# ECONOMIC ANALYSIS
# LPMEOH™ PROCESS
# AS AN ADD-ON TO INTEGRATED GASIFICATION COMBINED CYCLE (IGCC) FOR COPRODUCTION

**Prepared by**

**Air Products and Chemicals, Inc.**
**Allentown, Pennsylvania**

**for the**
**Air Products Liquid Phase Conversion Company, L.P.**

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# DISCLAIMER

## Abstract

The Liquid Phase Methanol (LPMEOH™) Demonstration Project at Kingsport, Tennessee, is a \$213.7 million cooperative agreement between the U.S. Department of Energy (DOE) and Air Products Liquid Phase Conversion Company, L.P. (the Partnership) to produce methanol from coal-derived synthesis gas (syngas). Air Products and Chemicals, Inc. (Air Products) and Eastman Chemical Company (Eastman) formed the Partnership to execute the Demonstration Project. The LPMEOH™ Process Demonstration Unit was built at a site located at the Eastman chemicals-from-coal complex in Kingsport.

This Topical Report compares the cost of methanol as produced from the LPMEOH™ Process and from a conventional gas phase process as applied to a generic 500 short tons-per-day methanol plant as part of an integrated gasification combined cycle (IGCC) coproduction facility. The cost of methanol is calculated as the sum of three terms: the methanol conversion cost (which includes the fixed and operating costs for the methanol unit), the distillation cost, and the syngas cost from the IGCC facility. A proprietary cost estimation screening program was used to calculate the methanol conversion cost and the distillation cost from the LPMEOH™ Process and the gas phase process for various syngas supply pressures and on-stream factors. The methanol conversion cost from the LPMEOH™ Process is \$0.02 to \$0.08 per gallon lower than from the gas phase methanol process.

A major component of the methanol conversion cost in an IGCC complex is the cost to distill the crude methanol product in order to meet the final specification. Based upon the current applications of methanol, there are three grades of methanol product (Chemical Grade AA, Fuel Grade, and MTBE (methyl tertiary-butyl ether) Grade) which could be used in downstream chemical or power applications. The Fuel Grade and MTBE Grade products have a water specification of 1 wt%, while the Chemical Grade AA methanol has a maximum water content of 0.1 wt%.

The LPMEOH™ Process, which can directly handle coal-derived syngas which is rich in carbon monoxide (CO), produces a crude methanol product with nominally about 1 wt% water. Whereas, gas phase methanol synthesis results in a crude methanol product with 2-20 wt% water. This results in lower purification cost for the LPMEOH™ process for the Fuel Grade and MTBE Grade products. By applying the same cost estimation screening program, the distillation cost to produce Fuel Grade methanol from the LPMEOH™ Process which directly utilizes CO-rich syngas is about \$0.02 per gallon less than from the gas phase methanol process.

When the cost of the syngas stream is included, the results indicate that the LPMEOH™ Process has a \$0.04 - \$0.11 per gallon advantage over a gas phase process in the production of Fuel Grade methanol in an IGCC coproduction facility. Sensitivity studies performed as part of this Report indicate that the magnitude of the advantage in the methanol conversion cost for the LPMEOH™ Process when compared with the conventional gas phase process is increased when:

a) the syngas is rich in CO,
b) syngas is available at higher pressures,
c) only modest syngas conversion to methanol is required,
d) syngas is available with low hydrogen sulfide ($H_2S$) and carbonyl sulfide (COS) content,
e) inerts in the syngas (such as nitrogen in the oxygen from the air separation unit feeding the gasifier) are relatively high, and
f) Fuel Grade or MTBE Grade (low water) Methanol is required.

These results are based upon the operating expectations of the LPMEOH™ Process at the LPMEOH™ Demonstration Unit in Kingsport. Actual operation and subsequent evaluation is expected to lead to a number of improvements in future designs which will result in lower capital and operating costs. Studies show that the cost advantage of the LPMEOH™ Process over the conventional gas phase process could be increased by an additional \$0.02 - \$0.05 per gallon if increases in the rated capacity of the plant and improvements from the original plant design can be demonstrated as part of the four-year operating test phase of the Project.

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## ACRONYMS AND DEFINITIONS

| **Term:** | **Definition:** |
|---|---|
| Balanced Gas | A synthesis gas with a composition of hydrogen ($H_2$), carbon monoxide (CO) and carbon dioxide ($CO_2$) in stoichiometric balance for the production of methanol (approximately 2:1). |
| Catalyst Age ($\eta$) | Described in terms of the ratio of the catalyst rate constant at a given time to the rate constant of fresh catalyst; this is a design parameter used to set the catalyst addition/withdrawal frequency for a given methanol production rate. |
| CO Gas | A synthesis gas containing primarily CO. |
| CO-rich | A synthesis gas containing more CO than that required for the stoichiometric production of methanol. |
| CSTR | Continuous Stirred Tank Reactor, an important design parameter in terms of required reactor volume. |
| Feed Gas (Feed) | Synthesis gas “fed” to a reactor for synthesis. |
| Fresh Feed | Make-up synthesis gas to the methanol loop. |
| Fuel Gas | Total of all purge gas streams (reactor loop, distillation, catalyst reduction) returned to a fuel gas header. |
| $H_2$ Gas | A synthesis gas with a $H_2$ to CO ratio greater than 2. |
| IGCC | Integrated Gasification Combined Cycle, a type of electric power generation plant. |
| LHV | Abbreviation for Lower Heating Value. |
| LPMEOH™ | Liquid Phase Methanol (the technology to be demonstrated). |
| MEOH | Abbreviation for methanol. |
| Methanol Conversion Cost | The cost of production of methanol based upon the capital and operating charges within the battery limits of the methanol facility (excluding distillation). |
| M | Abbreviation for thousand. |
| MM | Abbreviation for million. |
| MTBE | Abbreviation for methyl tertiary butyl ether. |
| POX | Abbreviation for partial oxidation process. |
| ppmv | parts per million (volume basis). |
| ppmw | parts per million (weight basis). |
| psig | pounds per square inch (gauge). |
| Reactor Feed | The sum of the “Fresh Feed” and “Recycle Gas”. |
| Recycle Gas | The portion of unreacted synthesis gas effluent from the reactor, “recycled” as a feed gas. |
| Recycle Ratio | The ratio of “Recycle Gas” to “Fresh Feed”. |
| Space Velocity | Defined as the rate of inlet gas flow to the reactor per mass of catalyst; typical units are Sl/hr-kg catalyst oxide, or standard liters (standard conditions are 1 atm absolute and 0 deg C) per hour per kilogram catalyst (on an oxide basis). |
| sT/D | Abbreviation for short ton(s) per day. |
| Syngas | Abbreviation for Synthesis Gas. |

## ACRONYMS AND DEFINITIONS (cont'd)

| Term: | Definition: |
|---|---|
| Synthesis Gas | A gas containing primarily hydrogen ($H_2$), carbon monoxide (CO), or mixtures of $H_2$ and CO; intended for "synthesis" in a reactor to form methanol and/or other hydrocarbon products (Synthesis gas may also contain carbon dioxide ($CO_2$), water, and other gases). |
| Synthesis Gas Conversion | The percentage of lower heating value (LHV) energy content of the Fresh Feed which is converted to liquid product. |
| Syngas Usage | The synthesis gas used to produce the methanol product, defined as the amount of lower heating value (LHV) energy content of the Fresh Feed, less the (LHV) energy content of the Fuel Gas, per volume of Methanol; typical units are BTU/gallon. |
| vol% | Abbreviation for Volume %. |
| wt% | Abbreviation for Weight %. |

## Executive Summary

The Liquid Phase Methanol (LPMEOH™) Demonstration Project at Kingsport, Tennessee, is a $213.7 million cooperative agreement between the U.S. Department of Energy (DOE) and Air Products Liquid Phase Conversion Company, L.P. (the Partnership) to produce methanol from coal-derived synthesis gas (syngas). Air Products and Chemicals, Inc. (Air Products) and Eastman Chemical Company (Eastman) formed the Partnership to execute the Demonstration Project. The LPMEOH™ Process Demonstration Unit was designed, constructed, and is in operation at a site located at the Eastman chemicals-from-coal complex in Kingsport.

On 04 October 1994, Air Products and Eastman signed the agreements that would form the Partnership, secure the demonstration site, and provide the financial commitment and overall project management for the project. These partnership agreements became effective on 15 March 1995, when DOE authorized the commencement of Budget Period No. 2 (Modification No. A008 to the Cooperative Agreement). The Partnership has subcontracted with Air Products to provide the overall management of the project, and to act as the primary interface with DOE. As subcontractor to the Partnership, Air Products provided the engineering design, procurement, construction, and commissioning of the LPMEOH™ Process Demonstration Unit, and is providing the technical and engineering supervision needed to conduct the operational testing program required as part of the project. As subcontractor to Air Products, Eastman is responsible for operation of the LPMEOH™ Process Demonstration Unit, and for the interconnection and supply of syngas, utilities, product storage, and other needed services.

The project involves the operation of an 80,000 gallons per day (260 short tons per day (sT/D)) methanol unit utilizing coal-derived syngas from Eastman's integrated coal gasification facility. The new equipment consists of syngas feed preparation and compression facilities, the liquid phase reactor and auxiliaries, product distillation facilities, and utilities.

The technology to be demonstrated is the product of a cooperative development effort by Air Products and DOE in a program that started in 1981. Developed to enhance electric power generation using integrated gasification combined cycle (IGCC) technology, the LPMEOH™ process is ideally suited for directly processing gases produced by modern day coal gasifiers. Originally tested at a small, DOE-owned experimental unit in LaPorte, Texas, the technology provides several improvements essential for the economic coproduction of methanol and electricity directly from gasified coal. This liquid phase process suspends fine catalyst particles in an inert liquid, forming a slurry. The slurry dissipates the heat of the chemical reaction away from the catalyst surface, protecting the catalyst and allowing the methanol synthesis reaction to proceed at higher rates.

At the Eastman chemicals-from-coal complex, the technology is integrated with existing coal gasifiers. A carefully developed test plan will allow operations at Eastman to simulate electricity demand load-following in coal-based IGCC facilities. The operations will also demonstrate the enhanced stability and heat dissipation of the conversion process, its reliable on/off operation, and its ability to produce methanol as a clean liquid fuel without additional upgrading. An off-site, product-use test program will be conducted to demonstrate the suitability of the methanol product as a transportation fuel and as a fuel for stationary applications for small modular electric power generators for distributed power.

The four-year operating test phase and off-site product-use test program will demonstrate the commercial viability of the LPMEOH™ process and allow utilities to evaluate the application of this technology in the coproduction of methanol with electricity. A typical commercial-scale IGCC coproduction facility, for example, could be expected to generate 200 to 350 MW of electricity, and to also manufacture 45,000 to 300,000 gallons per day of methanol (150 to 1,000 sT/D). A successful demonstration at Kingsport will show the ability of a local resource (coal) to be converted in a reliable (storable) and environmentally preferable way to provide the clean energy needs of local communities for electric power and transportation.

This Topical Report compares the cost of methanol as produced from the LPMEOH™ Process and from a conventional gas phase process as applied to a generic 500 sT/D methanol plant as part of an IGCC coproduction facility. The cost of methanol is calculated as the sum of three terms: the methanol conversion cost (which includes the fixed and operating costs for the methanol unit), the distillation cost, and the syngas cost from the IGCC facility. A proprietary cost estimation screening program developed by R. B. Moore of Air Products was used to calculate the methanol conversion cost and the distillation cost from the LPMEOH™ Process and the gas phase process for various syngas supply pressures and on-stream factors. The methanol conversion cost from the LPMEOH™ Process is $0.02 to $0.07 per gallon lower than from the gas phase methanol process.

A major component of the methanol conversion cost in an IGCC complex is the cost to distill the crude methanol product in order to meet the final specification. It is typical for methanol to be stabilized (either by distillation or by deep flashing) to remove volatile components (such as carbon dioxide ($CO_2$)) and permit shipment and transport in atmospheric vessels. Beyond stabilization, other distillation may be necessary so that the final methanol product meets the specification for the designated end-use. Based upon the current applications of methanol, there are three grades of methanol product (Chemical Grade AA, Fuel Grade, and MTBE (methyl tertiary-butyl ether) Grade) which could be used in downstream chemical or power applications. These grades of methanol differ in the amounts of water and higher alcohols which are present in the final product. In particular, the Fuel Grade and MTBE Grade products have a maximum water specification of 1 wt%, while the Chemical Grade AA methanol has a maximum water content of 0.1 wt%.

The LPMEOH™ Process, which can directly process coal-derived syngas which is rich in carbon monoxide (CO), produces a crude methanol product with nominally about 1 wt% water. Whereas, gas phase methanol synthesis results in a crude methanol product with 2-20 wt% water, depending on the amount of $CO_2$ in the syngas which is converted to methanol and water. This results in lower purification cost for the LPMEOH™ Process for the Fuel Grade and MTBE Grade products. By applying the same cost estimation screening program, the distillation cost to produce Fuel Grade methanol from the LPMEOH™ Process which directly utilizes CO-rich syngas is about $0.02 per gallon less than from the gas phase methanol process.

When the cost of the syngas stream is included, the results indicate that the LPMEOH™ Process has a $0.04 - $0.11 per gallon advantage over a gas phase process in the production of methanol in an IGCC coproduction facility. Sensitivity studies performed as part of this Report indicate that the magnitude of the advantage in the methanol conversion cost for the LPMEOH™ Process when compared with the conventional gas phase process is increased when:

a) the syngas is rich in CO,
b) syngas is available at higher pressures,
c) only modest syngas conversion to methanol is required,
d) syngas is available with low hydrogen sulfide ($H_2S$) and carbonyl sulfide (COS) content,
e) inerts in the syngas (such as nitrogen in the oxygen from the air separation unit feeding the gasifier) are relatively high, and
f) Fuel Grade or MTBE Grade (low water) Methanol is required.

These results are based upon the operating expectations of the LPMEOH™ Process at the LPMEOH™ Demonstration Unit in Kingsport. The current design of the LPMEOH™ Demonstration Unit at Kingsport is based upon conservative performance calculations and also on an extensive amount of extra equipment and instrumentation required for evaluation of a new technology. Actual operation and subsequent evaluation is expected to lead to a number of improvements in future designs which will result in lower capital and operating costs. Studies show that the cost advantage of the LPMEOH™ Process over the conventional gas phase process could be increased by an additional $0.02 - $0.05 per gallon if increases in the rated capacity of the plant and the feasibility of these design improvements can be demonstrated as part of the four-year operating test phase of the Project.

## 1. Introduction

The LPMEOH™ Process is a very effective technology for converting much of the hydrogen ($H_2$) and CO in syngas generated by an integrated gasification combined cycle (IGCC) complex to methanol. The process is very flexible in being able to process many variations in syngas composition. Its biggest advantage over conventional gas phase methanol technology is the ability to process syngas which is rich in CO, without the need for shift conversion and $CO_2$ removal to adjust to the stoichiometric $(H_2\text{-}CO_2)/(CO+CO_2)$ ratio of about 2.1. This is possible as the liquid phase serves as a temperature moderator and heat removal medium. In a gas phase reactor, circulating $H_2$-rich gas is required for temperature moderation. Of course the richer the syngas is in CO the more the production is limited by the availability of $H_2$. Normally the least expensive methanol conversion cost comes from converting as much $H_2$ as is practical on a once-through basis without compression, recycle or further processing of the gas. The higher the pressure which the syngas is available, the greater is the degree of conversion and the lower the conversion cost. Also, the closer the $(H_2\text{-}CO_2)/(CO+CO_2)$ ratio is to 2.1, the greater the conversion and the lower the conversion cost.

As greater amounts of conversion are required, different techniques can be used to increase conversion. The simplest is the compression of the syngas to a higher pressure. Most of the test work to date has been for operating the LPMEOH™ Process at pressures under 1,000 psig; however, there is confidence in extrapolating up to pressures of 1,250 psig. The second method used to increase the degree of conversion is to condense methanol from the reactor effluent and to recycle part of the effluent gas back to the reactor inlet. A recycle ratio of 1:1, recycle:fresh feed, is usually quite effective in optimizing the methanol production. Above a 2:1 recycle little is gained with CO-rich syngas as most of the available $H_2$ has already been consumed. If higher conversion is desired it becomes necessary to generate additional $H_2$. This is done by the addition of water/steam to the syngas before it passes through the reactor. The methanol catalyst performs under these conditions a (partial) water gas shift. In this case, the increase in conversion is accompanied with a modest increase of water in the crude methanol product and $CO_2$ in the recycle gas. The $CO_2$ produced may be partially or totally removed from the recycle stream, as required, to further improve the degree of conversion.

### 1.1 Process Flow Diagram

Figure 1 shows the **Process Flow Diagram LPMEOH™ Synthesis**. In this Report, the IGCC complex is assumed to be operating in conjunction with an air separation unit producing a syngas stream with an inerts concentration of 1 vol%. In the simplest configuration, syngas at its maximum available pressure from the IGCC power plant (Stream 1) is passed through the LPMEOH™ Process and unreacted syngas (Stream 3) is returned to the IGCC plant. As additional conversion is required beyond the kinetic limit of once-through syngas at delivery pressure, the syngas is compressed to increase the production. Current economic analysis indicate that the limit might be about 1,250 psig for once-through operation. If higher conversion is still required, unreacted syngas can be recycled back to the reactor inlet. At this point, the process reaches the limitation imposed by chemical equilibrium for a specific syngas composition. Equilibrium can be shifted favorably by addition of a steam/water injection system to the reactor feed gas and a $CO_2$ removal unit to the recycle gas stream; this combination is required for coal-derived syngas conversion greater than about 50% by volume of the ($H_2$+CO) in the syngas. Each of these parameters will be evaluated as they impact the methanol conversion cost as produced in an IGCC complex.

### 1.2 Typical Synthesis Gas Conversion

The following discussion of syngas conversion for the production of methanol provides information on the ($H_2$+CO) conversion as a function of syngas composition and of two other reactor design variables: syngas feed pressure and space velocity (SV). The calculations are made at two fixed design conditions: the assumption that the slurry reactor is operating as a single continuous stirred tank reactor (CSTR), implying a perfectly back-mixed system; and that the methanol synthesis catalyst has been aged to a target value. The “age” of the methanol synthesis catalyst can be expressed in terms of a dimensionless variable eta ($\eta$), which is defined as the ratio of the rate constant at any time to the rate constant for freshly reduced catalyst (as determined in the laboratory autoclave). For all cases in this Report, the value used for $\eta$ is 0.5.

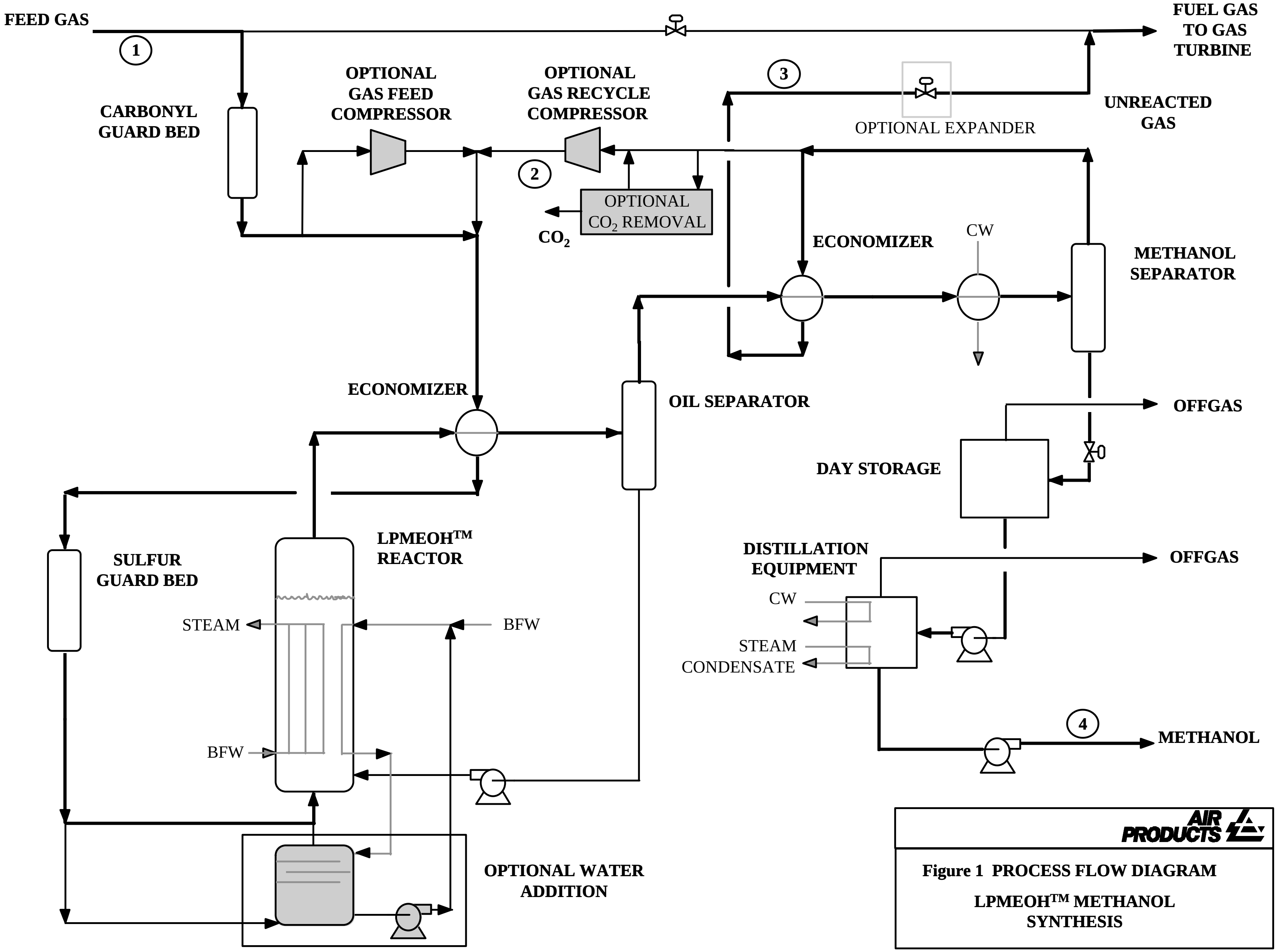



**Figure 1 PROCESS FLOW DIAGRAM**

**LPMEOH™ METHANOL SYNTHESIS**

### 1.2.1 Balanced Gas

As the syngas composition approaches the ideal $(H_2-CO_2)/(CO+CO_2)$ ratio of 2.1 the conversion per pass is relatively high. A syngas close to this composition is obtained with the partial oxidation of natural gas, or the reforming of natural gas supplemented with about 30% imported $CO_2$ . Figure 2, **Balanced Synthesis Gas Conversion Vs Pressure,** shows that at an inlet pressure of 500 psig and a space velocity of 8,000 Sl/hr-kg catalyst oxide, about 15% of the $(H_2+CO)$ is converted to methanol. With a lower space velocity of 4,000 (more catalyst in a bigger reactor) the conversion per pass is about 23%. Greater conversion up to about 52% can be achieved with feed gas compressed to 1,250 psig. Of course, with a syngas at the stoichiometric ratio and with a high recycle of unconverted syngas, it is possible to convert virtually all of the syngas to methanol. Also shown is the conversion at a space velocity of 0, which represents equilibrium and the maximum conversion possible without condensing product and recycling gas.

