COMMERCIAL-SCALE DEMONSTRATION OF THE LIQUID PHASE METHANOL (LPMEOH™) PROCESS

PUBLIC DESIGN REPORT
(Final Report, Volume 1: Public Design)

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DISCLAIMER

ABSTRACT

The subject of this report is a U.S. Department of Energy (DOE) sponsored demonstration under the Clean Coal Technology Program of the production of methanol from coal-derived synthesis gas. The technology being demonstrated is the Liquid Phase Methanol (LPMEOH™) process.

A 80,000 gallon/day facility was installed in Kingsport, Tennessee; this facility has completed a successful start-up and began four years of demonstration operations in April of 1997.

This Report provides the public with non-proprietary design information for the facility. The report includes a project description and history along with information on the participants and project objectives. Information on the technology is provided first in a brief form and later in detail. The report includes flow diagrams, a material and energy balance, facility layout drawings, photographs, and details of the equipment that comprise the facility.

Details of the capital cost of the facility and estimated operating costs are provided.

The final section of the report is dedicated to describing in some detail the expected commercial applications of this technology.

ACKNOWLEDGEMENTS

We would like to acknowledge the contributions made to the project by the following individuals:

Eastman Chemical Company
William Jones
Laurie Paulonis
Larry Daniels
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The following individuals have made contributions to this report as well as to the project.

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Robert Kornosky
William O'Dowd

Air Products and Chemicals, Inc.
Edward Heydorn
VanEric Stein
Robert Moore

POINT OF CONTACT

Air Products and Chemicals, Inc., is the licensor of the LPMEOH™ technology. Listed below are two individuals to contact for more information on this Demonstration project or commercial applications:

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Air Products	-	Air Products and Chemicals, Inc.
AFDU	-	Alternative Fuels Development Unit
BFW	-	Boiler Feed Water
BOD	-	Biochemical Oxygen Demand
CO	-	Carbon Monoxide
CO_2	-	Carbon Dioxide
COTR	-	Contracting Officer's Technical Representative
DME	-	Dimethyl Ether
DOE	-	United States Department of Energy
DVT	-	Design Verification Testing
EA	-	Environmental Assessment
Eastman	-	Eastman Chemical Company
EIV	-	Environmental Information Volume
EMP	-	Environmental Monitoring Plan
H_2	-	Hydrogen
H_2S	-	Hydrogen Sulfide
HHV	-	Higher Heating Value
IGCC	-	Integrated Gasification Combined Cycle
LHV	-	Lower Heating Value
LPMEOH™	-	Liquid Phase Methanol (the technology to be demonstrated)
MTBE	-	Methyl Tertiary Butyl Ether
N_2	-	Nitrogen
NEPA	-	National Environmental Policy Act
NH_3	-	Ammonia
NO_x	-	Nitrogen Oxides
Partnership	-	Air Products Liquid Phase Conversion Company, L.P.
PON	-	Program Opportunity Notice
Project	-	Production of Methanol/DME Using the LPMEOH™ Process at an Integrated Coal Gasification Facility
P&ID(s)	-	Piping and Instrumentation Diagram(s)
SBC	-	Standard Building Code
Sl/hr-kg	-	Standard Liter(s) per Hour per Kilogram of Catalyst (oxide basis)
SO_2	-	Sulfur Dioxide
Tie-in(s)	-	the interconnection(s) between the LPMEOH™ Process Demonstration Facility and the Eastman Facility
TPD, sT/D or T/D	-	Short Ton(s) per Day (2,000 lbs/ton)
TSP	-	Particulate Matter
VOC	-	Volatile Organic Compounds
vol	-	Volume
WBS	-	Work Breakdown Structure
wt	-	Weight

