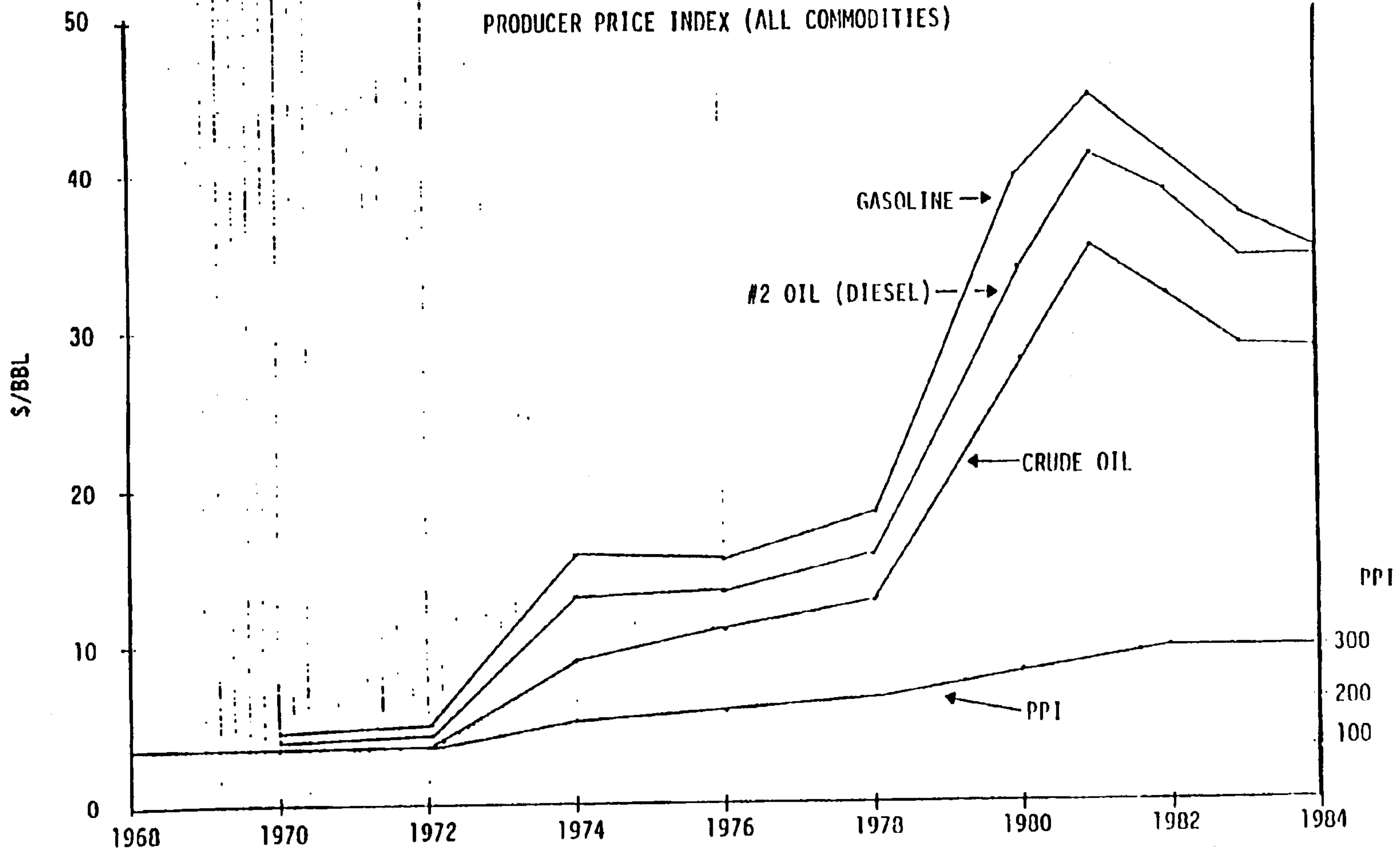


FIGURE 4

RECENT PRICE HISTORY OF CRUDE OIL AND LIQUID FUELS
AND
PRODUCER PRICE INDEX (ALL COMMODITIES)