### 1.2.2 Heavy Oil Residue Synthesis Gas

A common oil residue-based syngas composition has a $(H_2-CO_2)/(CO+CO_2)$ ratio of 1. In a refinery this is obtained by the partial oxidation (POX) of heavy oils. Figure 3, **Oil Synthesis Gas Conversion Vs Pressure,** shows that a syngas of this composition will be produce less methanol in a single pass when compared with a stoichiometrically balanced gas. For example, even with the more favorable SV of 4,000, only 17% can be converted at 500 psig on a once-through basis. Even with compression to 1,250 psig and a 1:1 recycle to fresh feed ratio, the maximum practical conversion appears to be about 61%. Also shown is the conversion at a space velocity of 0, which represents equilibrium and the maximum conversion possible without recycle.

### 1.2.3 Coal-derived Synthesis Gas

Most of the LPMEOH™ development work has been directed towards converting a portion of coal-derived syngas produced in a coal gasifier to methanol. This syngas will typically have a $H_2/CO$ ratio below 0.7; the amount of $CO_2$ varies with the type of gasification technology. The most likely application for the LPMEOH™ Process is the coproduction of methanol as part of a coal based IGCC facility. As can be seen from Figure 4, **Coal Synthesis Gas Conversion Vs Pressure,** that with gas this lean in $H_2$, at a SV of 2,000 and 500 psig, about 18% of the $(H_2+CO)$ can be converted to methanol on a once-through basis. Even with compression, 1:1 recycle:fresh feed and an optimum SV of 4,000, only about 46% of the $(H_2+CO)$ can be converted to methanol, as there is insufficient $H_2$ for further conversion. This Figure shows that with 1:1 recycle, 15 vol% water addition for partial shift conversion and $CO_2$ removal for recycled gas, 51% of the $(H_2+CO)$ can be converted to methanol. Of course with additional recycle, shift conversion and $CO_2$ removal to provide a stoichiometric 2.1:1 balanced syngas virtually all of the syngas can be converted to methanol. Also shown is the conversion at a space velocity of 0, which is the maximum conversion possible with once-through operation.

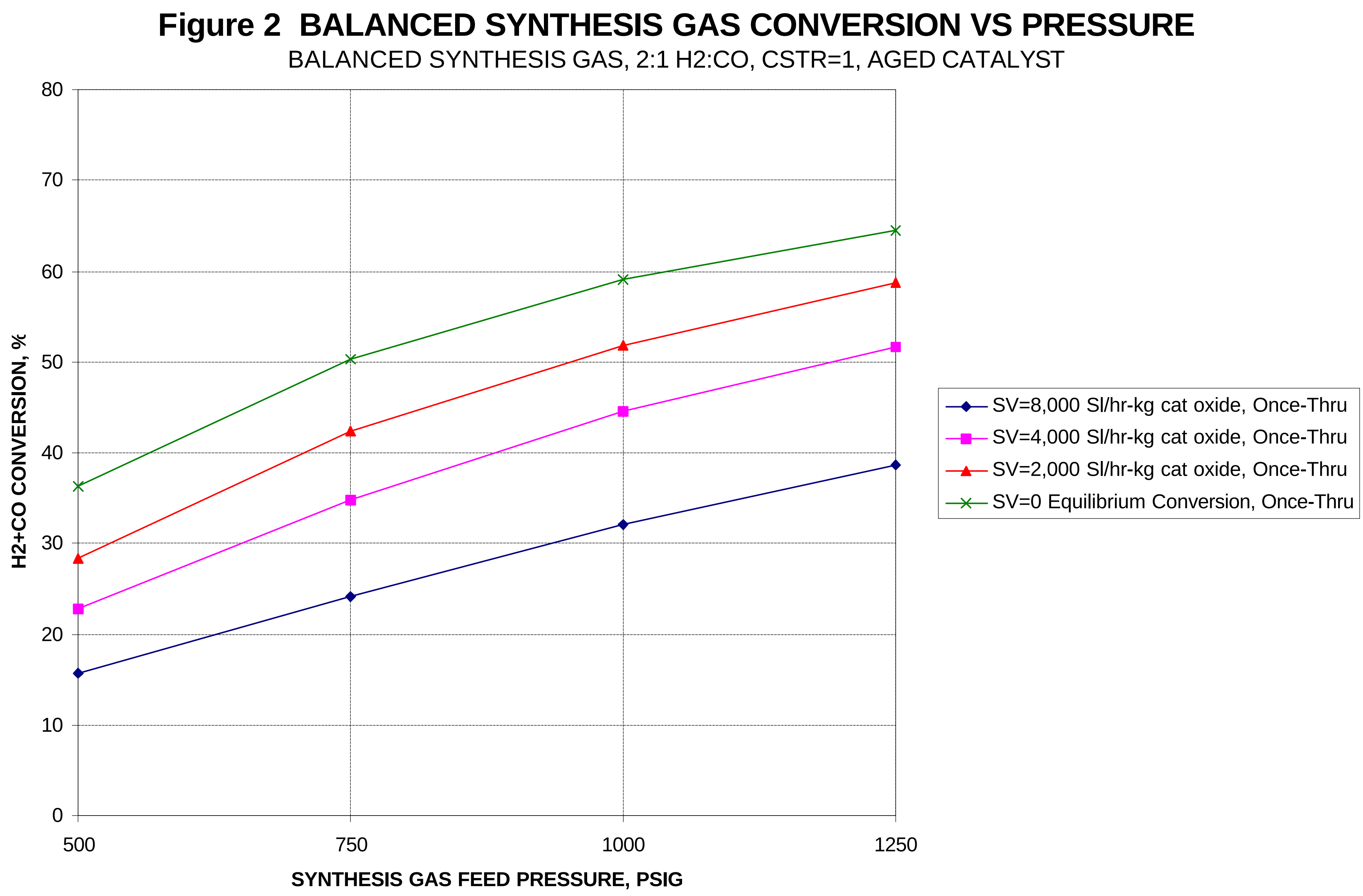
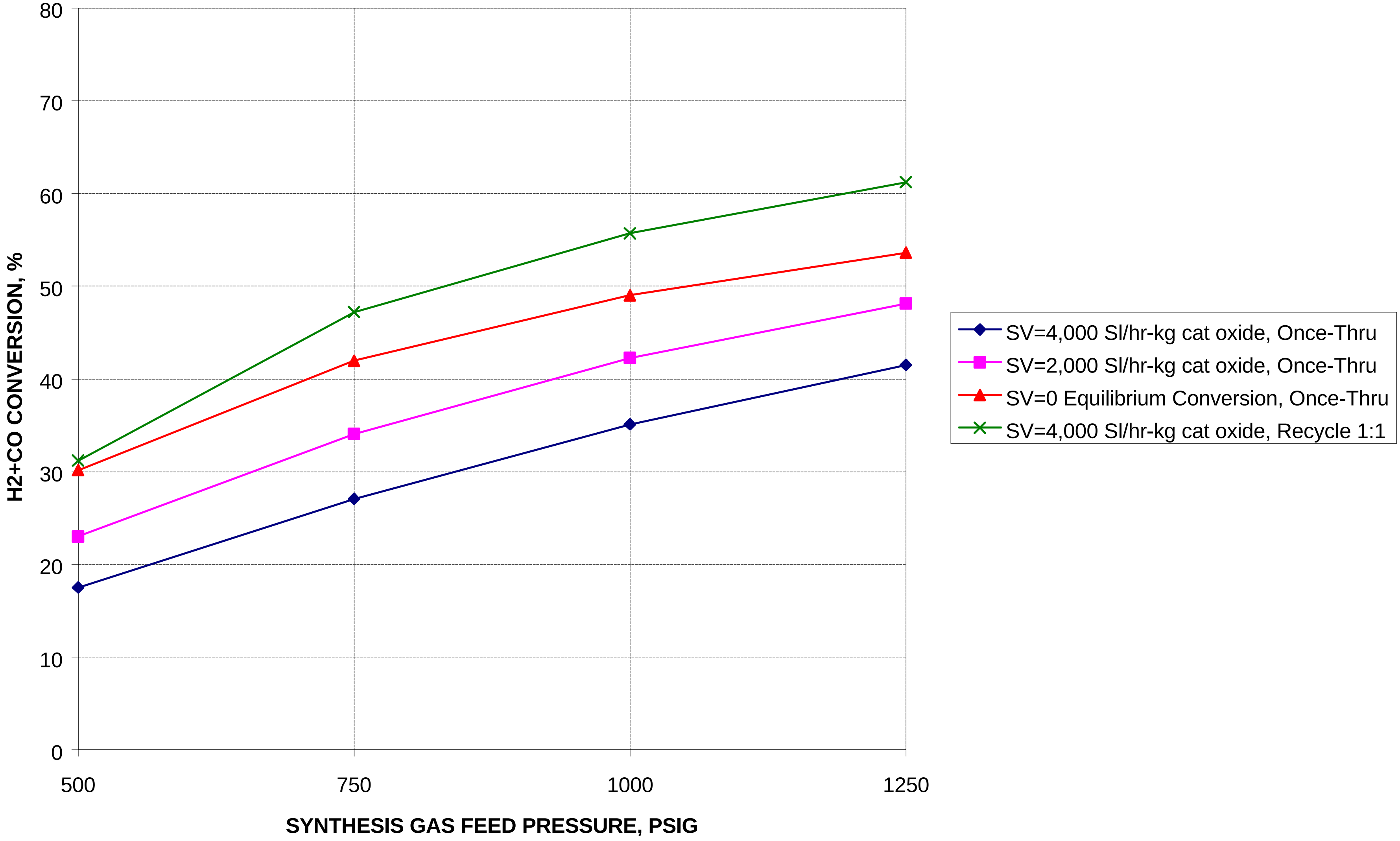


Figure 2 BALANCED SYNTHESIS GAS CONVERSION VS PRESSURE
BALANCED SYNTHESIS GAS, 2:1 H2:CO, CSTR=1, AGED CATALYST
H2+CO CONVERSION, %
0
10
20
30
40
50
60
70
80
500
750
1000
1250
SYNTHESIS GAS FEED PRESSURE, PSIG
SV=8,000 Sl/hr-kg cat oxide, Once-Thru
SV=4,000 Sl/hr-kg cat oxide, Once-Thru
SV=2,000 Sl/hr-kg cat oxide, Once-Thru
SV=0 Equilibrium Conversion, Once-Thru

## Figure 3 OIL SYNTHESIS GAS CONVERSION VS PRESSURE

HEAVY RESIDUE POX SYNTHESIS GAS, 0.97:1 H2:CO, CSTR=1, AGED CATALYST



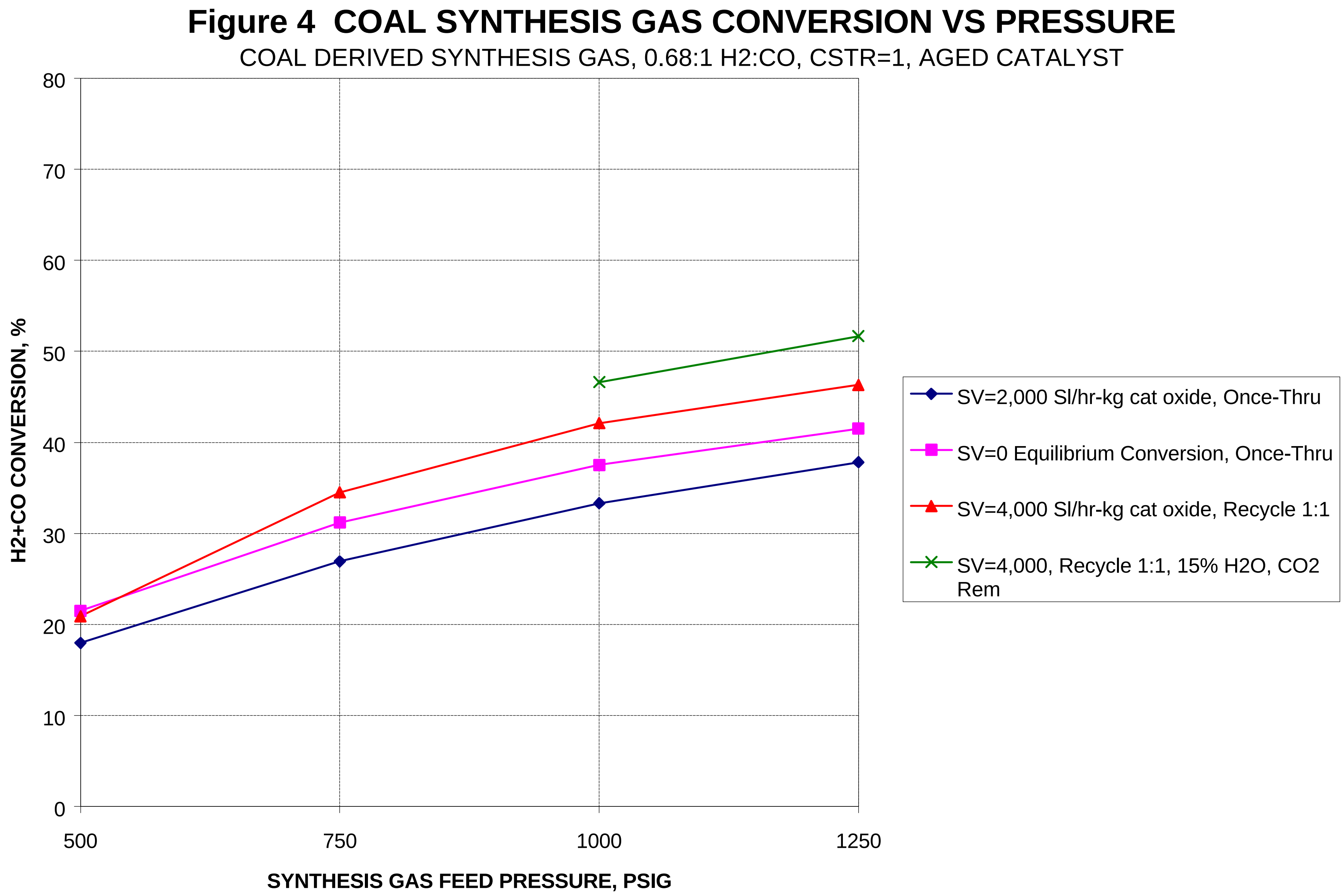


Figure 4 COAL SYNTHESIS GAS CONVERSION VS PRESSURE
COAL DERIVED SYNTHESIS GAS, 0.68:1 H2:CO, CSTR=1, AGED CATALYST
80
70
60
50
40
30
20
10
0
H2+CO CONVERSION, %
500
750
1000
1250
SYNTHESIS GAS FEED PRESSURE, PSIG
SV=2,000 Sl/hr-kg cat oxide, Once-Thru
SV=0 Equilibrium Conversion, Once-Thru
SV=4,000 Sl/hr-kg cat oxide, Recycle 1:1
SV=4,000, Recycle 1:1, 15% H2O, CO2 Rem

## 1.3 <u>Distillation</u>

A major component of the methanol conversion cost in an IGCC complex is the cost to distill the crude methanol product in order to meet the final specification.

It is typical for methanol to be stabilized (either by distillation or by deep flashing) to remove volatile components (such as $CO_2$) and permit shipment and transport in atmospheric vessels. Beyond stabilization, other distillation may be necessary so that the final methanol product meets the specification for the designated end-use. Based upon the current applications of methanol, there are three grades of methanol product which could be used in downstream chemical or power applications.

### 1.3.1 Chemical Grade AA Methanol

This grade of methanol product meets the Federal Specifications for 99.85% by weight (wt%) minimum purity of methanol, 0.1 wt% water maximum, and concentrations of higher alcohols at parts-per-million levels. Significant capital investment (either a two- or a three-column distillation system) and significant consumption of low-pressure steam are required to upgrade the as-produced methanol from gas-phase technologies to this conventional methanol product.

### 1.3.2 Fuel Grade Methanol

Fuel Grade Methanol is defined as a methanol with a relaxed specification on water and other impurities when compared with Chemical Grade AA Methanol. Studies in both automotive and power applications have shown that the presence of heavier hydrocarbons (such as ethanol and the inert liquid medium from the LPMEOH™ Process) do not impact either the operation of this equipment or the environmental characteristics of the combustion gas stream. When methanol is produced directly from a CO-rich syngas in an IGCC application, the water content in the crude methanol can meet this specification. This can result in significant capital and operating savings when compared to conventional gas phase processes, which produce a crude methanol with water concentrations well in excess of 1 wt%. The specification used in this study to produce the Fuel Grade Methanol is a minimum of 97 wt% methanol, a maximum of 1 wt% water, 1.5 wt% higher alcohols, and 0.5 wt% process oil (the inert liquid medium from the LPMEOH™ Process).

### 1.3.3 MTBE Grade Methanol

An intermediate grade of methanol (between Chemical Grade AA and Fuel Grade) has the potential to meet the needs for the production of methyl tertiary butyl ether (MTBE), an additive which improves both the oxygen content and octane of unleaded gasoline. Preliminary testing of Fuel Grade Methanol (as defined in Section 1.3.2) has indicated that the concentrations of acidic components in the crude methanol (in particular, methyl formate and methyl acetate) can adversely impact the MTBE production plant by corrosion within the methanol recovery equipment. As a result, additional stabilization of the crude methanol beyond flash separation or distillation of $CO_2$ is required. Again, the presence of higher alcohols, the inert liquid medium from the LPMEOH™ Process, and water up to 1 wt% concentration appeared to have no impact on the operation of the MTBE production unit. The specification used in this study to produce the MTBE Grade Methanol is a minimum of 97 wt% methanol, a maximum of 1 wt% water, 2 wt% higher alcohols, 150 ppmw methyl acetate, and 0.3 wt% inert liquid medium.

## 2. Cost Estimation Screening Program

The methods used in developing costs estimates for the various cases studied in this Report are combined in a single, proprietary, PC-based spreadsheet program developed by R. B. Moore, Senior Engineering Associate at Air Products. This program is a screening tool which provides costs for applications of the LPMEOH™ Process based upon the capital and expected operating costs from the LPMEOH™ Demonstration Unit at Kingsport. The cost basis within the screening program includes first-time costs associated with the execution of the Demonstration Test Plan; the potential impact of these costs are quantified in Section 7.

### 2.1 Spreadsheet Inputs

The spreadsheet requires the following information as inputs:

1) Flow, composition, temperature, and pressure of the fresh feed syngas
2) LPMEOH™ Reactor operating conditions, including pressure, space velocity, and recycle ratio
3) Desired CO conversion
4) Desired methanol product purity
5) Capacity factor for the LPMEOH™ plant
6) Required on-site methanol storage

The spreadsheet performs an overall material balance using a proprietary kinetic model developed by Air Products to calculate the performance of the LPMEOH™ Reactor for a specific set of operating conditions. One of the important parameters in the sizing of the LPMEOH™ Reactor is the age of the catalyst, $\eta$, used in the design calculations. As noted in Section 1.2, the value used in this Report for $\eta$ is 0.5. Unless otherwise noted, the capacity factor for the LPMEOH™ plant is assumed to be 90%.

### 2.2 Capital Cost Calculations

Once the material balance has been calculated, the spreadsheet then calculates the investment cost summary. These are categorized as follows:

Capital costs for equipment, valves, and instrumentation are calculated using costs from the LPMEOH™ Demonstration Unit as the basis (calendar year 1995). Different multipliers are used to scale the costs from the 80,000 gallons-per-day (260 sT/D) design for the Kingsport facility. For example, costs associated with methanol distillation to the desired grade of methanol might scale directly with methanol production rates, while some of the costs within the LPMEOH™ Reactor system may depend upon the volumetric gas flowrate.

Construction costs assume the same schedule (15 months) and site preparation as the LPMEOH™ Demonstration Unit.

Freight and Miscellaneous refer to project execution costs, again based upon the costs from the LPMEOH™ Demonstration Project.