LIST OF UNITS

Btu	-	British Thermal Units
Btu/hr	-	British Thermal Units per hour
Btu/lb	-	British Thermal Units per pound
Btu/SCF	-	British Thermal Units per standard cubic foot
Ft	-	Foot
°F	-	Degrees Fahrenheit
g	-	Grams
Gal or gal	-	Gallons
G/D	-	Gallons per day
GPH	-	Gallons per hour
gmol	-	Gram-mole
gpm	-	Gallons per minute
Hp	-	Horsepower
hr	-	Hour
KCAL	-	Kilocalories
KSCFH	-	Acronym for 1,000 standard cubic feet per hour; for this term, standard conditions are 14.7 psia and 60°F
KW	-	Kilowatt
KWH	-	Kilowatt-hour
l	-	Liter
lb	-	Pound
lbs	-	Pounds
mg	-	Milligrams
MM or mm	-	Million
MWe	-	Megawatts of electricity
ppb	-	Parts per billion
ppm	-	Parts per million
psia	-	Pounds per square inch (Absolute)
psig	-	Pounds per square inch (Gauge)
RPM	-	Revolutions per minute
SCF	-	Standard cubic feet (60°F, 14.7 psia)
SCFM	-	Standard cubic feet per minute
SCFH	-	Standard cubic feet per hour
SCFD	-	Standard cubic feet per day
x'-y"	-	x feet - y inches

GLOSSARY OF TERMS

Term:	**Definition:**
Balanced Gas	A synthesis gas with a composition of hydrogen (H_2), carbon monoxide (CO) and carbon dioxide (CO_2) in stoiciometric balance for the production of methanol (approximately 2:1 H_2/CO).
Catalyst Age	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 Conversion	Defined as the percentage of CO consumed across the reactor.
CO Gas	A synthesis gas containing primarily CO.
Crude Grade Methanol	Underflow from rectifier column, defined as 80 wt% minimum purity; requires further distillation in existing Eastman Chemical equipment prior to use.
Feed Gas (Feed)	Synthesis gas "fed" to a reactor for synthesis.
Fresh Feed	Sum of "Balanced Gas", "H_2 Gas", and "CO Gas."
Fuel Gas	Total of all purge gas streams (reactor loop, distillation, catalyst reduction) returned to Eastman Chemical's fuel gas header; energy content of this stream is used in calculation of total energy content of feed gas streams converted to methanol.
Gas Holdup	Percentage of the reactor volume up to the Gassed Slurry Level which is gas; this is an important parameter in terms of reactor hydrodynamics and required reactor volume.
Gassed Slurry Level	Level of gassed slurry; one of the major process control loops; typical units are feet.

Term:	**Definition:**
H_2 Gas	A synthesis gas with a H_2 to CO ratio greater than 2.
Inlet Superficial Velocity	Ratio of the actual cubic feet of gas at the reactor inlet (calculated at the reactor temperature and pressure) to the reactor cross-sectional area (excluding the area contribution by the internal heat exchanger); typical units are feet per second.
Methanol Productivity	Defined as the gram-moles of methanol produced per hour per kilogram catalyst (on an oxide basis).
Raw Methanol	Sum of Refined Grade and Crude Grade Methanol; represents total methanol which is produced after stabilization.
Reactor Feed	The sum of the "Fresh Feed" and "Recycle Gas."
Reactor O-T-M Conversion	Percentage of energy in the Reactor Feed converted to methanol (Once-Through-Methanol basis).
Reactor Volumetric Productivity	Defined as the quantity of methanol produced (tons per day) per cubic foot of reactor volume up to the Gassed Slurry Level.
Recycle Gas	The portion of unreacted syngas effluent from the reactor, "recycled" as a feed gas.
Recycle Ratio	The ratio of "Recycle Gas" to "Fresh Feed"
Reduction Gas	A nitrogen/carbon monoxide mixture used to activate fresh catalyst.
Refined Grade Methanol	Refined Methanol, defined as 99.8 wt% minimum purity; will be used directly in downstream Eastman Chemical processes.
Slurry Concentration Syn: (Catalyst Concentration) (Catalyst Loading)	Percentage of weight of slurry (solid + liquid) which is catalyst. Catalyst weight is defined on an oxide (unreduced) basis. Design slurry concentration is 40 wt%; target is >40 (45 wt%).