SOURCE: APCI CORPORATE ENERGY

FIGURE 5

MOBILE MTG PROCESS

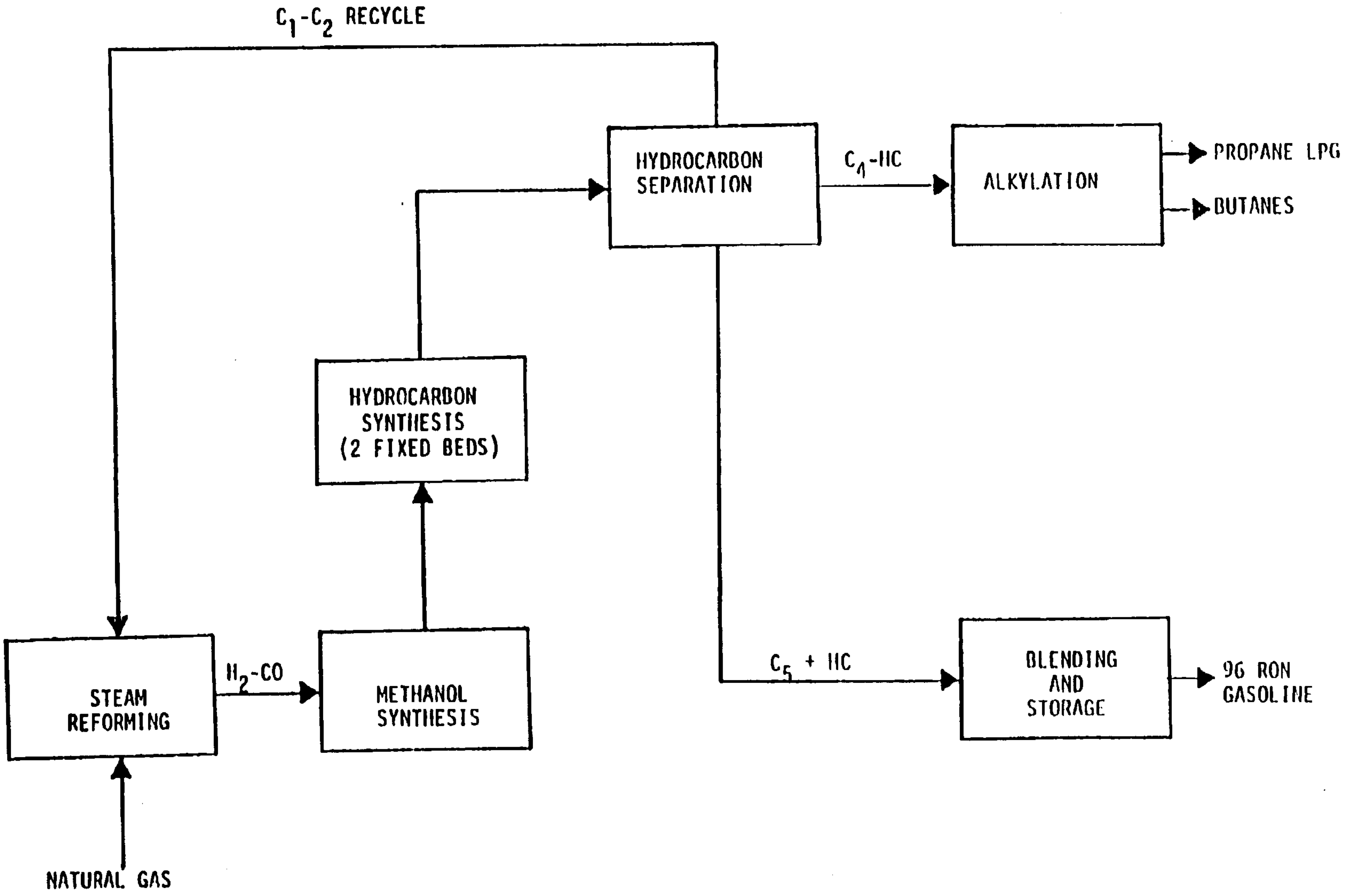


TABLE A.1

Equipment Parameters in F-T Synthesis Areas
Basis: 875 MM SCFD Synthesis Gas to Reactors

	Slurry	Synthol	ARGE
Reactors			
Type	Slurry	Entrained Bed	Fixed Bed
Number	12	6(1)	24
Size	12'4" I.D. x 72' T x T	10' ID x 170' T x T	11'6" ID x 90' T X T
Internal Heat Exchange Surface Per Reactor (Ft^2)	74,700	108,000	117,000
Recycle Compressor BHP	21,600	65,340	47,320
External Heat Exchanger Surface Area (Ft^2)	187,000	408,000	549,000

(1) Each reactor also requires a 29'3" ID x 30' T x T catalyst settling hopper and a 10' ID x 125' standpipe.