Air Products Process Studies refer to initial process evaluations which need to be performed in order to optimize the process cycle selected for a particular IGCC application. For example, optimization of feed and recycle compression and an evaluation of the need for water injection are performed during these studies.

Air Products Technical Package refers to the final process design package for a site-specific application. This includes all heat and material balances, detailed equipment specifications for the LPMEOH™ Reactor, and process specifications for other equipment items.

Project Engineering refers to the overall management of the engineering, design, and construction work, based upon the costs from the LPMEOH™ Demonstration Project.

Design Engineering refers to the detailed design work (civil, mechanical, instrument and electrical), again based upon the costs from the LPMEOH™ Demonstration Project.

Field Engineering refers to all field engineering services, including construction management at the site. These costs are again based upon the costs from the LPMEOH™ Demonstration Project.

Travel and Living costs are included as a line item.

Reserve is provided as a line item, but is not included in these evaluations.

License Fee is provided as a line item, but is not included in these evaluations.

Once these items are computed, an allowance for Owner's Cost (25% of the equipment cost for the evaluated system) is provided. This is a nominal cost of project execution which includes costs for:

- Initial charge of catalyst
- Initial supply of chemicals and lubricants
- Any applicable taxes and insurance
- Cost of land
- Legal and other overhead costs

Costs associated with Methanol Storage and $CO_2$ Removal (if required) are separated from other equipment items, since the costs are site-specific and are highlighted in this manner to provide a potential customer with the opportunity to optimize the required storage.

The sum of these items is the Capital Cost for the LPMEOH™ Facility. The costs are also tabulated by process area (excluding Owner's Cost, Methanol Storage Cost, and $CO_2$ Removal Costs) to provide information on the relative costs of the different process steps, including all valves, instruments, and associated construction costs.

A calculation of the Plot Area Required, based upon the layout from the LPMEOH™ Demonstration Unit, is also given.

### 2.3 Operating Cost Calculations

Operating costs are evaluated as annual costs ($/Year) and in the cost per gallon of methanol. The annual cost is computed from the cost per unit time and the annual capacity factor. The final cost of methanol is computed as follows:

$$\text{\$/Gallon Methanol} = \frac{\text{Annual Costs (\$/Year)}}{365 * LF * 303 * X}$$

where:

LF = Annual Capacity Factor
X = Methanol Production Rate, short Tons per Day

The conversion factor 303 gallons per short ton of methanol is used in this equation.

The following is a summary of the components of the Operating Costs:

Syngas refers to the value (lower heating value basis) of the syngas produced in the IGCC facility. When a specific application is defined, the cost ($/MMBtu) can be specified.

Unreacted Gas refers to the value (lower heating value basis) of the syngas returned from the LPMEOH™ plant to the IGCC facility. When a specific application is defined, the cost ($/MMBtu) can be specified.

Power used by the LPMEOH™ Process is evaluated at $0.04/kilowatt-hour, which is assumed to be the electricity costs from the adjacent IGCC facility. Categories of power consumers (compressors, pumps, etc.) are tabulated separately.

Steam is a credit from the LPMEOH™ Plant; a typical value of $4.00/1,000 pounds of steam production (including the cost for boiler feed water and the credit for condensate return) is used for the 200 psig steam produced by the LPMEOH™ Reactor.

Cooling Water is evaluated at a 20°F temperature rise across all cooling water heat exchangers. A cost of $0.12/1,000 gallons of cooling water is used in this study.

Other Miscellaneous Utilities (such as instrument air and nitrogen) are calculated based upon the costs from the LPMEOH™ Demonstration Project.

Catalyst, Chemicals and Lubes refers to estimated costs for the methanol synthesis catalyst, process mineral oil, and other lubricants. These values are based upon the budgeted costs from the LPMEOH™ Demonstration Project. Once operational history from the LPMEOH™ Demonstration Unit is acquired, these costs will be adjusted.

Zinc Oxide is used as the absorbent for the hydrogen sulfide which is present with the fresh feed syngas or is produced by the COS (carbonyl sulfide) Hydrolysis catalyst within the LPMEOH™ Process. Each of these consumables are evaluated based upon present costs for each material.

Operating Labor is calculated from the original budgetary forecast for the LPMEOH™ Demonstration Project.

Maintenance costs are evaluated based upon an annual budget of 2% of the total capital cost for the evaluated LPMEOH™ Facility.

Property Taxes and Insurance are estimated at an annual budget of 1.5% of the total capital cost for the evaluated LPMEOH™ Facility.

Overhead is computed at a nominal rate of 15% of the operating costs (not including the costs for syngas or the credit for the unreacted gas).

Recovery of Capital Cost, Depreciation, and ROI (Return on Investment) are computed based upon a 15 year depreciation and a ROI of about 14%, which results in a total charge of 20% of the investment per year.

For purposes of this report, Fixed Costs are defined as sum of the Capital Cost, Depreciation, and ROI, the Operating Labor, the Maintenance, the Property Taxes and Insurance, and the associated Overhead for these items. All other costs are totaled as Variable Costs.

### 2.3.1 Distillation Costs

Distillation Costs are provided separately based upon the type of methanol product (Fuel Grade, MTBE Grade, or Chemical Grade):

Power costs are evaluated at $0.04/kilowatt-hour, which is assumed to be the electricity costs from the adjacent IGCC facility.

Steam is a cost for distillation of the methanol from the LPMEOH™ Plant; a typical value of $3.00/1,000 pounds of steam (including the cost for boiler feed water and the credit for condensate return) is used for the 100 psig steam used in the distillation area.

Cooling Water is evaluated at a 20°F temperature rise across all cooling water heat exchangers in the distillation area. A cost of $0.12/1,000 gallons of cooling water is used in this study.

The cost of incremental Operating Labor for the added distillation equipment is evaluated at an additional operator per year, at an annual cost of $50,000 per operator.

Maintenance costs for distillation are evaluated based upon an annual budget of 2% of the total capital cost in the distillation area.

Property Taxes and Insurance are estimated at an annual budget of 1.5% of the total capital cost for the distillation area.

Overhead is computed at a nominal rate of 15% of the operating costs within the distillation area.

Recovery of Capital Cost, Depreciation, and ROI (Return on Investment) are computed based upon a 15 year depreciation and a ROI of about 14%, which results in a total charge of 20% of the investment per year.

### 2.4 Sample from Cost Estimation Screening Program - Case 5-HW

In order to provide an example of the detailed output from the cost estimation screening program, a Case designated 5-HW using compression of syngas from 500 to 1,250 psig, 1:1 recycle:fresh feed ratio and 5 vol% water addition to the LPMEOH™ Reactor feed gas, was chosen for illustration of the generated material balance, investment and operating costs (refer to Section 2 for definition of terms). Table 1 contains the **LPMEOH™ Material Balance** for this case, a 500 sT/D LPMEOH™ facility, operating with a space velocity of 4,000 Sl/hr-kg catalyst oxide, processing "Texaco type" coal-derived syngas ($H_2/CO$ = 0.68, 13 vol% $CO_2$, 1 vol% inerts, represented by nitrogen). Table 2, **Facility Investment Summary**, is a summary of the investment costs for Case 5-HW, broken down by type of cost as well as major section costs (as defined in Section 2.2). As noted earlier, the Investments for Area A Reactor, and Area B Condensation and Separation, carry a high cost for steel structure, instrumentation and analytical equipment currently used in the LPMEOH™ Demonstration Unit at Kingsport. Although future commercial facilities will have none of these first-time costs, no other basis is available for developing costs for the LPMEOH™ Process; these costs will be used in all cases evaluated in this Report. Table 3, **Operating Cost Data**, summarizes the resulting methanol conversion cost, $0.230/gal. Also shown is the purification costs for distilling the final product to Fuel Grade Methanol, $0.017/gal, for a total of $0.247/gal.

## Table 1 LPMEOH™ MATERIAL BALANCE

Case 5-HW
SV 4000

| | | |
|---|---|---|
| LPMEOH™ PRODUCTION | CRUDE | 500 sT/D |
| | STABILIZED | 500 sT/D |
| CO Conversion | 38 % | SV = 4000 |
| H2 + CO Conversion | 51.61 % | |
| Water Addition, LB-MOL/HR | 464 | |
| CO2 Removal, 1=YES, 0=NO | 0 | |
| Syngas Utilization | 70.1 MBTU/GAL | |

| | (1) FEED FRESH SYNGAS | | (2) RECYCLE | |
|---|---|---|---|---|
| **COMP.** | **LB-MOL/HR** | **Vol%** | **LB-MOL/HR** | **Vol%** |
| H2 | 3245 | 35 | 1583 | 17 |
| CO | 4728 | 51 | 4958 | 53 |
| CO2 | 1205 | 13 | 2476 | 27 |
| C1+C2+C3 | 0 | 0 | 7 | 0 |
| N2 | 93 | 1 | 157 | 2 |
| H2O | 0 | 0 | 1 | 0 |
| CH3OH | 0 | 0 | 89 | 1 |
| OTHER | 0 | 0 | 0 | 0 |
| COS, PPMV | 5 | 0 | 0 | 0 |
| TOTAL | 9270 | 100 | 9270 | 100 |
| TEMP, F | 103 | | 100 | |
| PSIA | 515 | | 1189 | |
| MW | 20.7 | | 27.4 | |
| MMBTU HHV | 975 | | | |
| MMBTU LHV | 913 | | | |

| | (3) UNREACTED GAS | | (4) METHANOL | |
|---|---|---|---|---|
| **COMP.** | **LB-MOL/HR** | **Vol%** | **LB-MOL/HR** | **Vol%** |
| H2 | 933 | 17 | 0 | 0 |
| CO | 2924 | 53 | 0 | 0 |
| CO2 | 1460 | 27 | 0 | 0 |
| CH4 | 4 | 0 | 0 | 0 |
| N2 | 93 | 2 | 0 | 0 |
| H2O | 0 | 0 | 13 | 1 |
| CH3OH | 52 | 1 | 1300 | 98 |
| OTHER | 0 | 0 | 13 | 1 |
| TOTAL | 5467 | 100 | 1326 | 100 |
| TEMP, F | 100 | | 100 | |
| PSIA | 1189 | | 14.7 | |
| MW | 33.0 | | 32.0 | |
| MMBTU HHV | 489 | | 358 | |
| MMBTU LHV | 469 | | | |

## Table 2 FACILITY INVESTMENT SUMMARY

Case 5-HW
SV 4000

| INVESTMENT BREAKDOWN | | | MM-$ |
|---|---|---|---|
| COMPRESSION | | | $1.52 |
| LPMEOH™ EQUIPMENT | | | $5.14 |
| VALVES & INSTRUMENTS | | | $4.15 |
| CONSTRUCTION | | | $11.02 |
| FREIGHT & MISCELLANEOUS | | | $0.50 |
| AIR PRODUCTS PROCESS STUDIES | | | $0.22 |
| AIR PRODUCTS TECHNICAL PACKAGE | | | $0.97 |
| PROJECT ENGINEERING | | | $2.18 |
| DESIGN ENGINEERING | | | $5.20 |
| FIELD ENGINEERING | | | $1.54 |
| TRAVEL & LIVING | | | $0.39 |
| RESERVE | | | $0.00 |
| LICENSE FEE | | | not incl |
| | | Sub-total Turnkey Plant | $32.82 |
| OWNER'S COST | 25% of Equipment | | $1.67 |
| METHANOL STORAGE | 30 days | 5.0 MM Gallons | $2.46 |
| CO2 REMOVAL | | | $0.00 |
| | | TOTAL CAPITAL | $36.94 |

**PLOT AREA REQUIRED, ACRES**

| | | | |
|---|---|---|---|
| LPMEOH™ = 0.11 | Storage = 5.13 | TOTAL | 5.24 |

**INVESTMENT BY AREA**

| AREA | INVESTMENT | DESIGN | BASIS | MM-$ |
|---|---|---|---|---|
| A | REACTOR LOOP & CATALYST REDUCTION | 500 | sT/D | $19.06 |
| B | FEED COMPRESSION | 5665 | BHP | $4.37 |
| C | RECYCLE COMPRESSION | 370 | BHP | $2.86 |
| D | FRONT-END GAS CLEANUP | 9270 | LB-MOL/HR | $2.18 |
| E | COMMON EQUIPMENT | 500 | sT/D | $2.58 |
| F | SATURATOR | | | $0.58 |
| | RESERVE | 0.0 | % | $0.00 |
| | AIR PRODUCTS PROCESS STUDIES | | | $0.22 |
| | AIR PRODUCTS TECHNICAL PACKAGE | | | $0.97 |
| | LICENSE FEE | | | not incl |
| | | Sub-total Turnkey Plant | | $32.82 |

## Table 3 OPERATING COST DATA

Case 5-HW
SV 4000

**METHANOL CONVERSION ANNUAL OPERATING COST**

| | | | | | M-$/Yr | $/Gal |
|---|---|---|---|---|---|---|
| MeOH Production | 500 | sT/D | | | | |
| Annual Load Factor | 90 | % | | | | |
| Syngas (LHV) | 913 | MMBTU/hr | $0.00 | /MMBTU | $0 | $0.000 |
| Unreacted Gas (LHV) | 469 | MMBTU/hr | $0.00 | /MMBTU | $0 | $0.000 |
| Power | | | | | | |
| Feed Compressor | 4532 | kW | $0.04 | /kWh | $1,429 | $0.029 |
| Recycle Compressor | 296 | kW | $0.04 | /kWh | $93 | $0.002 |
| Pumps, Heaters, etc. | 382 | kW | $0.04 | /kWh | $121 | $0.002 |
| CO2 Removal | 0 | kW | $0.04 | /kWh | $0 | $0.000 |
| MP Steam, 200 psig | (49300) | lb/hr | $4.00 | /M-lb | ($1,555) | ($0.031) |
| C Water, 20oF Delta T | 2100 | gpm | $0.12 | /M-gal | $119 | $0.002 |
| Misc Utilities | | | | | $240 | $0.005 |
| Catalyst, Chemicals & Lubes | | | | | $989 | $0.020 |
| Sulfur Removal, Zinc Oxide | 66 | M-lb/yr | $3.58 | /lb | $214 | $0.004 |
| COS Hydrolysis Catalyst | 4 | M-lb/yr | $3.63 | /lb | $12 | $0.000 |
| Operating Labor | | | | | $683 | $0.014 |
| Maintenance | 2% | of Investment/yr | | | $739 | $0.015 |
| | | | | Sub-Total | $3,085 | $0.062 |
| Property Taxes, Insurance | 1.5% | of Investment/yr | | | $554 | $0.011 |
| Overhead | 15% | of Oper Costs (Less Feed) | | | $463 | $0.009 |
| Capital Costs, Depr & ROI | 20% | of Investment/yr | | | $7,388 | $0.148 |
| | | Methanol Conversion Cost, Total | | | $11,490 | $0.230 |

**DISTILLATION ANNUAL OPERATING COST** FUEL GRADE - Texaco Syngas

| | | | | | M-$/Yr | $/Gal |
|---|---|---|---|---|---|---|
| MeOH Production | 500 | sT/D | | | | |
| Annual Load Factor | 90 | % | | | | |
| Electric Power | 43 | kW | $0.04 | /kWh | $13 | $0.000 |
| LP Steam, 100 psig | 3515 | lb/hr | $3.00 | /M-lb | $83 | $0.002 |
| C Water, 20oF Delta T | 252 | gpm | $0.12 | /M-gal | $14 | $0.000 |
| Operating Labor | 1 | man/yr | 50 | $-M/yr | $50 | $0.001 |
| Maintenance | 2% | of Investment/yr | | | $56 | $0.001 |
| Property Taxes, Insurance | 1.5% | of Investment/yr | | | $42 | $0.001 |
| Overhead | 15% | of Oper Costs (Less Feed) | | | $39 | $0.001 |
| Capital Costs, Depr & ROI | 20% | of Investment/yr | | | $555 | $0.011 |
| | | Purification Cost, Total | | | $852 | $0.017 |

## 3. Results and Discussion

### 3.1 Conversion Cost for Methanol Production from Texaco-Type Synthesis Gas

In this section, the methanol conversion cost is calculated for a variety of cases. The composition for Texaco-type (CO-rich) syngas is provided in Section 2.4. For purposes of this Report, methanol conversion cost is defined as the cost of production of methanol based upon the capital and operating charges within the battery limits of the methanol facility. Distillation costs are excluded from this analysis, and are covered in Section 3.2.

#### 3.1.1 Optimum Space Velocity as a Function of Operating Pressure

In developing the series of cases to evaluate the methanol conversion cost from a CO-rich syngas in an IGCC complex, a separate evaluation of the impact of space velocity on methanol conversion cost was performed. Three cases were evaluated (Table 4):

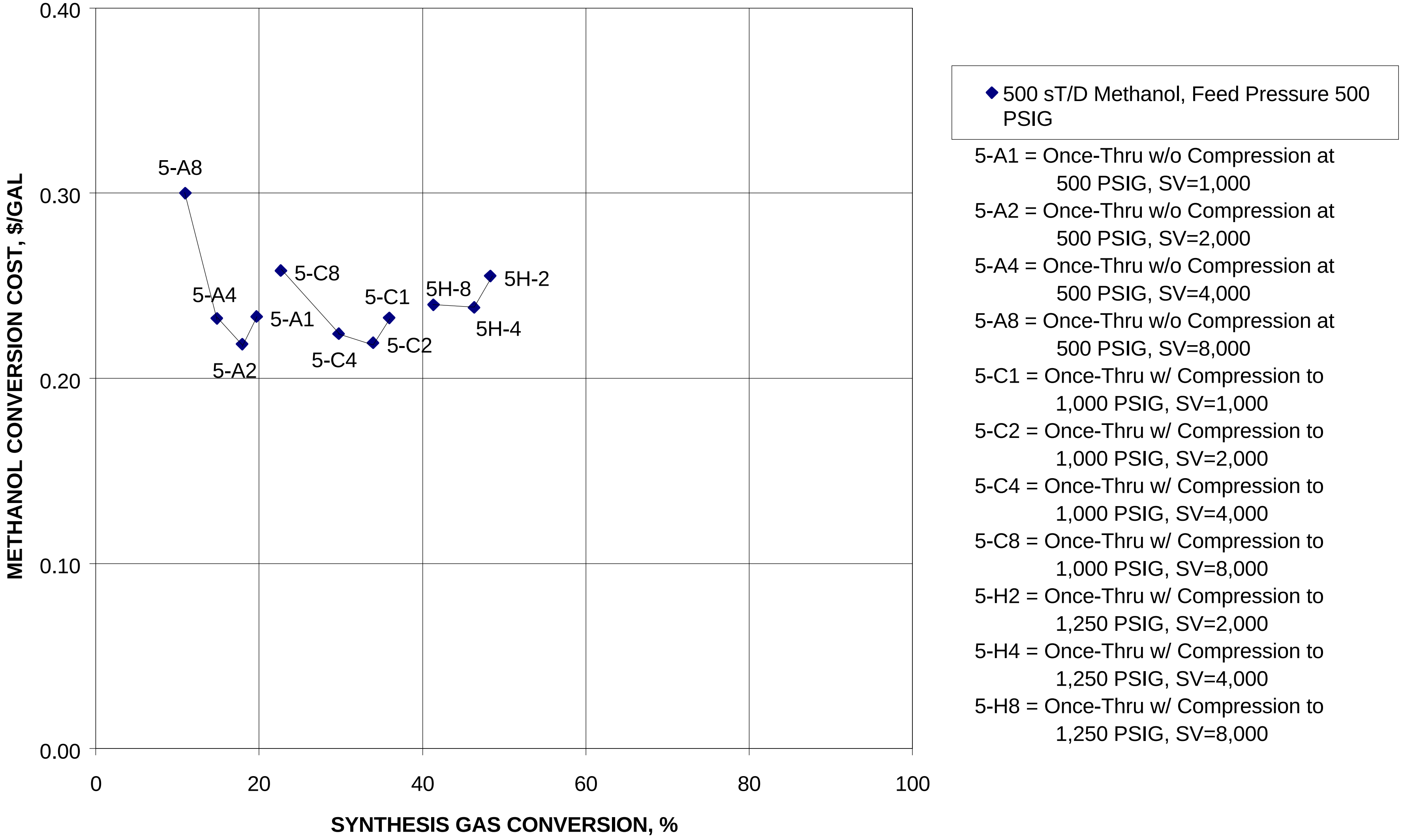
**TABLE 4**
**STUDY CASES - SPACE VELOCITY VS. PRESSURE**

| Case | Flow Scheme | Compression | Pressure, psig |
|---|---|---|---|
| 5-A | Once-Through | Without | 500 |
| 5-C | Once-Through | With | 1,000 |
| 5-H | 1:1 Recycle | With | 1,250 |

Figure 5, **Space Velocity Optimization**, for syngas available at **500 psig** shows the methanol conversion cost for once-through operation (Case 5-A) at space velocities of 1,000, 2,000, 4,000 and 8,000. Also shown are the results for once-through operation with the syngas compressed to 1,000 psig (Case 5-C) for the same space velocities. Cases for syngas compressed to 1,250 psig and with 1:1 recycle:fresh feed (Case 5-H) are shown for space velocities of 2,000, 4,000, and 8,000. For the once-through cases it appears that a space velocity of about 2,000 is the optimum and for the 1:1 recycle:fresh feed case the optimum space velocity is about 4,000. These simulations were made assuming design conditions of aged catalyst with a catalyst age ($\eta$) of 0.5 (50% of new catalyst activity). Also the LPMEOH™ Reactor was assumed to have an effective Continuous Stirred Tank Reactor (CSTR) = 1 stage. The operating test data from Kingsport facility will ultimately be used to determine an optimum ($\eta$) and the actual CSTR stages. This data is expected to increase the optimum space velocity for all cases.

## Figure 5 SPACE VELOCITY OPTIMIZATION

TEXACO-TYPE SYNTHESIS GAS, 0.68 H2:CO



### 3.1.2 Methanol Conversion Cost from CO-rich Syngas - 500 psig Feed Gas Pressure

The following table (Table 5) is a summary of the cases developed for operation on CO-rich syngas according to the process flowsheet in Figure 1. All cases assume that CO-rich syngas is available from the IGCC complex at 500 psig. Space velocity is selected based upon the analysis in Section 3.1.1.

**TABLE 5**
**STUDY CASES - 500 PSIG FEED GAS PRESSURE**

| Case | Flow Scheme | Compression | Reactor Pressure, psig | Space Velocity, Sl/hr-kg |
|---|---|---|---|---|
| 5-A | Once-Through | Without | 500 | 2,000 |
| 5-AW | Once-Through with 5 vol% water | Without | 500 | 2,000 |
| 5-B | Once-Through | With | 750 | 2,000 |
| 5-C | Once-Through | With | 1,000 | 2,000 |
| 5-D | Once-Through | With | 1,250 | 2,000 |
| 5-DW | Once-Through with 5 vol% water | With | 1,250 | 2,000 |
| 5-H | 1:1 Recycle | With | 1,250 | 4,000 |
| 5-HW | 1:1 Recycle with 5 vol% water | With | 1,250 | 4,000 |
| 5-HC5 | 1:1 Recycle with 5 vol% water and $CO_2$ removal | With | 1,250 | 4,000 |
| 5-HC15 | 1:1 Recycle with 15 vol% water and $CO_2$ removal | With | 1,250 | 4,000 |
| 5-M | 4.86:1 Recycle with Shift and $CO_2$ removal | With | 1,040 | |

In Case 5-A, the LPMEOH™ facility the CO-rich syngas from the IGCC complex in a once-though mode of operation. The syngas is passed once-through a LPMEOH™ facility. Case 5-AW is the same as Case 5-A except that 5 vol% water is added with the syngas to increase methanol production by shifting CO to $H_2$ in the LPMEOH™ Reactor. Case 5-B is again once-through operation with the feed compressed from 500 psig to 750 psig to increase methanol production. Case 5-C is once-through operation with the feed compressed from 500 psig to 1,000 psig and in Case 5-D the feed is compressed to 1,250 psig. The effect of 5 vol% water addition at 1,250 psig operation to increase methanol production was developed in Case 5-DW.

Case 5-H was developed to show the greater production achieved by both compressing the syngas from 500 psig to 1,250 psig, and recycling the reactor effluent syngas back to the reactor after first condensing the produced methanol. This case used a 1:1 recycle:feed ratio. Case 5-HW is similar to Case 5-H except 5 vol% water vapor was added to the incoming syngas. As a greater fraction of the syngas is converted to methanol with water addition and recycle operation, the amount of $CO_2$ builds up to a reaction limiting level. To eliminate this limit, $CO_2$ removal from the recycle gas is shown in Case 5-HC5, which is the same otherwise as Case 5-HW. With $CO_2$ removal it is practical to use greater amounts of water addition as shown in Case 5-HC15 which has 15 vol% water addition. A final reference case was developed to show the comparable conversion cost using a conventional gas phase methanol synthesis loop. In this Case 5-M, CO-rich syngas available at 500 psig is shifted and the resulting $CO_2$ is removed to produce a balanced syngas, $(H_2\text{-}CO_2)/(CO+CO_2) = 2.1$. This gas was then compressed to 1,040 psig and converted to methanol in a gas phase reactor using a recycle ratio of 4.86:1 recycle:fresh feed. These are typical conditions for the production of methanol in conventional gas-phase processes.

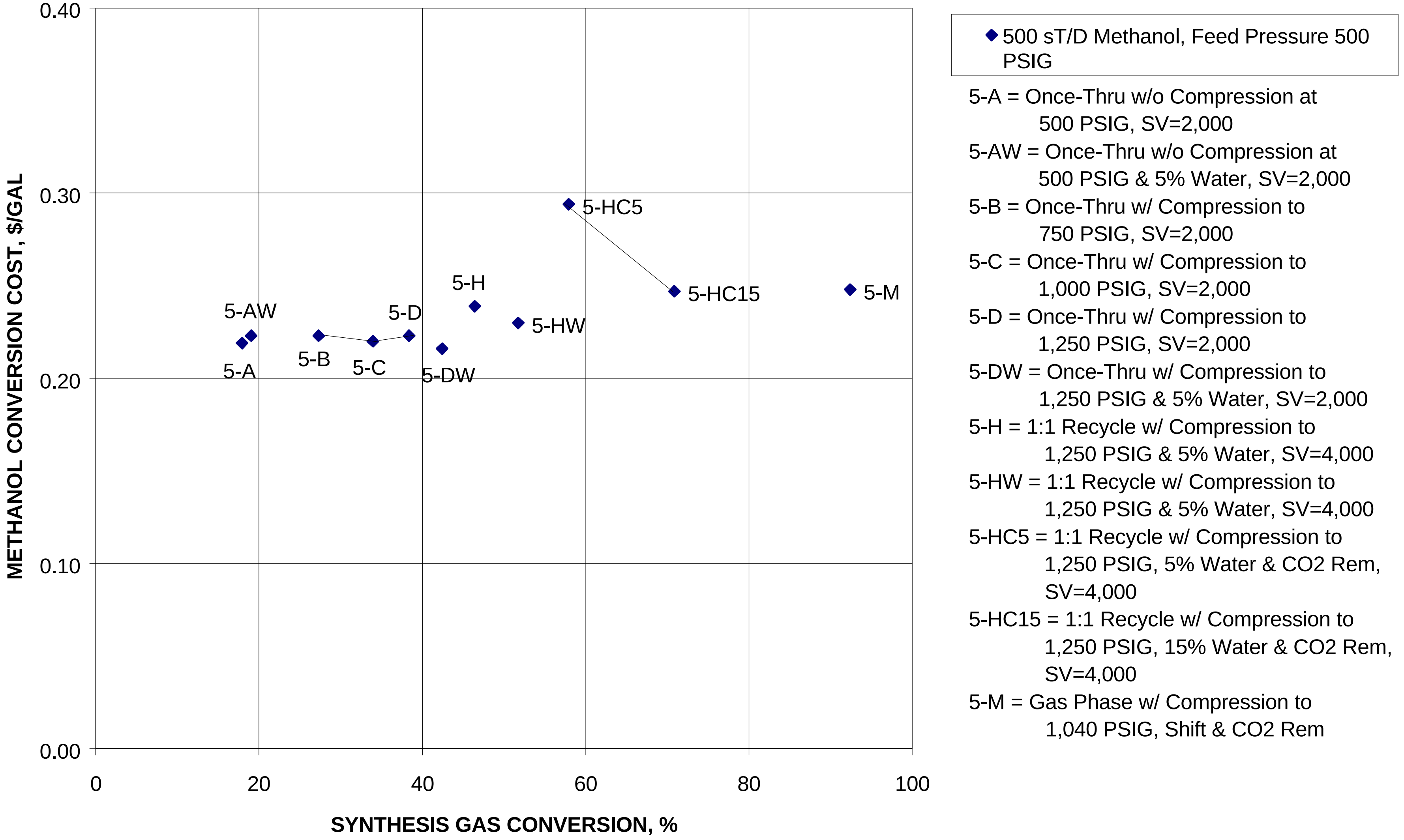
In Figure 6, **Conversion Cost Vs Percent Conversion**, for syngas available at **500 psig**, the least expensive conversion cost is for uncompressed syngas. Syngas conversion of about 18% can be obtained for a conversion cost of about $0.219/gal (Case 5-A). Additional conversion up to about 38% can be obtained at a modest cost ($0.223/gal) by compressing the feed gas up to 1,250 psig (Case 5-D). With compression to about 1,250 psig and 1:1 recycle:fresh feed (Case 5-H) about 46% syngas conversion can be obtained at a conversion cost of about $0.239/gal. Additional conversion to about 52% can be obtained with water addition (Case 5-HW) with a marginally lower methanol conversion cost of $0.230/gal. Greater amount of conversion can only be obtained at the expense of more water addition and of $CO_2$ removal. At a pressure of 1,250 psig, with 1:1 recycle to feed, water addition and $CO_2$ removal (Case 5-HC15), the syngas conversion can be pushed to 71% at an average conversion cost of about $0.247/gal.

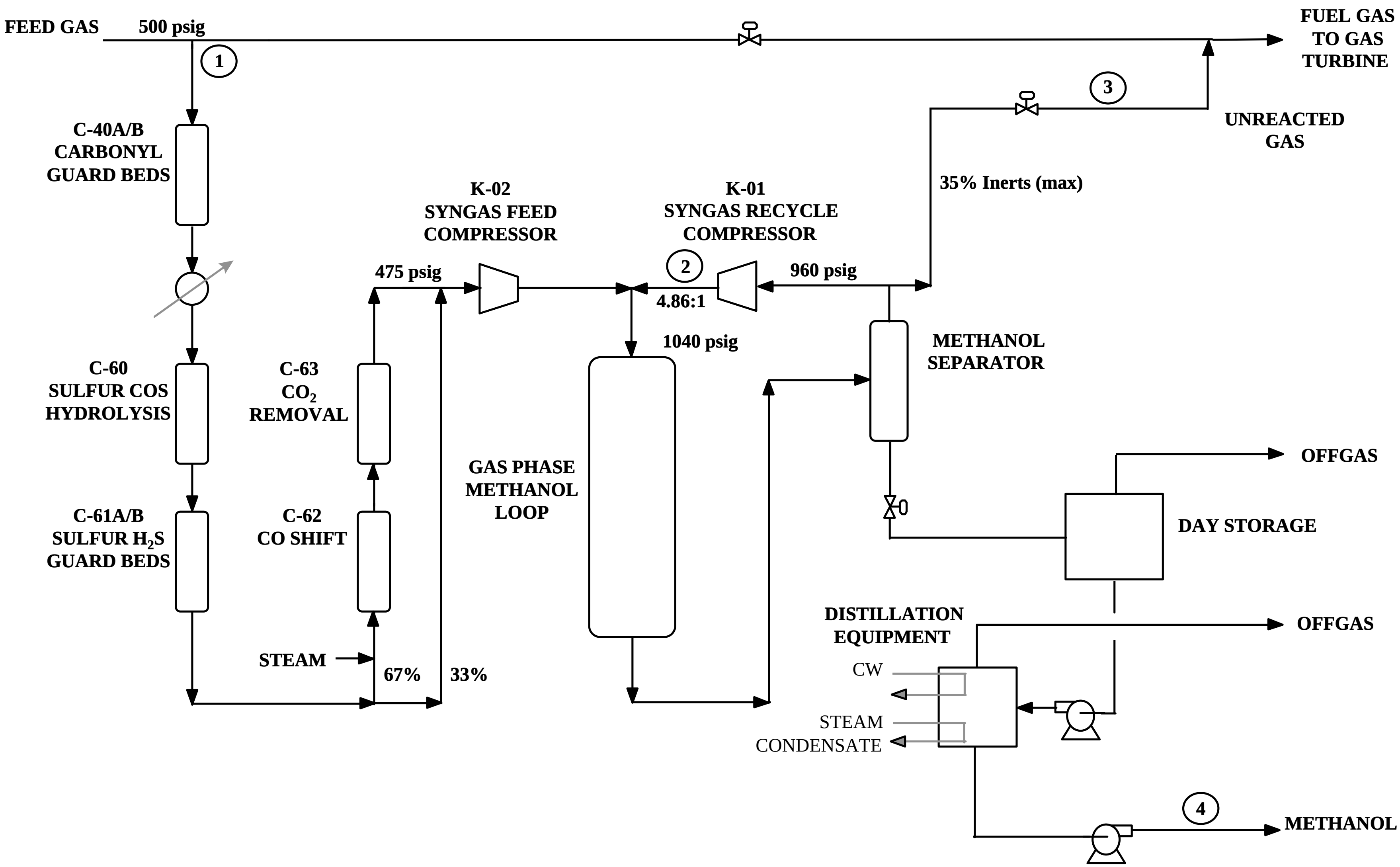
With the gas phase process, shift conversion and $CO_2$ removal are required to produce a balanced gas feeding the methanol synthesis loop. This is shown on Figure 7, **Process Flow Diagram Gas Phase Methanol Synthesis**. Using recycle about 92% of the balanced gas can then be converted to methanol. Table 6, **Gas Phase Investment Summary**, is a summary of the investment costs for the gas phase process (Case 5-M) divided into major section costs. When comparing the gas phase process with the LPMEOH™ Process the cost of the reactor and loop for the gas phase may be somewhat less than for the LPMEOH™ Process. This difference is expected to be reduced in future designs as the LPMEOH™ Process is further optimized. However, more than offsetting the difference in cost of the methanol synthesis reactors is the cost of the shift conversion and $CO_2$ removal equipment required for the production of a balanced syngas feed required by the gas phase process. Table 7, **Gas Phase Operating Cost Data**, summarizes the resulting conversion cost to methanol, $0.248/gal.

Also noted on Table 5 is an example of the distillation costs for methanol from a gas phase facility. The distillation costs for purifying the final product to Fuel Grade would add $0.041/gal for a total of $0.289/gal. A detailed discussion of distillation costs is provided in Section 3.2.

**Figure 6 CONVERSION COST VS PERCENT CONVERSION**

TEXACO-TYPE SYNTHESIS GAS, 0.68:1 H2:CO






AIR PRODUCTS

**Figure 7 PROCESS FLOW DIAGRAM**

**GAS PHASE METHANOL SYNTHESIS**

## Table 6 GAS PHASE INVESTMENT SUMMARY

| **INVESTMENT BREAKDOWN** | Production sT/D | 500 | | **MM-$** |
|---|---|---|---|---|
| SULFUR REMOVAL | | | | $1.38 |
| SHIFT CONVERSION | | | | $3.91 |
| CO2 REMOVAL | | | | $7.83 |
| SYNGAS COMPRESSION & RECYCLE | | | | $5.21 |
| METHANOL LOOP | Conventional quench cooled converter * | | | $16.78 |
| RESERVE | | | | $0.00 |
| LICENSE FEE | | | | included |
| | | | Sub-total Turnkey Plant | $35.12 |
| OWNER'S COST | 25% of Equipment | | | $1.76 |
| METHANOL STORAGE | 30 days | 5.0 MM Gallons | | $2.46 |
| | | | TOTAL CAPITAL | $39.33 |

**PLOT AREA REQUIRED, ACRES**

| | | | |
|---|---|---|---|
| Methanol = 0.11 | Storage = 5.13 | TOTAL | 5.24 |

* Scaled from Chem Systems "Methanol 93-1" Report

# Table 7 GAS PHASE OPERATING COST DATA

**METHANOL CONVERSION ANNUAL OPERATING COST**

| | | | | | M-$/Yr | $/Gal |
|---|---|---|---|---|---|---|
| MeOH Production | 500 | sT/D | | | | |
| Annual Load Factor | 90 | % | | | | |
| Syngas (LHV) | 486 | MMBTU/hr | $0.00 | /MMBTU | $0 | $0.000 |
| Unreacted Gas (LHV) | 41 | MMBTU/hr | $0.00 | /MMBTU | $0 | $0.000 |
| Power | | | | | | |
| Feed Compressor | 1687 | kW | $0.04 | /kWh | $532 | $0.011 |
| Recycle Compressor | 1083 | kW | $0.04 | /kWh | $341 | $0.007 |
| Pumps, Heaters, etc. | 0 | kW | $0.04 | /kWh | $0 | $0.000 |
| CO2 Removal | 781 | kW | $0.04 | /kWh | $246 | $0.005 |
| LP Steam, 100 psig | 36461 | lb/hr | $3.00 | /M-lb | $862 | $0.017 |
| IP Steam, 200 psig | (49300) | lb/hr | $4.00 | /M-lb | ($1,555) | ($0.031) |
| MP Steam, 600 psig | 18127 | lb/hr | $4.50 | /M-lb | $643 | $0.013 |
| C Water, 20oF Delta T | 4152 | gpm | $0.12 | /M-gal | $236 | $0.005 |
| Catalyst, Chemicals & Lubes | | | | | $499 | $0.010 |
| Sulfur Removal, Zinc Oxide | 35 | M-lb/yr | $3.58 | /lb | $114 | $0.002 |
| COS Hydrolysis Catalyst | 2 | M-lb/yr | $3.63 | /lb | $6 | $0.000 |
| Operating Labor | | | | | $683 | $0.014 |
| Maintenance | 2% | of Investment/yr | | | $787 | $0.016 |
| | | | | Sub-Total | $3,395 | $0.068 |
| Property Taxes, Insurance | 1.5% | of Investment/yr | | | $590 | $0.012 |
| Overhead | 15% | of Oper Costs (Less Feed) | | | $509 | $0.010 |
| Capital Costs, Depr & ROI | 20% | of Investment/yr | | | $7,866 | $0.158 |
| | | Methanol Conversion Cost, Total | | | $12,361 | $0.248 |

**DISTILLATION ANNUAL OPERATING COST** FUEL GRADE - Balanced Gas

| | | | | | M-$/Yr | $/Gal |
|---|---|---|---|---|---|---|
| MeOH Production | 500 | sT/D | | | | |
| Annual Load Factor | 90 | % | | | | |
| Electric Power | 87 | kW | $0.04 | /kWh | $27 | $0.001 |
| LP Steam, 100 psig | 35954 | lb/hr | $3.00 | /M-lb | $850 | $0.017 |
| C Water, 20oF Delta T | 2985 | gpm | $0.12 | /M-gal | $169 | $0.003 |
| Operating Labor | 1 | man/yr | 50 | $-M/yr | $50 | $0.001 |
| Maintenance | 2% | of Investment/yr | | | $67 | $0.001 |
| Property Taxes, Insurance | 1.5% | of Investment/yr | | | $50 | $0.001 |
| Overhead | 15% | of Oper Costs (Less Feed) | | | $182 | $0.004 |
| Capital Costs, Depr & ROI | 20% | of Investment/yr | | | $671 | $0.013 |
| | | Purification Cost, Total | | | $2,068 | $0.041 |

The following table (Table 8) summarizes the investment, fixed and variable costs, and methanol conversion costs for the 11 cases at a Feed Gas pressure of 500 psig:

**TABLE 8**
**METHANOL CONVERSION COST - 500 PSIG FEED GAS PRESSURE**

| Case | Investment, $-M | Fixed Cost, $/Gal | Variable Cost, $/Gal | Conversion Cost, $/Gal |
|---|---|---|---|---|
| 5-A | 39,540 | 0.205 | 0.014 | 0.219 |
| 5-AW | 40,559 | 0.210 | 0.013 | 0.223 |
| 5-B | 35,320 | 0.185 | 0.038 | 0.223 |
| 5-C | 33,211 | 0.175 | 0.045 | 0.220 |
| 5-D | 32,906 | 0.173 | 0.050 | 0.223 |
| 5-DW | 32,424 | 0.171 | 0.045 | 0.216 |
| 5-H | 37,851 | 0.197 | 0.042 | 0.239 |
| 5-HW | 36,940 | 0.192 | 0.038 | 0.230 |
| 5-HC5 | 42,279 | 0.242 | 0.052 | 0.294 |
| 5-HC15 | 40,565 | 0.210 | 0.044 | 0.247 |
| 5-M | 39,329 | 0.204 | 0.044 | 0.248 |

For the majority of the cases utilizing the LPMEOH™ Process, the methanol conversion cost is approximately $0.02/gallon less than the gas phase process. This indicates that the initial optimization work for considering the coproduction of methanol in an IGCC application can focus upon the desired syngas conversion, as many of the operating scenarios result in a similar methanol conversion cost of about $0.22/gallon.

### 3.1.3 Methanol Conversion Cost from CO-rich Syngas - 1,000 psig Feed Gas Pressure

The following table (Table 9) is a summary of the cases developed for operation on CO-rich syngas available from the IGCC complex at 1,000 psig, according to the process flowsheet in Figure 1. As in the prior analysis, space velocity is selected based upon the analysis in Section 3.1.1.