TABLE A.2

ANNUAL OPERATING COSTS FOR SYNTHESIS GAS PRODUCTION AND F-T SYNTHESIS
1990 DOLLARS MILLIONS
330 DAYS/YR OPERATION

SYNTHESIS GAS PRODUCTION

	Partial Oxidation		Steam Reforming
H_2/CO Ratio	0.92	1.7	2.0
Natural Gas @ $1.00/MMBTU	$120.8	$114.9	$160.9
Power @ 30 mills/kwh	60.9	45.8	26.3
Steam @ $2.60/M lb	(37.5)	(48.6)	(9.0)
Catalysts, Chems & Lubes	5.1	4.2	4.0
Operating Labor	1.8	1.8	4.2
Maintenance	35.2	30.5	18.1
Miscellaneous	3.2	3.2	9.2
	$189.5	$151.8	$213.7

F-T SYNTHESIS

	Slurry	Synthol	ARGE
Synthesis Gas	$416.0	$332.0	$352.0
Power @ 30 mills/kwh	17.0	26.5	20.1
Steam @ $2.60/MLB Superheated @ $1.15/MLB Saturated(1)	(33.7)	(39.8)	(17.5)(2)
Fuel Gas @ $1.00/MMBTU	7.9	10.5	6.3
Catalyst, Chems and Lubes	53.1	57.9	53.4
Operating Labor	20.7	20.7	20.7
Maintenance	40.8	62.8	58.2
	$521.8	$470.6	$493.2

(1) The value of saturated steam was established by the amount of power which could be extracted from it relative to the superheated steam.

(2) Value is reduced since 1,100,000 lb/hr saturated steam (36% of total available) is condensed without power recovery.

Appendix 2

REVISED PRODUCT COST SUMMARY

Air Products and Chemicals, Inc.
7201 Hamilton Boulevard
Allentown, PA 18195-1501

Telephone (215) 481-4911
Telex 847416
Fax: (215) 481-5900

AIR PRODUCTS

1 July 1991

Mr. George Cinquegrane
U.S. Department of Energy
Pittsburgh Energy Technology Center
P.O. Box 10940, MS 920-L
Pittsburgh, PA 15236

Subject: Contract No. DE-AC22-85PC80011
"Development and Process Evaluation of Improved Fischer-Tropsch Slurry Catalyst"

Dear Mr. Cinquegrane:

In response to DOE's comments emanating from the draft final report for the subject contract, Air Products has reviewed the cost to produce synthesis gas from coal and the impact this has on Fischer-Tropsch process economics. The enclosed "Product Cost Summary" shows the crude oil break-even price for F-T product using recently updated coal derived synthesis gas costs. These syngas costs are approximately 1/3 of those that were used in the original draft final report, and are based on extensive data developed under the Liquid Phase Methanol Technology Program (1987-1990).

For the record, the original process and economic evaluations performed under this contract were based on a CO-rich feedstock derived from natural gas. The process had included a water gas shift reactor to adjust the syngas to the CO-rich composition typical of that derived from coal. This approach was used since earlier studies at Air Products (ca. 1985) had indicated that syngas derived directly from coal would cost up to three times that derived from natural gas. The recent studies completed under the LPMEOH program show this assumption to be incorrect, with natural gas and coal-based syngas costs being more equal ($4 - 5.50 MM/BTU).

The Projected Revenues shown in the summary were determined based on a January 1990 crude price of $18/BBL and include anticipated (at that time) price increases. As can be seen, Revenue Shortfalls are now similar. The break-even selling price for F-T product compared to crude oil is also shown. If crude oil prices should rise to $35-40/BBL, Fischer-Tropsch processing looks promising, especially since natural gas costs were assumed to be $1.00 MM BTU and in most instances natural gas prices are higher.

If you have any questions concerning this information, please call me at (215) 481-3352.

Sincerely yours,

Howard P. Withers

Howard P. Withers
Program Manager

2605J-IA

PRODUCT COST SUMMARY - $MM

Basis: Remote Natural Gas @ $1.00 MM Btu
40,000 BBL/day Liquid Fuel Production
Products Selling Prices Based on $18/BBL Crude Oil

	Natural Gas	COAL @$4.30 MM BTU	@$5.50 MM BTU
Costs			
Syngas	$370.50	$392.70	$503.10
Other	48.50	48.50	48.50
Capital	202.30	202.30	202.30
Total Annual Cost	621.30	643.50	753.90
Projected Shortfall			
$MM/Yr	299.20	324.40	431.80
$/BBL	22.50	24.20	32.50
Crude Oil Price Required to Break Even	$35/BBL	$36/BBL	$42/BBL

2605J-IA