**TABLE 9**
**STUDY CASES - 1,000 PSIG FEED GAS PRESSURE**

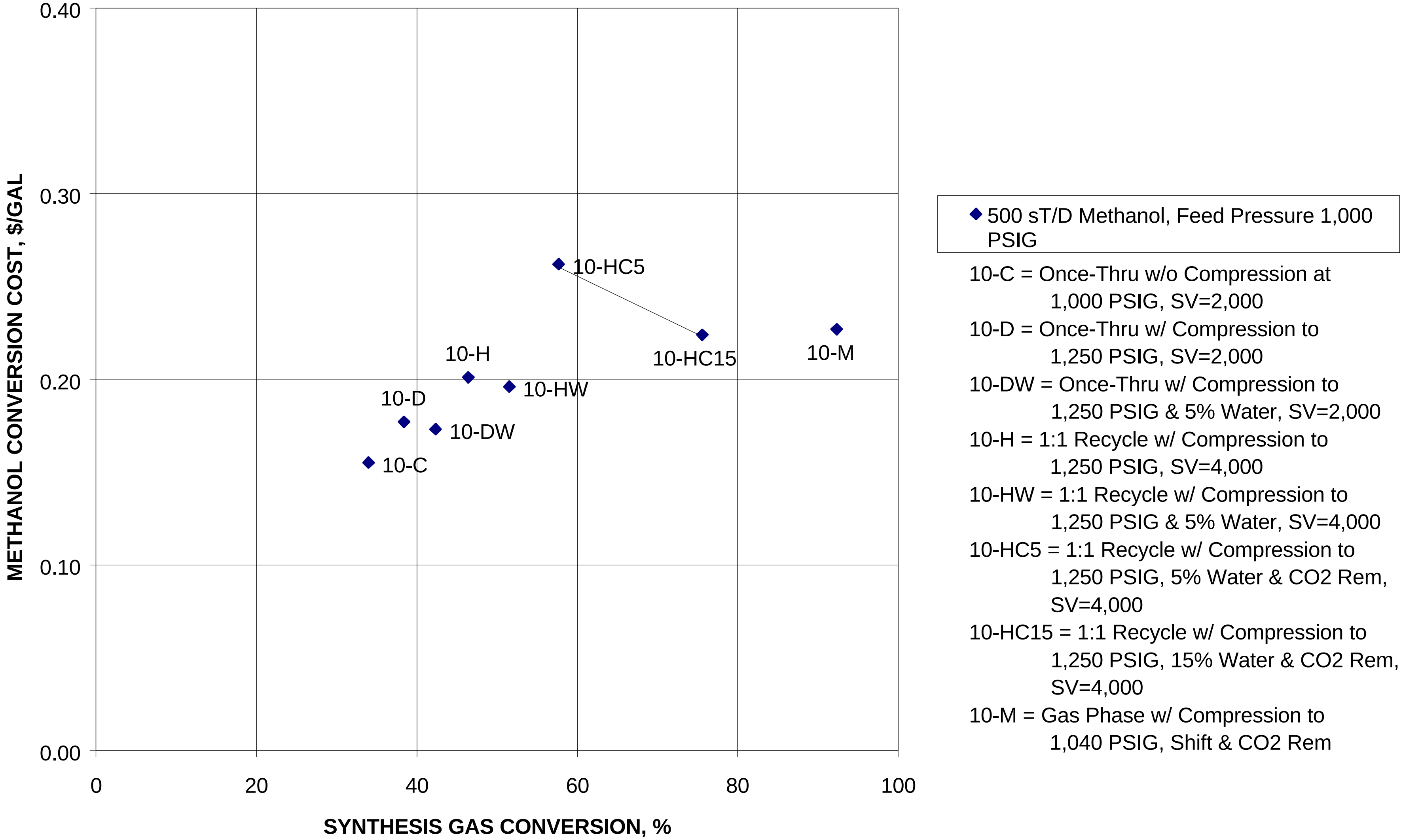
| Case | Flow Scheme | Compression | Reactor Pressure, psig | Space Velocity, Sl/hr-kg |
|---|---|---|---|---|
| 10-C | Once-Through | Without | 1,000 | 2,000 |
| 10-D | Once-Through | With | 1,250 | 2,000 |
| 10-DW | Once-Through with 5 vol% water | With | 1,250 | 2,000 |
| 10-H | 1:1 Recycle | With | 1,250 | 4,000 |
| 10-HW | 1:1 Recycle with 5 vol% water | With | 1,250 | 4,000 |
| 10-HC5 | 1:1 Recycle with 5 vol% water and $CO_2$ removal | With | 1,250 | 4,000 |
| 10-HC15 | 1:1 Recycle with 15 vol% water and $CO_2$ removal | With | 1,250 | 4,000 |
| 10-M | 4.86:1 Recycle with Shift and $CO_2$ removal | With | 1,040 | |

The descriptions for the letter designation of these cases can be found in Section 3.1.2.

Figure 8 shows the **Conversion Cost Vs Percent Conversion**, for syngas available at **1,000 psig**. The results are similar to the results for 500 psig feed gas pressure (Figure 6), other than that the product cost is \$0.03 to \$0.07/gal lower as the result of reduced compression requirements. The effect is the greatest at 34% syngas conversion, (Case 10-C), \$0.065/gal, and falls to about \$0.035/gal at 52% syngas conversion (Case 10-HW).

**Figure 8 CONVERSION COST VS PERCENT CONVERSION**

TEXACO-TYPE SYNTHESIS GAS, 0.68:1 H2:CO



The following table (Table 10) summarizes the investment, fixed and variable costs, and methanol conversion costs for the 8 cases at a Feed Gas pressure of 1,000 psig:

**TABLE 10**
**METHANOL CONVERSION COST - 1,000 PSIG FEED GAS PRESSURE**

| Case | Investment, $-M | Fixed Cost, $/Gal | Variable Cost, $/Gal | Conversion Cost, $/Gal |
|---|---|---|---|---|
| 10-C | 28,287 | 0.151 | 0.004 | 0.155 |
| 10-D | 30,714 | 0.163 | 0.014 | 0.177 |
| 10-DW | 30,402 | 0.161 | 0.012 | 0.173 |
| 10-H | 36,012 | 0.188 | 0.013 | 0.202 |
| 10-HW | 35,272 | 0.184 | 0.012 | 0.196 |
| 10-HC5 | 45,687 | 0.234 | 0.028 | 0.262 |
| 10-HC15 | 39,900 | 0.207 | 0.017 | 0.224 |
| 10-M | 37,653 | 0.196 | 0.031 | 0.227 |

Results for the gas phase process (Case 10-M) show a similar level of improvement at the CO-rich syngas supply pressure of 1,000 psig over the 500 psig case (Case 5-M). However, the methanol conversion cost for the gas phase process is still greater ($0.07/gal) than the once-through LPMEOH™ Process (Case 10-C), due to the higher capital cost for the shift conversion and $CO_2$ removal system.

Also provided for reference is Table 11, **Facility Investment Summary**, and Table 12, **Operating Cost Data**, which are outputs from the Cost Estimation Screening Spreadsheet for Case 10-C for **1,000 psig** syngas feed pressure, once-through, with ~ 34% syngas conversion. The resulting conversion cost to methanol is $0.155/gal. Including the cost to distill the methanol to the Fuel Grade specification, the total is $0.172/gal. This shows the economics of a LPMEOH™ plant built in conjunction with an IGCC facility where syngas is produced at 1,000 psig.

# Table 11 FACILITY INVESTMENT SUMMARY

Case 10-C
SV 2000

| INVESTMENT BREAKDOWN | | | | MM-$ |
|---|---|---|---|---|
| COMPRESSION | | | | \$0.00 |
| LPMEOH™ EQUIPMENT | | | | \$4.81 |
| VALVES & INSTRUMENTS | | | | \$3.10 |
| CONSTRUCTION | | | | \$8.22 |
| FREIGHT & MISCELLANEOUS | | | | \$0.37 |
| AIR PRODUCTS PROCESS STUDIES | | | | \$0.22 |
| AIR PRODUCTS TECHNICAL PACKAGE | | | | \$0.97 |
| PROJECT ENGINEERING | | | | \$1.63 |
| DESIGN ENGINEERING | | | | \$3.88 |
| FIELD ENGINEERING | | | | \$1.15 |
| TRAVEL & LIVING | | | | \$0.29 |
| RESERVE | | | | \$0.00 |
| LICENSE FEE | | | | not incl |
| | | | Sub-total Turnkey Plant | \$24.63 |
| OWNER'S COST | 25% of Equipment | | | \$1.20 |
| METHANOL STORAGE | 30 days | 5.0 MM Gallons | | \$2.46 |
| CO2 REMOVAL | | | | \$0.00 |
| | | | TOTAL CAPITAL | \$28.29 |

**PLOT AREA REQUIRED, ACRES**

| | | | | |
|---|---|---|---|---|
| | LPMEOH™ = 0.11 | Storage = 5.13 | TOTAL | 5.24 |

**INVESTMENT BY AREA**

| AREA | INVESTMENT | DESIGN | BASIS | MM-$ |
|---|---|---|---|---|
| A | REACTOR LOOP & CATALYST REDUCTION | 500 | sT/D | \$18.60 |
| B | FEED COMPRESSION | 0 | BHP | \$0.00 |
| C | RECYCLE COMPRESSION | 0 | BHP | \$0.00 |
| D | FRONT-END GAS CLEANUP | 14552 | LB-MOL/HR | \$3.00 |
| E | COMMON EQUIPMENT | 500 | sT/D | \$1.85 |
| F | SATURATOR | | | \$0.00 |
| | RESERVE | 0.0 | % | \$0.00 |
| | AIR PRODUCTS PROCESS STUDIES | | | \$0.22 |
| | AIR PRODUCTS TECHNICAL PACKAGE | | | \$0.97 |
| | LICENSE FEE | | | not incl |
| | | | Sub-total Turnkey Plant | \$24.63 |

## Table 12 OPERATING COST DATA

Case 10-C
SV 2000

**METHANOL CONVERSION ANNUAL OPERATING COST**

| | | | | | M-$/Yr | $/Gal |
|---|---|---|---|---|---|---|
| MeOH Production | 500 | sT/D | | | | |
| Annual Load Factor | 90 | % | | | | |
| Syngas (LHV) | 1434 | MMBTU/hr | $0.00 | /MMBTU | $0 | $0.000 |
| Unreacted Gas (LHV) | 995 | MMBTU/hr | $0.00 | /MMBTU | $0 | $0.000 |
| Power | | | | | | |
| Feed Compressor | 0 | kW | $0.04 | /kWh | $0 | $0.000 |
| Recycle Compressor | 0 | kW | $0.04 | /kWh | $0 | $0.000 |
| Pumps, Heaters, etc. | 382 | kW | $0.04 | /kWh | $121 | $0.002 |
| CO2 Removal | 0 | kW | $0.04 | /kWh | $0 | $0.000 |
| MP Steam, 200 psig | (49300) | lb/hr | $4.00 | /M-lb | ($1,555) | ($0.031) |
| C Water, 20oF Delta T | 400 | gpm | $0.12 | /M-gal | $23 | $0.000 |
| Misc Utilities | | | | | $240 | $0.005 |
| Catalyst, Chemicals & Lubes | | | | | $989 | $0.020 |
| Sulfur Removal, Zinc Oxide | 104 | M-lb/yr | $3.58 | /lb | $337 | $0.007 |
| COS Hydrolysis Catalyst | 6 | M-lb/yr | $3.63 | /lb | $18 | $0.000 |
| Operating Labor | | | | | $683 | $0.014 |
| Maintenance | 2% | of Investment/yr | | | $566 | $0.011 |
| | | | | Sub-Total | $1,422 | $0.028 |
| Property Taxes, Insurance | 1.5% | of Investment/yr | | | $424 | $0.008 |
| Overhead | 15% | of Oper Costs (Less Feed) | | | $213 | $0.004 |
| Capital Costs, Depr & ROI | 20% | of Investment/yr | | | $5,657 | $0.113 |
| | | Methanol Conversion Cost, Total | | | $7,717 | $0.155 |

**DISTILLATION ANNUAL OPERATING COST** FUEL GRADE - Texaco Syngas

| | | | | | M-$/Yr | $/Gal |
|---|---|---|---|---|---|---|
| MeOH Production | 500 | sT/D | | | | |
| Annual Load Factor | 90 | % | | | | |
| Electric Power | 43 | kW | $0.04 | /kWh | $13 | $0.000 |
| LP Steam, 100 psig | 3515 | lb/hr | $3.00 | /M-lb | $83 | $0.002 |
| C Water, 20oF Delta T | 252 | gpm | $0.12 | /M-gal | $14 | $0.000 |
| Operating Labor | 1 | man/yr | 50 | $-M/yr | $50 | $0.001 |
| Maintenance | 2% | of Investment/yr | | | $56 | $0.001 |
| Property Taxes, Insurance | 1.5% | of Investment/yr | | | $42 | $0.001 |
| Overhead | 15% | of Oper Costs (Less Feed) | | | $39 | $0.001 |
| Capital Costs, Depr & ROI | 20% | of Investment/yr | | | $555 | $0.011 |
| | | Purification Cost, Total | | | $852 | $0.017 |

### 3.1.4 Effect of Plant Size on Methanol Conversion Cost

In Figure 9, **Conversion Cost Vs Methanol Plant Size**, for **500 psig** coal-derived syngas was plotted. This plot assumes an unlimited amount of syngas available for methanol synthesis. The once-through process (Case 5-A) at 18% syngas conversion yields the least expensive methanol conversion cost at $0.265/gal at 300 sT/D, and $0.173/gal for a 1,100 sT/D facility. Comparable gas phase conversion costs which include shift conversion and $CO_2$ removal are $0.309 and $0.187/gal.

Figure 10, **Conversion Cost Vs Methanol Plant Size**, plots similar cases for **1,000 psig** coal-derived syngas. In all cases, methanol conversion costs are less than the comparable cases at 500 psig supply pressure, and the difference in methanol conversion cost between the LPMEOH™ Process and the gas phase process is greater at the higher supply pressure.

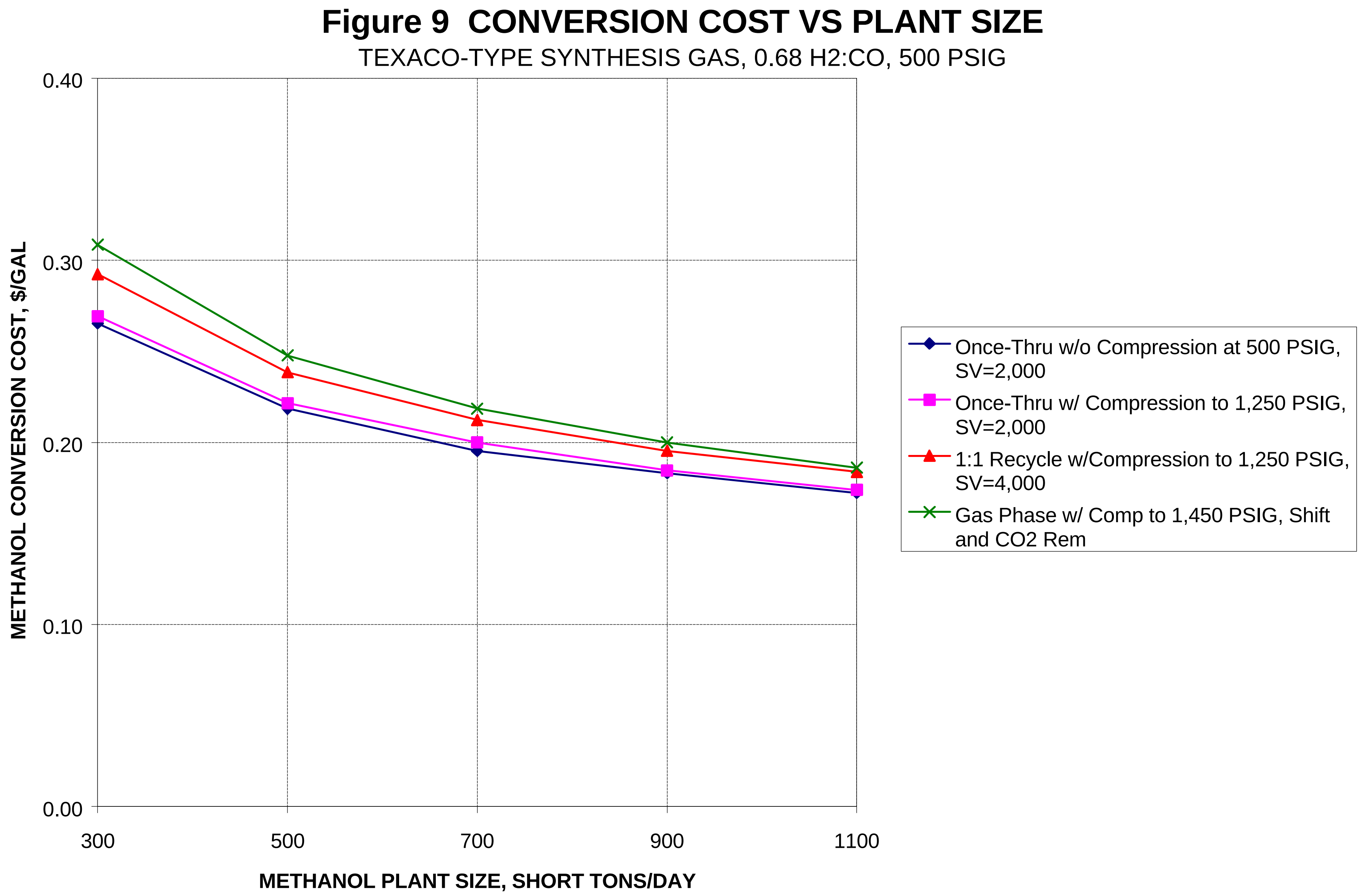
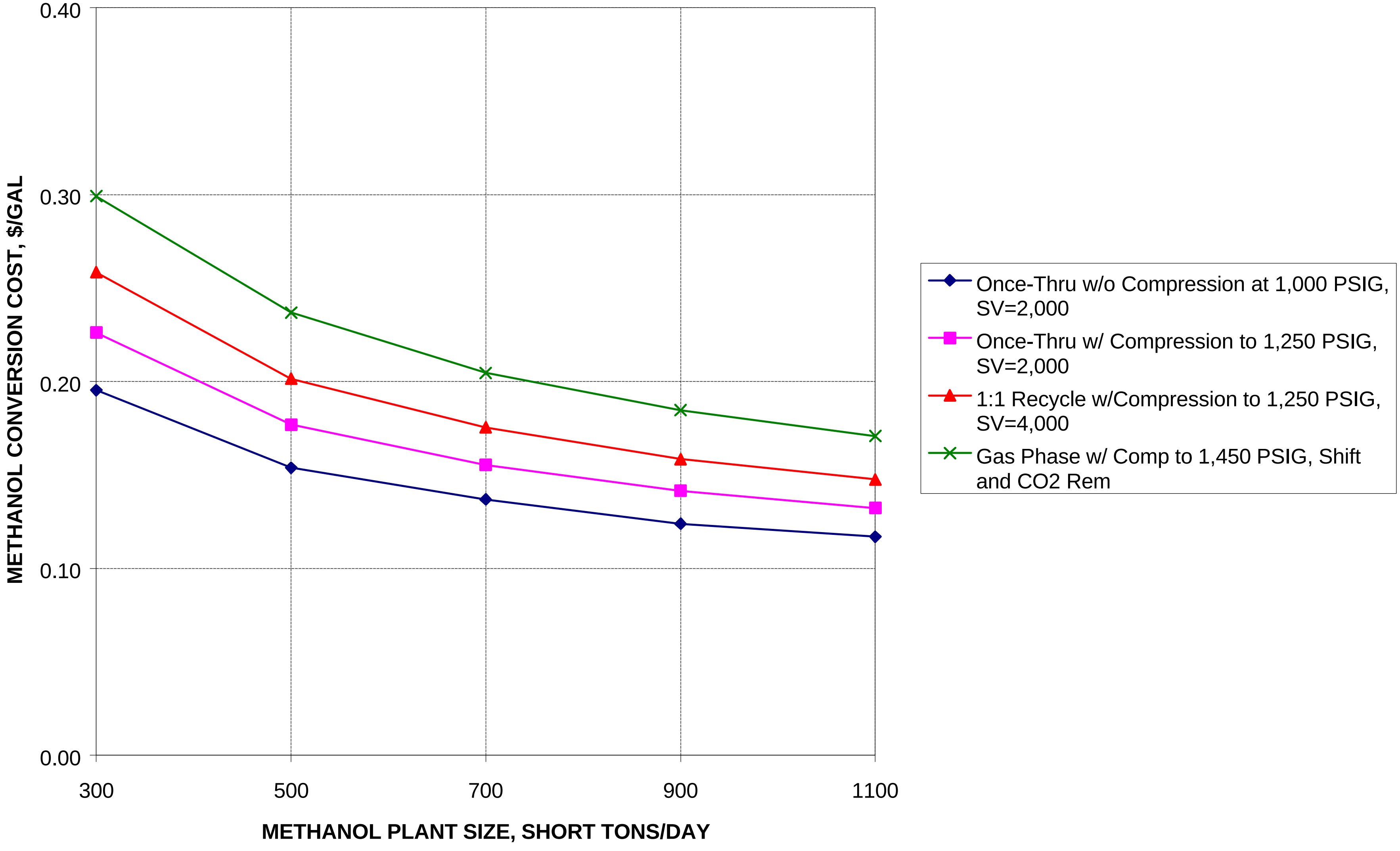


Figure 9 CONVERSION COST VS PLANT SIZE
TEXACO-TYPE SYNTHESIS GAS, 0.68 H2:CO, 500 PSIG
0.40
0.30
0.20
0.10
0.00
METHANOL CONVERSION COST, $/GAL
300
500
700
900
1100
METHANOL PLANT SIZE, SHORT TONS/DAY
Once-Thru w/o Compression at 500 PSIG, SV=2,000
Once-Thru w/ Compression to 1,250 PSIG, SV=2,000
1:1 Recycle w/Compression to 1,250 PSIG, SV=4,000
Gas Phase w/ Comp to 1,450 PSIG, Shift and CO2 Rem

# Figure 10 CONVERSION COST VS PLANT SIZE

TEXACO-TYPE SYNTHESIS GAS, 0.68:1 H2:CO, 1,000 PSIG



## 3.2 Methanol Product Purification Cost

The following is an economic analysis of the distillation cost for different product purities for balanced, Texaco-type ($H_2$/CO = 0.68), and Shell-type ($H_2$/CO = 0.50) coal-derived syngas. This analysis includes purification of methanol to 1) Fuel Grade, 2) MTBE Grade, and 3) Chemical Grade AA. The Chemical Grade includes alternatives for using either a 2 column or a 3 column design. The LPMEOH™ Process, utilizing coal-derived syngas which is CO-rich, produces a crude methanol product with nominally about 1 wt% water. Whereas, gas phase methanol synthesis results in a crude methanol product with 2-20 wt% water, depending on the amount of $CO_2$ in the syngas which is converted to methanol and water. This results in lower purification cost for the LPMEOH™ process for the Fuel Grade and MTBE Grade products. The LPMEOH™ Process makes considerably less ethanol in the crude methanol with Texaco-type syngas than either with balanced syngas or the Shell-type syngas, probably as a result of greater $CO_2$ concentrations in the feed gas. This results in considerably less cost in purifying the methanol produced from Texaco-type syngas to Chemical Grade AA than for other syngases.

### 3.2.1 Balanced Synthesis Gas

The following applies to the LPMEOH™ Process when operating on balanced syngas. It always applies to gas phase methanol as this technology runs on balanced or hydrogen-rich syngas. Figure 11, **Methanol Distillation Investment, Balanced Synthesis Gas** shows the investment cost for the different sizes and different product purity cases. Figure 12, **Methanol Distillation Cost, Balanced Synthesis Gas**, shows the operating cost of distillation for each of the cases as a function of plant size. The crude methanol produced by balanced syngas is assumed to contain 94.1 wt% methanol, 3.0 wt% water, 2.2 wt% carbon dioxide, 0.1 wt% ethanol, 0.2 wt% methyl formate, and 0.4 wt% other.

#### Fuel Grade

As defined, Fuel Grade Methanol is a methanol product containing less than 1 wt% water and is suitable for firing in a boiler or gas turbine. The capital investment for producing Fuel Grade Methanol from the balanced syngas was the lowest of the three cases. The investment is estimated at $3.4 MM for a distillation unit to produce 500 sT/D of methanol. The resulting product purification cost is $0.039/gal of methanol. The largest single operating cost component was for 100 psig steam for the reboilers. With this steam valued at $3.00/1,000 lb, the resulting steam usage cost is $0.015/gal of methanol. Note that, for this study, the balanced syngas contained 3.8 vol% $CO_2$, and produced a crude methanol that contained 3.0 wt% water. Other balanced syngas streams which contain more $CO_2$, will produce a crude methanol stream containing water, as high as 10-20 wt%. The water in the crude methanol from balanced syngas was reduced to 0.8 wt% by distillation to meet a nominal 1 wt% water content for Fuel Grade. This resulted in higher capital and steam usage cost than for the distillation of crude methanol produced directly from Texaco-type syngas by the LPMEOH™ Process, which normally contains less than 1 wt% water.

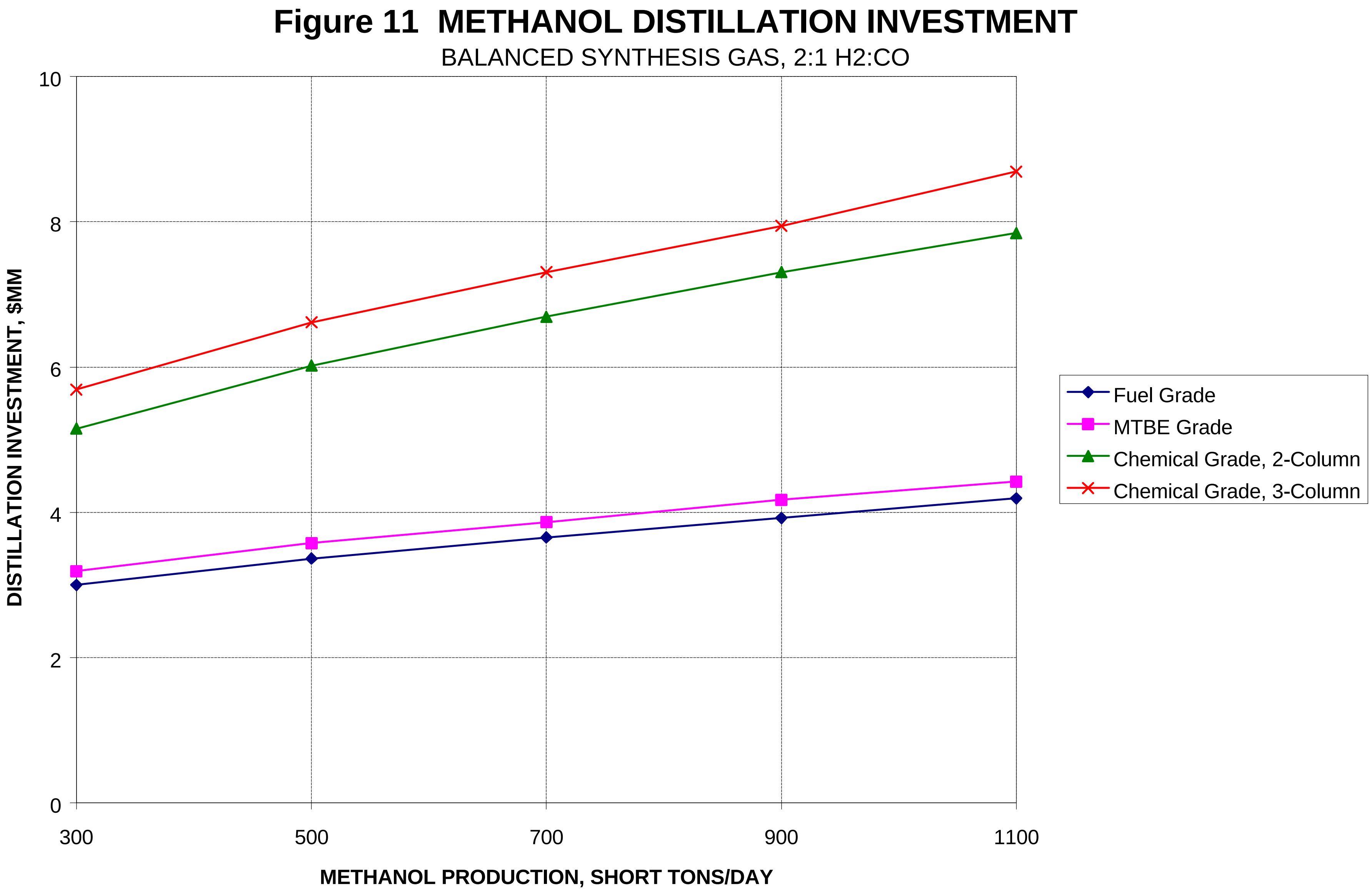
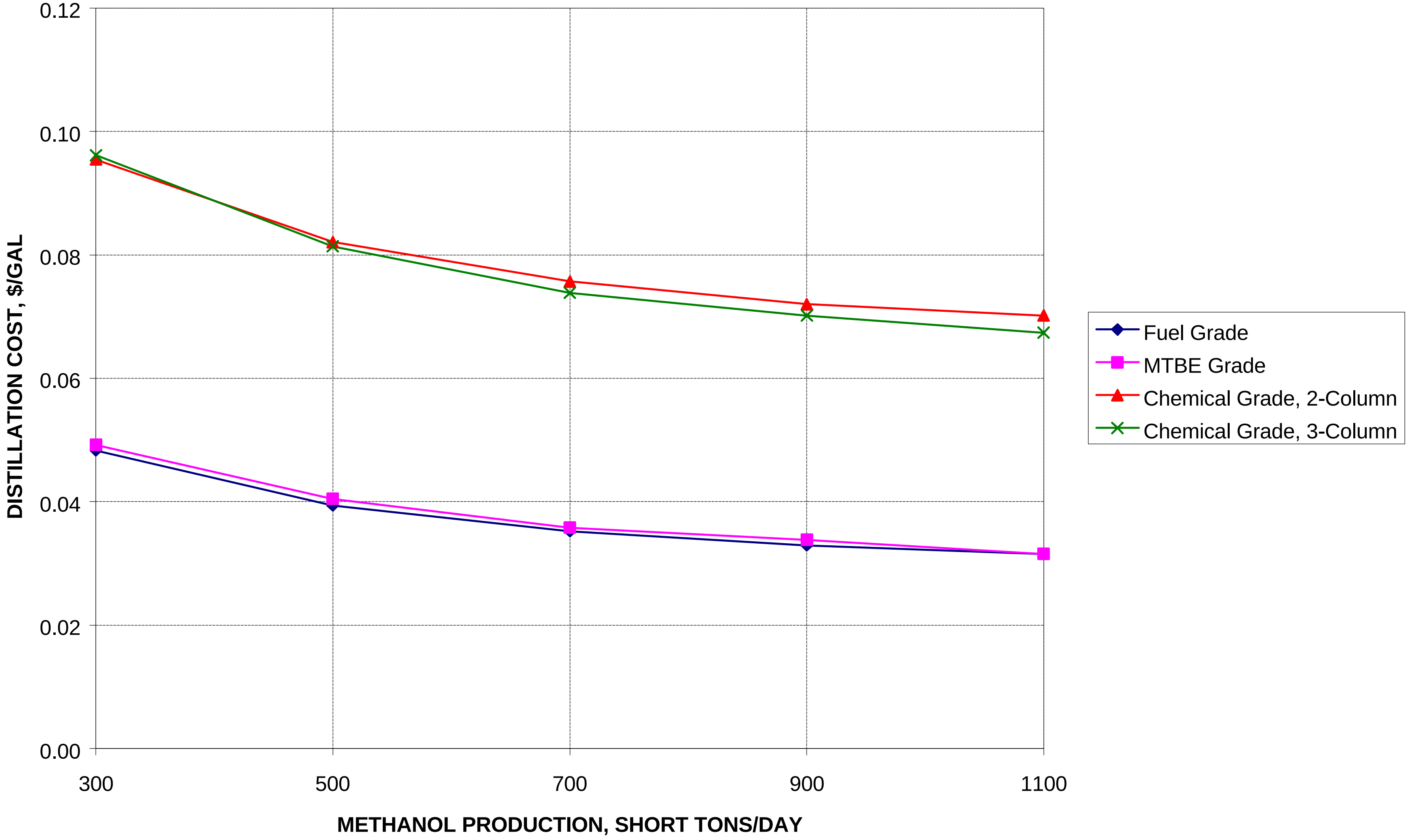


Figure 11 METHANOL DISTILLATION INVESTMENT
BALANCED SYNTHESIS GAS, 2:1 H2:CO
DISTILLATION INVESTMENT, $MM
METHANOL PRODUCTION, SHORT TONS/DAY
0
2
4
6
8
10
300
500
700
900
1100
Fuel Grade
MTBE Grade
Chemical Grade, 2-Column
Chemical Grade, 3-Column

**Figure 12 METHANOL DISTILLATION COST**

BALANCED SYNTHESIS GAS, 2:1 H2:CO



### MTBE Grade

MTBE Grade Methanol is defined as a partially refined methanol suitable for producing MTBE. The capital investment for the production of MTBE Grade Methanol from balanced syngas was only modestly higher than for Fuel Grade Methanol. The investment is estimated at $3.6 MM for a distillation unit to produce 500 sT/D of contained methanol. The resulting product purification cost is $0.040 gal of methanol. Optimization of the column height with steam usage was not done for this study. In an actual case this optimization should be done and it will result in somewhat lower cost for both grades, with the Fuel Grade being modestly lower than the MTBE Grade. For the purposes of this analysis it can be concluded that for Fuel Grade and MTBE Grade products, the investment and operating costs are almost identical at about $3.6 MM capital cost and $0.040/gal purification cost, respectively, at 500 sT/D.

### Chemical Grade AA

Chemical Grade AA methanol meets Federal Specifications for 99.85% by weight methanol. Two cases were developed for the production of Chemical Grade AA methanol from balanced syngas. These consisted of a two column design which requires the least capital investment, and a three column design which requires less steam and cooling water at the expense of increased capital investment. The investment for the two column design is estimated at $6.0 MM for a distillation unit to produce 500 sT/D of contained methanol. The investment for the three column design is estimated at $6.6 MM. With steam valued at $3.00/1,000 lb, the two cases result in about the same total operating cost of $0.081/gal. In an actual project, lower cost steam would favor the two column design. A larger size facility would favor the three column design as the scale factor for investment is less than for steam usage. For screening purposes the two designs can be considered equal.

The following table (Table 13) is a summary of the distillation cost for the four different cases.

**TABLE 13**
**DISTILLATION COST FOR METHANOL FROM BALANCED SYNTHESIS GAS**

| sT/D | Fuel Grade, $/Gal | MTBE Grade, $/Gal | Chem Grade (2 column), $/Gal | Chem Grade (3 column), $/Gal |
|---|---|---|---|---|
| 300 | 0.048 | 0.049 | 0.095 | 0.096 |
| 500 | 0.039 | 0.040 | 0.082 | 0.081 |
| 700 | 0.035 | 0.036 | 0.076 | 0.074 |
| 900 | 0.033 | 0.034 | 0.072 | 0.070 |
| 1,100 | 0.032 | 0.032 | 0.070 | 0.067 |

### 3.2.2 Texaco-Type Synthesis gas

Figure 13, **Process Flow Diagram, Methanol Distillation Fuel Grade** shows the equipment required for production of Fuel Grade Methanol produced by the LPMEOH™ Process from CO-rich syngas. Table 14, **MEOH Distillation Investment Summary** shows the total investment required. Shown on Table 3, **Operating Cost Data** is the annual operating costs for Fuel Grade methanol from Texaco-type syngas ($0.017/gal). The crude methanol produced directly from Texaco-type syngas is assumed to contain 95.4 wt% methanol, 0.6 wt% water, 3.0 wt% carbon dioxide, 0.9 wt% ethanol, 0.03 wt% methyl formate, and 0.07 wt% other. In the case of $H_2O$/steam injection to the reactor feed there is a modest increase in the water content of the crude methanol with the total reaching 1 wt% only in high addition cases.

Figure 14, **Methanol Distillation Investment, Texaco-Type Synthesis Gas** shows the investment cost for the different sizes and different product purity cases. Figure 15, **Methanol Distillation Cost, Texaco-Type Synthesis Gas**, also shows the operating cost of distillation for each of the cases as a function of plant size.

#### Fuel Grade

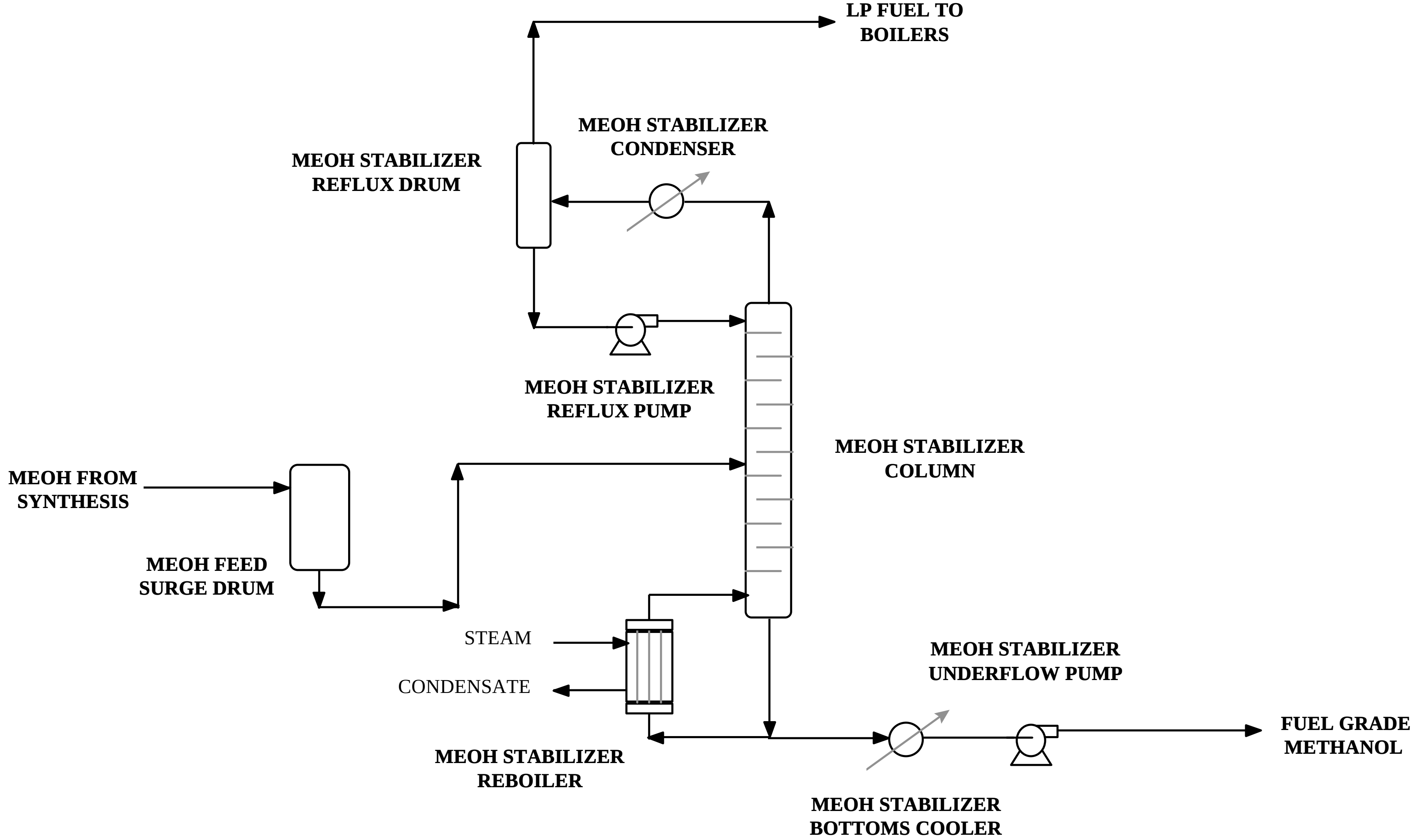
The capital investment for the Fuel Grade methanol with Texaco-type syngas was the lowest of the cases. The investment is estimated at $2.8 MM for a distillation unit to produce 500 sT/D of methanol. The resulting product purification cost is $0.017/gal of methanol. The Texaco-type syngas produces a crude methanol containing less than 1 wt% water directly from the LPMEOH™ Process; this as-produced methanol meets the specification for Fuel Grade Methanol.

#### MTBE Grade

The capital investment for the Texaco-type syngas MTBE Grade Methanol was only slightly higher than for Fuel Grade Methanol. The investment is estimated at $2.8 MM for a distillation unit to produce 500 sT/D of methanol. Since MTBE Grade purification uses more steam than Fuel Grade the resulting product purification cost is higher at $0.021/gal of methanol.

#### Chemical Grade AA

Two cases were developed for the distillation of crude methanol from Texaco-type syngas to Chemical Grade AA methanol. These consisted of a two column design which requires the least capital investment and a three column design which requires less steam and cooling water at the expense of increased capital investment. The investment for the two column design is estimated at $4.9 MM for a distillation unit to produce 500 sT/D of methanol. The investment for the three column design is estimated at $5.8 MM. With steam valued at $3.00/M-lb the two cases result in about the same total operating cost of $0.066/gal at the 500 sT/D scale. In an actual project, lower cost steam would favor the two column design while larger size facility would favor the three column design. For screening purposes the two designs can be considered equal.




AIR PRODUCTS

**Figure 13 PROCESS FLOW DIAGRAM**

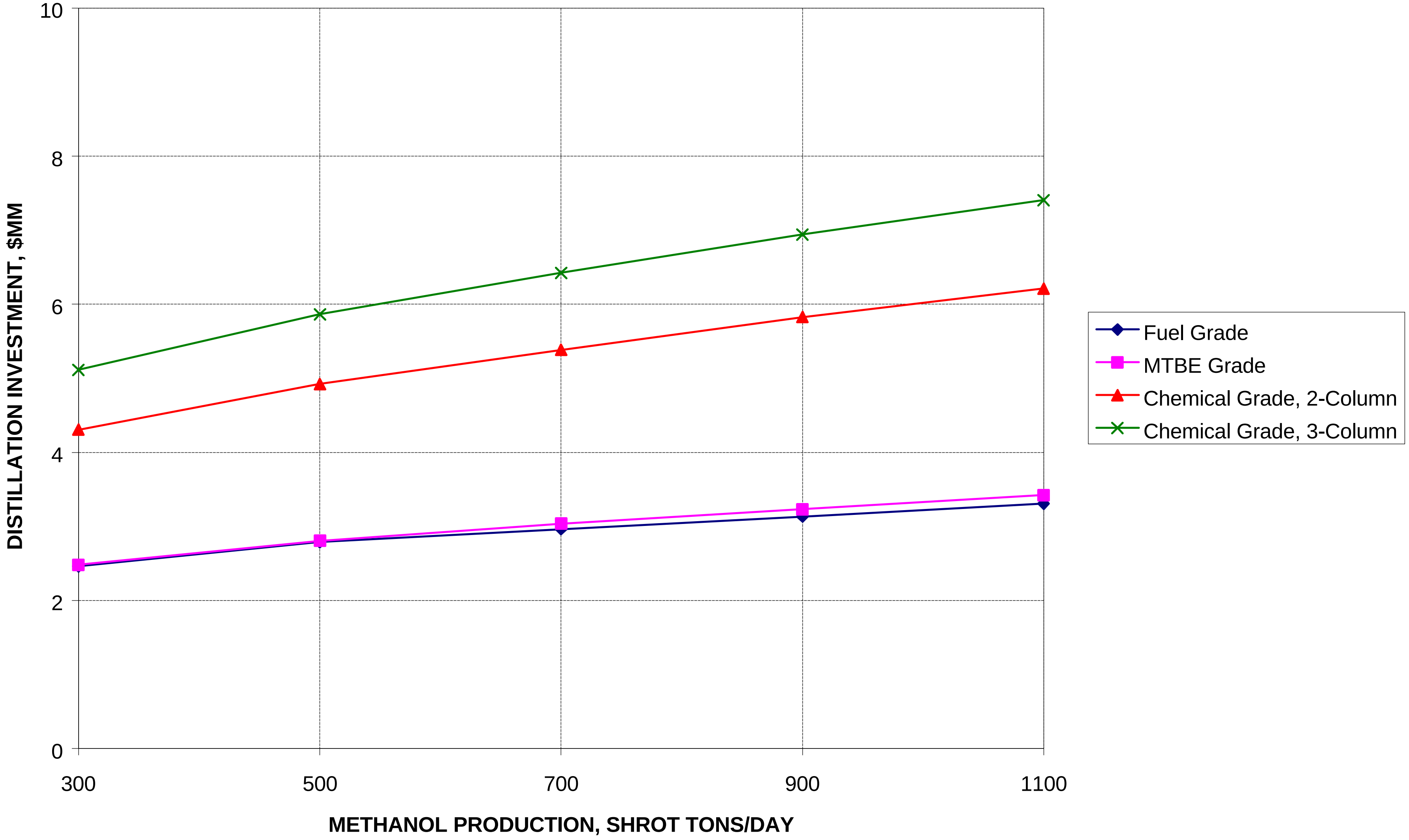
**METHANOL DISTILLATION**
**FUEL GRADE**

## Table 14 METHANOL DISTILLATION INVESTMENT

| | | | |
|---|---|---|---|
| FUEL GRADE - Texaco Syngas | | Production, sT/D | 500 |
| **INVESTMENT BREAKDOWN** | | | **M-$** |
| EQUIPMENT | | | $279 |
| EQUIPMENT SETTING | | | $6 |
| PIPING | | | $241 |
| CIVIL | | | $124 |
| STEEL | | | $20 |
| INSTRUMENTATION | | | $247 |
| ELECTRICAL | | | $128 |
| INSULATION | | | $128 |
| PAINT | | | $7 |
| OTHER | | | $1,302 |
| | | Sub-total Direct | $2,482 |
| G&A, OVERHEAD & FEES | | | $224 |
| CONTINGENCY | | | $0 |
| | | Sub-total Turnkey Plant | $2,706 |
| OWNER'S COST | 25% of Equipment | | $70 |
| | | TOTAL | $2,776 |

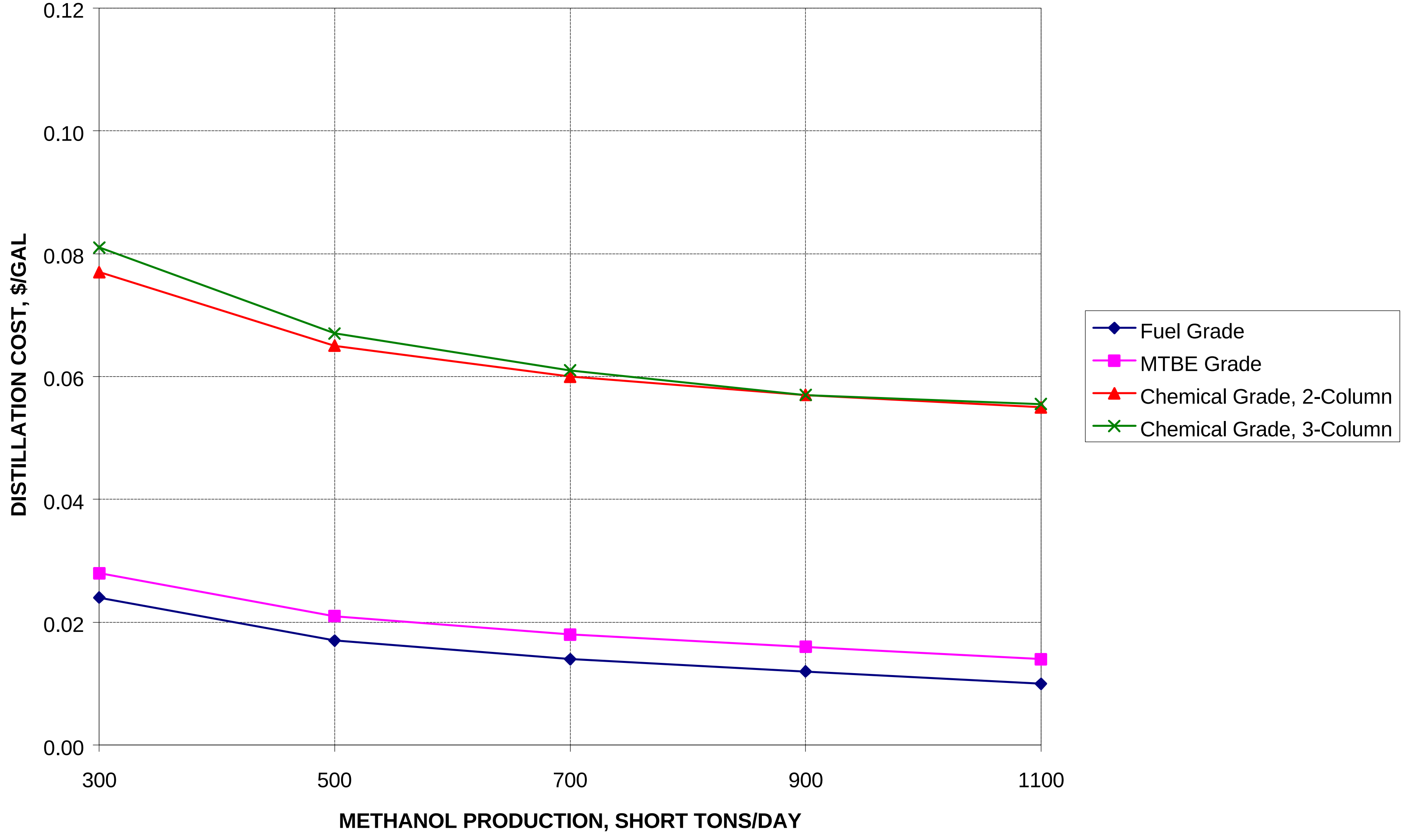
# Figure 14  METHANOL DISTILLATION INVESTMENT

TEXACO-TYPE SYNTHESIS GAS, 0.68:1 H2:CO



## Figure 15 METHANOL DISTILLATION COST

TEXACO-TYPE SYNTHESIS GAS, 0.68:1 H2:CO



The following table (Table 15) is a summary of the LPMEOH™ process distillation cost for the four different cases.

**TABLE 15**
**DISTILLATION COST FOR METHANOL FROM TEXACO-TYPE SYNTHESIS GAS**

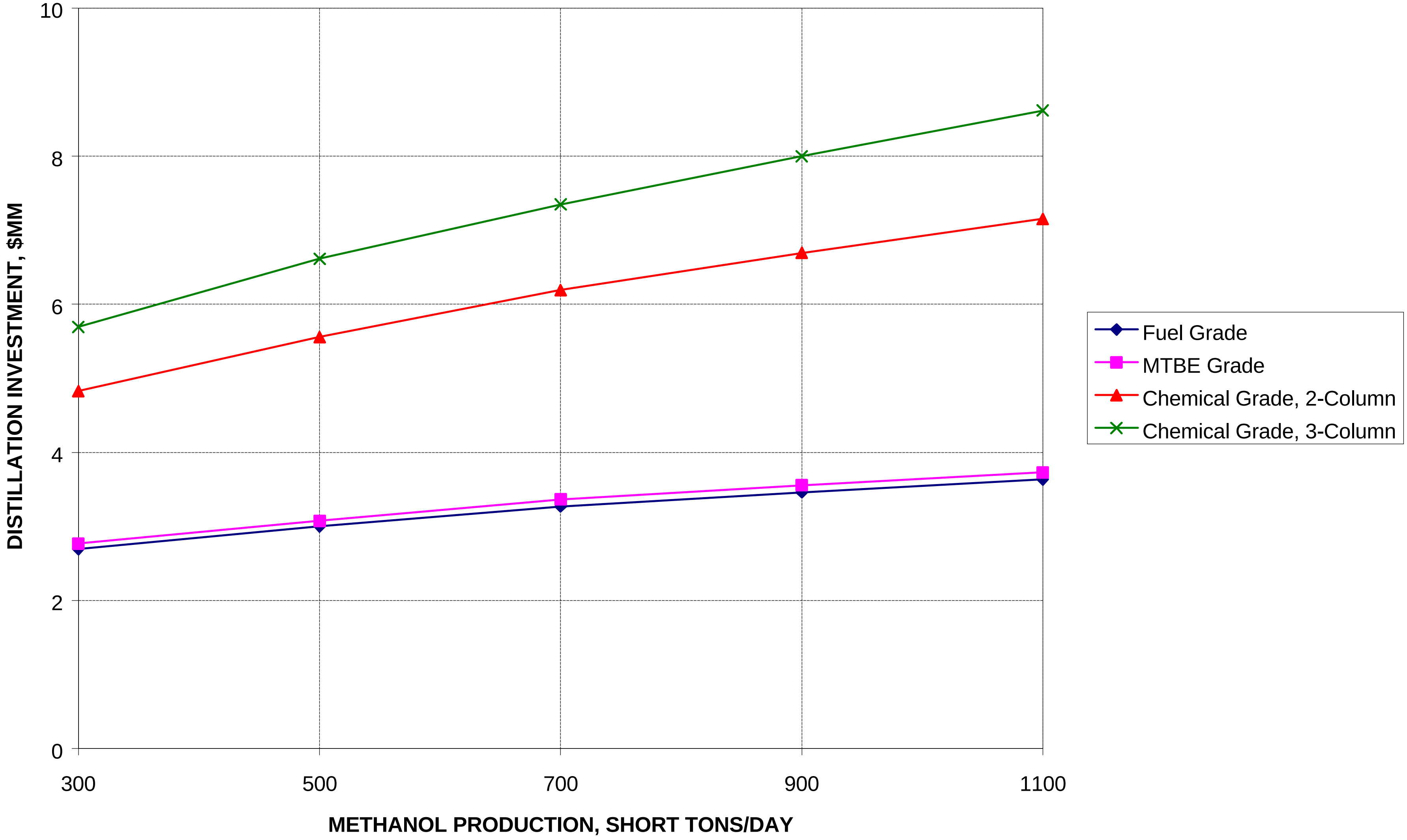
| sT/D | Fuel Grade, $/Gal | MTBE Grade, $/Gal | Chem Grade (2 column), $/Gal | Chem Grade (3 column), $/Gal |
|---|---|---|---|---|
| 300 | $0.024 | $0.028 | $0.077 | $0.081 |
| 500 | $0.017 | $0.021 | $0.065 | $0.067 |
| 700 | $0.014 | $0.018 | $0.060 | $0.061 |
| 900 | $0.012 | $0.016 | $0.057 | $0.057 |
| 1,100 | $0.010 | $0.014 | $0.055 | $0.055 |

### 3.2.3 Shell-Type Synthesis Gas

Distillation costs for Shell-type syngas are shown in Figure 16, **Methanol Distillation Investment, Shell-Type Synthesis Gas** and Figure 17, **Methanol Distillation Cost, Shell-Type Synthesis Gas**. Note that the operating cost for the production of Chemical Grade AA Methanol using a two column design crosses that for a three column design at about 550 sT/D. This shows that at larger sizes the reduced steam usage for three column distillation outweighs the increased investment required for the three column design. Since the $H_2$/CO ratio and $CO_2$ content is different for Shell-type syngas than with Texaco-type syngas the crude methanol product will be a different composition. The crude methanol produced directly from Shell-type syngas is assumed to contain 94.5 wt% methanol, 0.7 wt% water, 2.3 wt% carbon dioxide, 0.6 wt% ethanol, 1.0 wt% methyl formate, and 0.9 wt% other.

# Figure 16 METHANOL DISTILLATION INVESTMENT

SHELL-TYPE SYNTHESIS GAS, 0.53:1 H2:CO



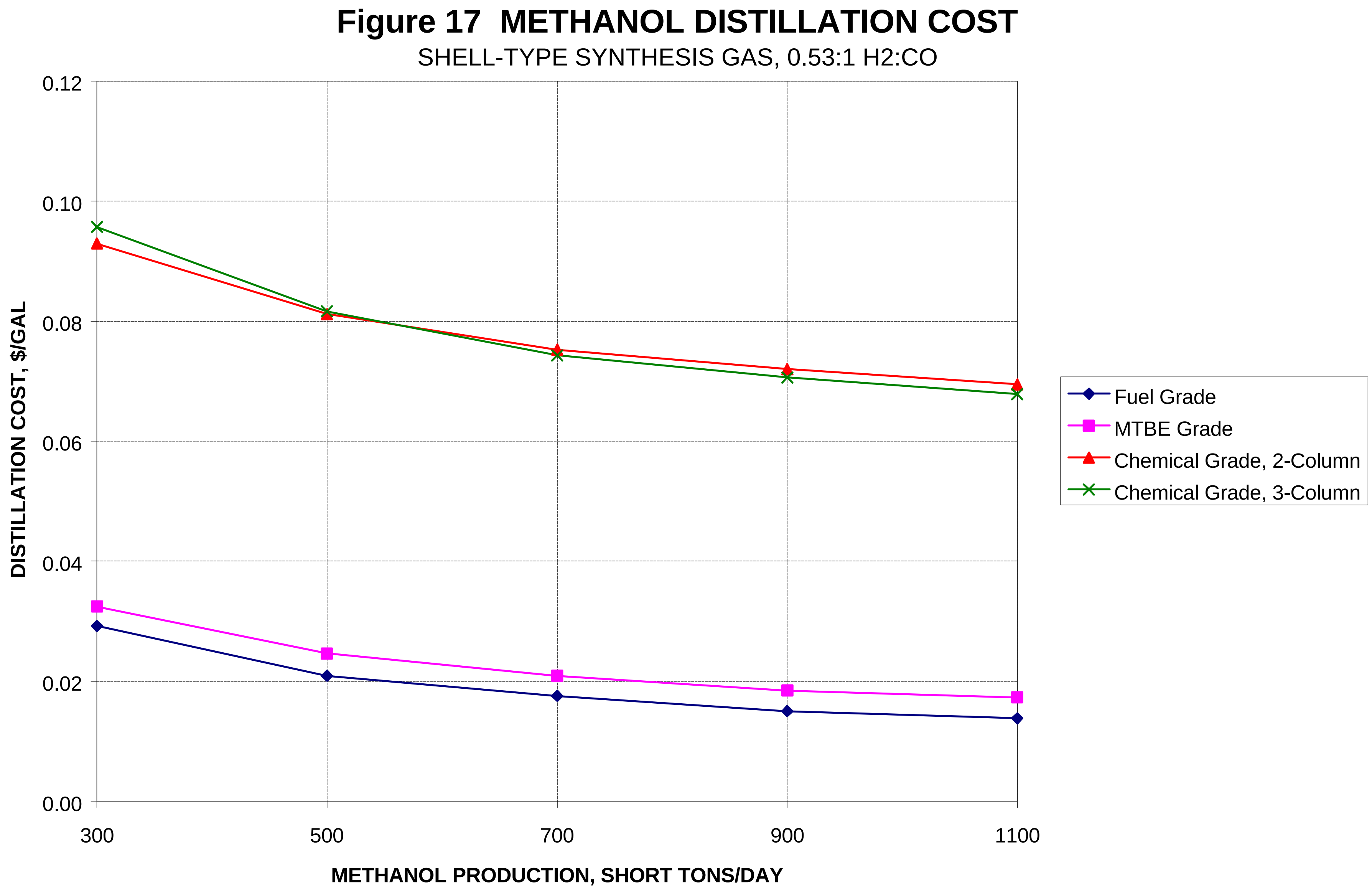


Figure 17  METHANOL DISTILLATION COST
SHELL-TYPE SYNTHESIS GAS, 0.53:1 H2:CO
DISTILLATION COST, $/GAL
0.12
0.10
0.08
0.06
0.04
0.02
0.00
300
500
700
900
1100
METHANOL PRODUCTION, SHORT TONS/DAY
Fuel Grade
MTBE Grade
Chemical Grade, 2-Column
Chemical Grade, 3-Column

# 4. <u>Sensitivity Studies</u>

## 4.1 <u>Syngas Composition Variations - Impact on Liquid Phase and Gas Phase</u>

### 4.1.1 Sulfur Content

For the cases considered in Section 3, the Fresh Feed was assumed to contain 5 ppmv of carbonyl sulfide (COS). To remove this catalyst poison the COS is first hydrolyzed over a metal oxide catalyst and then the hydrogen sulfide ($H_2S$) produced is removed with a zinc oxide bed. COS was selected to represent sulfur species in the syngas instead of $H_2S$ because of the higher capital cost associated with the hydrolysis step (operating costs for absorption of $H_2S$ and COS after hydrolysis are the same on a sulfur weight basis). Figure 18, **Effect of COS Content on Conversion Cost**, shows the effect of an increase to 20 ppmv of COS in the syngas feed. There is an increase of $0.05 to $0.08/gal in the methanol conversion cost at lower syngas conversion, as a large amount of gas must be treated. As the amount of conversion is increased (independent of the methanol conversion technology), the effect on cost becomes more modest.

### 4.1.2 Inert Content

For the cases considered in Section 3, the Fresh Feed was assumed to contain 1 vol% inerts. Increased levels of inerts can result from such parameters as the oxygen purity from the air separation unit feeding the coal gasifier. The greater the syngas conversion, the greater becomes the cost penalty of inert build up in the synthesis loop. Figure 19, **Effect of Inerts on Conversion Cost**, shows the effect of an increase to 10% inerts in the feed syngas. For the once-through case (Case 5-A) there is a $0.044/gal increase in cost along with a small reduction in conversion. With a 1:1 recycle at 1,250 psig (Case 5-H), the cost increase becomes less, $0.015/gal. For the gas phase the effect is much greater. In the case shown the inerts in the synthesis loop were limited to 35% to limit the impact on conversion due to lower partial pressure of reactants; this increases the methanol conversion cost by $0.027/gal and greatly reduced the syngas conversion to 77%.

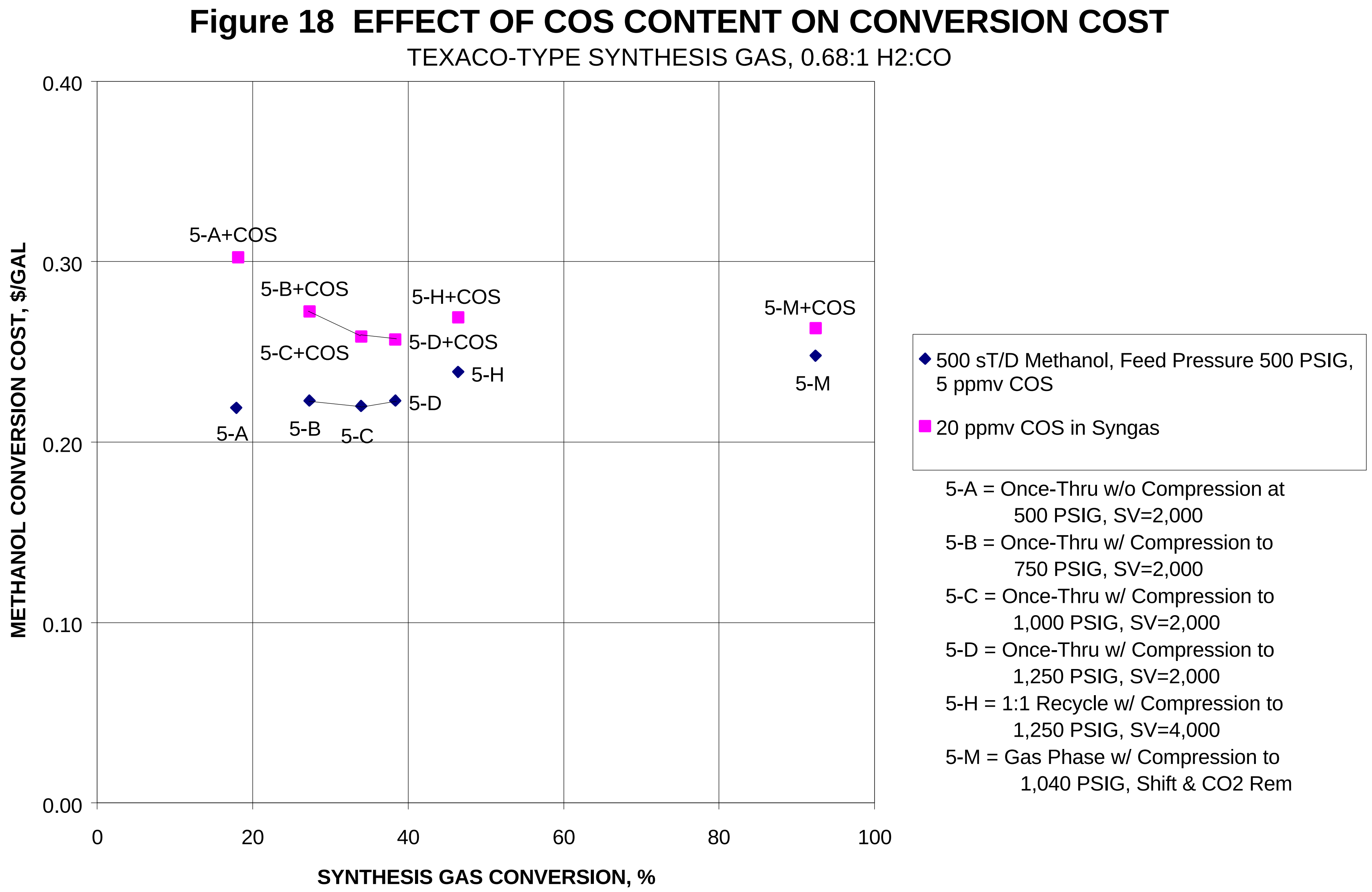
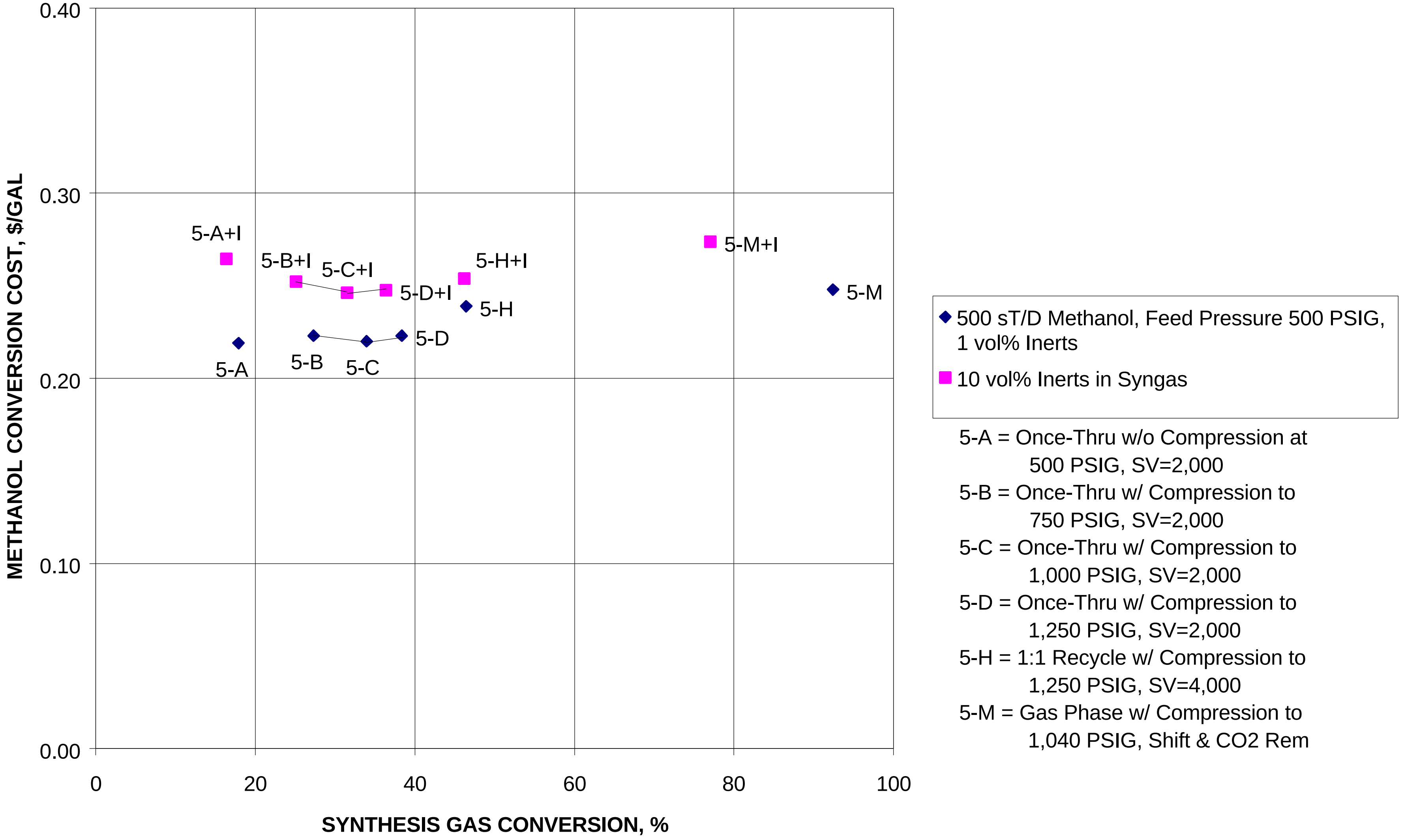


Figure 18 EFFECT OF COS CONTENT ON CONVERSION COST
TEXACO-TYPE SYNTHESIS GAS, 0.68:1 H2:CO
METHANOL CONVERSION COST, $/GAL
0.40
0.30
0.20
0.10
0.00
0
20
40
60
80
100
SYNTHESIS GAS CONVERSION, %
5-A+COS
5-B+COS
5-C+COS
5-D+COS
5-H+COS
5-M+COS
5-A
5-B
5-C
5-D
5-H
5-M
500 sT/D Methanol, Feed Pressure 500 PSIG, 5 ppmv COS
20 ppmv COS in Syngas
5-A = Once-Thru w/o Compression at 500 PSIG, SV=2,000
5-B = Once-Thru w/ Compression to 750 PSIG, SV=2,000
5-C = Once-Thru w/ Compression to 1,000 PSIG, SV=2,000
5-D = Once-Thru w/ Compression to 1,250 PSIG, SV=2,000
5-H = 1:1 Recycle w/ Compression to 1,250 PSIG, SV=4,000
5-M = Gas Phase w/ Compression to 1,040 PSIG, Shift & CO2 Rem

# Figure 19 EFFECT OF INERTS ON CONVERSION COST

TEXACO-TYPE SYNTHESIS GAS, 0.68:1 H2:CO



### 4.1.3 Effect of Capacity Factor

The depreciation and return on investment have a significant impact on the methanol conversion cost. These are magnified by the capacity factor, as this term is inversely related to the unit cost of methanol ($/gallon). All of the figures and tables presented to this point in this Report have used a capacity factor of 90%. The following table (Table 16) summarizes the change in methanol conversion cost when the capacity factor is lowered to 70%. This is more representative of the coproduction of methanol with electric power in an IGCC load-following application.

**TABLE 16**
**EFFECT OF CAPACITY FACTOR - 500 PSIG FEED GAS PRESSURE**

| | | **90% Capacity Factor** | | **70% Capacity Factor** | |
|---|---|---|---|---|---|
| Case | Variable Cost, $/Gal | Fixed Cost, $/Gal | Conversion Cost, $/Gal | Fixed Cost, $/Gal | Conversion Cost, $/Gal |
| 5-A | 0.014 | 0.205 | **0.219** | 0.263 | **0.278** |
| 5-AW | 0.013 | 0.210 | **0.223** | 0.270 | **0.283** |
| 5-B | 0.038 | 0.185 | **0.223** | 0.237 | **0.276** |
| 5-C | 0.045 | 0.175 | **0.220** | 0.224 | **0.270** |
| 5-D | 0.050 | 0.173 | **0.223** | 0.223 | **0.272** |
| 5-DW | 0.045 | 0.171 | **0.216** | 0.220 | **0.265** |
| 5-H | 0.042 | 0.197 | **0.239** | 0.253 | **0.295** |
| 5-HW | 0.038 | 0.192 | **0.230** | 0.247 | **0.285** |
| 5-HC5 | 0.052 | 0.242 | **0.294** | 0.311 | **0.363** |
| 5-HC15 | 0.037 | 0.210 | **0.247** | 0.270 | **0.307** |
| 5-M | 0.044 | 0.204 | **0.248** | 0.262 | **0.306** |

The effect of lowering the capacity factor from 90% to 70% for a Feed Gas pressure of 500 psig is to increase the methanol conversion cost by $0.05 to $0.06/gal. The advantage for the LPMEOH™ Process over the gas phase process remains at approximately $0.03/gal.

A similar table for applications with a Feed Gas pressure of 1,000 psig is provided below (Table 17). The methanol conversion cost increases by $0.04 to $0.05/gal when the capacity factor is lowered from 90% to 70%. The LPMEOH™ Process maintains a $0.07 to $0.08/gal advantage over the gas phase process at the lower capacity factor.

**TABLE 17**
**EFFECT OF CAPACITY FACTOR - 1,000 PSIG FEED GAS PRESSURE**

| | | **90% Capacity Factor** | | **70% Capacity Factor** | |
|---|---|---|---|---|---|
| Case | Variable Cost, $/Gal | Fixed Cost, $/Gal | Conversion Cost, $/Gal | Fixed Cost, $/Gal | Conversion Cost, $/Gal |
| 10-C | 0.004 | 0.151 | **0.155** | 0.194 | **0.198** |
| 10-D | 0.014 | 0.163 | **0.177** | 0.209 | **0.223** |
| 10-DW | 0.012 | 0.161 | **0.173** | 0.207 | **0.219** |
| 10-H | 0.013 | 0.188 | **0.201** | 0.242 | **0.255** |
| 10-HW | 0.012 | 0.184 | **0.196** | 0.237 | **0.249** |
| 10-HC5 | 0.028 | 0.234 | **0.262** | 0.301 | **0.329** |
| 10-HC15 | 0.017 | 0.207 | **0.224** | 0.266 | **0.283** |
| 10-M | 0.031 | 0.196 | **0.227** | 0.252 | **0.283** |

## 4.2 Syngas Usage - Impact on IGCC Power Plant

### 4.2.1 Impact of Process on Syngas Usage

Syngas usage is defined as the lower heating value (LHV) energy content of the fresh syngas feed, Stream (1) Figure 1, minus the lower heating value (LHV) energy content of the unreacted gas, Stream (3) Figure 1, Fuel to the IGCC power plant gas turbine. The usage changes as a result of process design considerations, such as feed compression, water addition, recycle, and $CO_2$ removal. The following (Table 18) is a table of the syngas usage per gallon of methanol for each of the cases at a Feed Gas pressure of 500 psig:

**TABLE 18**
**EFFECT OF ON SYNGAS USAGE - 500 PSIG FEED GAS PRESSURE**

| Case | Flow Scheme | Reactor Pressure, psig | Syngas Usage, Btu/Gal |
|---|---|---|---|
| 5-A | Once-Through | 500 | 71,700 |
| 5-AW | Once-Through with 5 vol% water | 500 | 75,000 |
| 5-B | Once-Through | 750 | 69,600 |
| 5-C | Once-Through | 1,000 | 69,200 |
| 5-D | Once-Through | 1,250 | 69,200 |
| 5-DW | Once-Through with 5 vol% water | 1,250 | 70,600 |
| 5-H | 1:1 Recycle | 1,250 | 69,100 |
| 5-HW | 1:1 Recycle with 5 vol% water | 1,250 | 70,100 |
| 5-HC5 | 1:1 Recycle with 5 vol% water and $CO_2$ removal | 1,250 | 70,800 |
| 5-HC15 | 1:1 Recycle with 15 vol% water and $CO_2$ removal | 1,250 | 71,800 |
| 5-M | 4.86:1 Recycle with Shift and $CO_2$ removal | 1,040 | 70,400 |

The general trend shows that the higher the pressure which is used, the higher is the production, and the lower is the syngas usage per gallon of methanol. Also, as the amount of water/steam injection is increased, the higher is the syngas usage per gallon of methanol.

### 4.2.2 Impact of Mass Flow on Power

In the LPMEOH™ cases, which do not utilize $CO_2$ removal, there is an increase in $CO_2$ content in the fuel gas to the gas turbine. This $CO_2$ serves as a temperature moderator in the combustion zone and as mass flow for power production in the expansion zone. An analysis of this effect for the 500 psig once-through case with 18% conversion of coal-derived syngas (Case 5-A) , results in an increase the capacity of the gas turbine by about 0.5%. The gas turbine heat rate was slightly poorer; however there was 0.8% more turbine mass flow exhaust available at 1100$^o$F for heat recovery for a steam turbine.

For the 1,250 psig, 1:1 recycle case with 5 vol% water addition and 52% syngas conversion (Case 5-HW), the increase in $CO_2$ resulted in an increased gas turbine capacity of 6.7% with a slight improvement in gas turbine heat rate. Also there was 6.9% more turbine mass flow exhaust available at 1100$^o$F for heat recovery for a steam turbine.

A comprehensive analysis of the effect of change in fuel composition on an IGCC facility is too complex for inclusion in this Report other than as shown above. It is expected to be a positive factor and should be analyzed for the specific case. Gas phase methanol production does not significantly change the fuel composition going to the gas turbine, thus there will be no beneficial effect.

## 5. Examples of Cost of Methanol in IGCC Applications

The objective of optimizing the methanol conversion cost, the cost of product distillation, and the syngas usage is to compute the cost of production of methanol. The following is a comparison of a simple 500 sT/D LPMEOH™ Facility which converts 34% of a syngas supply using once-through production (Case 10-C), with a more complex 500 sT/D gas phase methanol facility which shifts a portion of the incoming syngas, removes the $CO_2$ produced and blends with additional syngas to produce the balanced syngas needed for a gas phase reactor (94% conversion). The results of this analysis are provided in Table 19. In the case of baseload coproduction it is assumed that the syngas will normally be available 90% of the time and that the methanol product must bear part of the investment cost for the syngas. In this case syngas was charged at $4.50/MMBTU. Case 10-C has higher capital investment for the methanol synthesis loop and sulfur removal equipment, but lower overall capital cost due to the additional shift and $CO_2$ removal equipment in the gas phase process. The methanol savings are on the order of $0.109/gal when using the LPMEOH™ Process in this application. Tie-in costs are not included, but would be expected to be on the order of $0.01/gal for either methanol conversion technology.

In a load-following situation, syngas from the IGCC power plant would be available for methanol production only at off-peak periods; for this example, the availability of the syngas is assumed to be 70% of the time. On an as-available basis the methanol product would bear only the variable cost of syngas production, or $3.80/MMBTU in this case. The net effect is to increase the advantage for the LPMEOH™ Process to $0.123/gal.

## TABLE 19
## LPMEOH™ VS GAS PHASE METHANOL PRODUCTION

| **Case** | **LPMEOH™ Case 10-C** | **Gas Phase** |
|---|---|---|
| Methanol Plant Size: | 500 sT/D | 500 sT/D |
| Conversion: | 34% | 94% |
| Methanol Product: | Fuel Grade | Fuel Grade |
| CO-rich Syngas: | 1,000 psig | 1,000 psig |
| | 5 ppmv COS | 5 ppmv COS |
| **Methanol Technology** | **LPMEOH™** | **Gas Phase** |
| Syngas Usage, BTU/Gal | 69,200 | 70,400 |
| **Investment, millions of $** | | |
| Sulfur Removal | $3.00 | $0.42 |
| Shift Conversion | $0 | $3.91 |
| $CO_2$ Removal | $0 | $7.83 |
| Syngas Compression | $0 | $5.63 |
| Methanol Loop | $19.55 | $16.78 |
| Catalyst Reduction | $2.08 | $0 |
| Owner's Cost | $1.20 | $1.73 |
| Methanol Storage | $2.46 | $2.46 |
| Total Investment | $28.29 | $38.76 |
| **Base Load** | 90% On-Stream MEOH, $/Gal | 90% On-Stream MEOH, $/Gal |
| Conversion Cost | $0.155 | $0.236 |
| Distillation Cost | $0.017 | $0.039 |
| Syngas Cost @ $4.50/MMBTU | $0.311 | $0.317 |
| Baseload Methanol Cost | $0.483 | $0.592 |
| **LPMEOH™ Advantage** | **$0.109** | |
| **Load Following** | 70% On-Stream MEOH, $/Gal | 70% On-Stream MEOH, $/Gal |
| Conversion Cost | $0.199 | $0.294 |
| Distillation Cost | $0.021 | $0.044 |
| Syngas Cost @ $3.80/MMBTU | $0.263 | $0.268 |
| Baseload Methanol Cost | $0.483 | $0.606 |
| **LPMEOH™ Advantage** | **$0.123** | |

## 6. Conclusions - Summary of Cost Advantages (LPMEOH™ Vs Gas Phase)

This Topical Report compares the cost of methanol as produced from the LPMEOH™ Process and from a conventional gas phase process as applied to a generic 500 sT/D methanol plant as part of an IGCC coproduction facility. The cost of methanol is calculated as the sum of three terms: the methanol conversion cost (which includes the fixed and operating costs for the methanol unit), the distillation cost, and the syngas cost from the IGCC facility. A proprietary cost estimation screening program developed by R. B. Moore of Air Products was used to calculate the methanol conversion cost and the distillation cost from the LPMEOH™ Process and the gas phase process for various syngas supply pressures and on-stream factors. The methanol conversion cost from the LPMEOH™ Process is $0.02 to $0.07 per gallon lower than from the gas phase methanol process.

A major component of the methanol conversion cost in an IGCC complex is the cost to distill the crude methanol product in order to meet the final specification. It is typical for methanol to be stabilized (either by distillation or by deep flashing) to remove volatile components (such as $CO_2$) and permit shipment and transport in atmospheric vessels. Beyond stabilization, other distillation may be necessary so that the final methanol product meets the specification for the designated end-use. Based upon the current applications of methanol, there are three grades of methanol product (Chemical Grade AA, Fuel Grade, and MTBE (methyl tertiary-butyl ether) Grade) which could be used in downstream chemical or power applications. These grades of methanol differ in the amounts of water and higher alcohols which are present in the final product. In particular, the Fuel Grade and MTBE Grade products have a water specification of 1 wt%, while the Chemical Grade AA methanol has a maximum water content of 0.1 wt%.

The LPMEOH™ Process, which can directly process coal-derived syngas which is rich in CO, produces a crude methanol product with nominally about 1 wt% water. Whereas, gas phase methanol synthesis results in a crude methanol product with 2-20 wt% water, depending on the amount of $CO_2$ in the syngas which is converted to methanol and water. This results in lower purification cost for the LPMEOH™ process for the Fuel Grade and MTBE Grade products. By applying the same cost estimation screening program, the distillation cost to produce Fuel Grade methanol from the LPMEOH™ Process which directly utilizes CO-rich syngas is about $0.02 per gallon less than from the gas phase methanol process.

Sensitivity studies performed as part of this Report indicate that the magnitude of the advantage in the methanol conversion cost for the LPMEOH™ Process when compared with the conventional gas phase process is increased when:

a) the syngas is rich in CO,
b) syngas is available at higher pressures,
c) only modest syngas conversion to methanol is required,
d) syngas is available with low $H_2S$ and COS content,
e) inerts in the syngas (such as nitrogen in the oxygen from the air separation unit feeding the gasifier) are relatively high, and
f) Fuel Grade or MTBE Grade (low water) Methanol is required.

The following table (Table 20) summarizes several operating scenarios for a baseload 500 sT/D LPMEOH™ Facility, when compared with the more complex 500 sT/D gas phase methanol facility. As in the prior discussion, the baseload coproduction assumes that the syngas is normally be available 90% of the time and syngas is charged at \$4.50/MMBTU. Cost advantages of 4 to 11 cents per gallon of methanol can be realized by utilizing the LPMEOH™ Process.

**TABLE 20**
**Baseload Coproduction, 500 sT/D, Fuel Grade Methanol**
**Texaco-Type Syngas @ \$4.50/MMBTU**

| Syngas Feed Pressure, psig | 500 | 500 | 1,000 | 1,000 |
|---|---|---|---|---|
| Design Case | **5-C** | **5-H** | **10-C** | **10-H** |
| Flow Scheme | Once-Through | 1:1 Recycle | Once-Through | 1:1 Recycle |
| Syngas Conversion, % | 34 | 46 | 34 | 46 |
| Methanol Cost, \$/Gal | 0.548 | 0.567 | 0.483 | 0.529 |
| LPMEOH™ Advantage over gas phase, \$/Gal | 0.055 | 0.036 | 0.109 | 0.063 |

## 7. Future Improvements and Recommendations

The current design of the LPMEOH™ Demonstration Unit at Kingsport is based upon conservative performance calculations and also on an extensive amount of extra equipment and instrumentation required for evaluation of a new technology. Any increase in the rated output from the LPMEOH™ Demonstration Unit as a result of operating performance will lead to a reduction in the impact of capital charges on the unit price of methanol. For example, a 10% increase in the basis for determining the rated production capacity would reduce the methanol conversion cost by 1 to 2 cents per gallon of methanol at the 500 sT/D plant size.

Actual operation and subsequent evaluation is also expected to lead to a number of improvements in future designs which will result in lower operating costs. Some of the expected improvements are as follows:

1. Greater than the current design basis for catalyst life and number of reactor stages. The design is based on an average catalyst activity ($\eta = 0.5$) after six months operation and a single stirred reactor stage (CSTRs = 1). Increasing either of these parameters improves the per-pass conversion of syngas to methanol.
2. Reduced capital requirements by elimination of instrumentation and analytical requirements and general learning on unneeded facilities from the LPMEOH™ Demonstration Unit .
3. Demonstration of a higher than design 90% availability factor.

4. Elimination of on-stream catalyst removal and addition facilities, particularly for load-following operation, where ample time is available to change catalyst off-line.

The potential results of such improvements are shown in the following table.

**TABLE 21**
**POTENTIAL IMPROVEMENTS BASED ON DEMONSTRATION TESTS**

| **Case** | Base | 1 | 2 | 3 | 4 |
|---|---|---|---|---|---|
| Plant Size, sT/D | 500 | 500 | 500 | 500 | 500 |
| On-Stream Factor, % | 90 | 90 | 90 | 98 | 90 |
| Catalyst Age ($\eta$) | 0.5 | 0.7 | 0.5 | 0.5 | 0.5 |
| Number of CSTRs | 1 | 2 | 1 | 1 | 1 |
| Capital Required, % of Base | 100 | 96 | 80 | 100 | 92 |
| On-Line Catalyst Replacement | Yes | Yes | Yes | Yes | No |
| **Methanol Conversion Cost, $/Gal** | **0.155** | **0.149** | **0.128** | **0.142** | **0.144** |

Other process variations should be evaluated such as the use of staged reactors rather than recycle compression. This permits better utilization of catalyst and is advantageous when a single reactor exceeds shipping limitations. When on-off load following is required, the elimination of starting and stopping a recycle compressor will improve reliability and response time.

Based upon this analysis, the potential exists for reducing the methanol cost from the LPMEOH™ Process by an additional $0.02 - $0.05 per gallon if increases in the rated capacity of the plant and improvements from the original plant design can be demonstrated. The results from the operation of the LPMEOH™ Demonstration Unit will be used to support these changes in future plant designs.