Term:	Definition:
Slurry Level (ungassed)	Level of ungassed slurry; typical units are feet.
Space Velocity	Defined as the rate of inlet gas flow to the reactor per mass of catalyst; typical units are standard liters (standard conditions are 1 atm absolute and 32°F) per hour per kilogram catalyst (on an oxide basis).
Syngas	Abbreviation for Synthesis Gas.
Syngas Conversion	The percentage of lower heating value (LHV) energy content of the Fresh Feed (Streams 10, 10, and 30) which is converted to liquid product.
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 (Syngas may also contain carbon dioxide (CO_2), water and other gases).
Syngas Usage	The syngas use to produce the methanol product, defined as the amount of lower heating value (LHV) energy content of the Fresh Feed (Streams 10, 20, and 30), less the (LHV) energy content of the Fuel Gas (Stream 27, Figure 3-2), per volume of Raw Methanol; typical units are BTU/gallon.
Syngas Utilization	Defined as the number of standard cubic feet of Balanced Gas plus CO Gas to the Demonstration unit required to produce one pound of Raw Methanol.

EXECUTIVE SUMMARY

This project, which is sponsored by the U. S. Department of Energy (DOE) under the Clean Coal Technology Program to demonstrate the production of methanol from coal-derived synthesis gas (syngas), has completed start-up and entered the four-year operating phase of the program in April of 1997. The purpose of this Public Design Report for the "Commercial-Scale Demonstration of the Liquid Phase Methanol (LPMEOH™) Process" is to provide the public with details of the project. The LPMEOH™ Demonstration Project is a $213.7 million cooperative agreement between the DOE and Air Products Liquid Phase Conversion Company, L.P. (the Partnership). The LPMEOH™ Process Demonstration Unit is located at the Eastman Chemical Company (Eastman) chemicals-from-coal complex in Kingsport, Tennessee.

On 4 October 1994, Air Products and Chemicals, Inc. (Air Products) and Eastman signed the agreements that formed the Partnership, secured the demonstration site, and provided 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 (Mod. A008 to the Cooperative Agreement). The design and construction of the 80,000 gallon/day LPMEOH™ demonstration unit was completed in January of 1997.

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 process development facility 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 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, fuel-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 LPMEOH™ demonstration unit contains the following major areas:

Feed Gas Purification

- Feed gas supply
- Guard bed

Compression

- Recycle gas compressor

Reactor Loop and Catalyst Reduction

- LPMEOH™ Reactor
- Product cooling/feed gas heating
- Catalyst and oil handling
- Catalyst reduction
- Slurry transfer

Distillation

- Light gas removal/stabilization
- Heavy alcohols/water removal

Storage and Miscellaneous

- Oil storage
- Instrument air system
- Vent system

The following major utilities are supplied by the existing Eastman complex:

- Coal-derived syngas
- Cooling water
- Boiler feed water
- High pressure steam (for start-up)
- Low pressure steam
- Power
- Plant air and nitrogen

In addition, the existing Eastman boilers provide an efficient method for recovering energy from the purge gases from the demonstration unit.

The infrastructure provided by Eastman (coal-derived syngas, site, utilities, operating force, etc.) allowed for a moderate cost demonstration of this technology. The total estimated project cost including the four years of operation is $213,700,000. The DOE's cost share is $92,700,000 with the remaining funds coming from the Partnership. The capital cost for the demonstration unit (including tie-ins) is approximately $31,000,000.

The commercial application of the LPMEOH™ technology with the greatest long-term potential is the coproduction of methanol within an IGCC power plant. The ability of the slurry reactor to ramp rapidly and to start/stop easily makes it amenable to peak shaving scenarios in an IGCC power plant. Of significant interest is that relatively small quantities (30,000 gallons/day) of methanol can be coproduced from coal at world scale (600,000 gallons/day) economics.

1.0 PROJECT OVERVIEW

1.1 PURPOSE OF THE PUBLIC DESIGN REPORT

The purpose of the Public Design Report for the "Commercial-Scale Demonstration of the Liquid Phase Methanol (LPMEOH™) Process" project is to consolidate for public use in one document nonproprietary design information that is available at the end of construction and start-up. The report also contains background information, an overview of the project, and pertinent cost data.

The scope of the report is limited to nonproprietary information. Therefore, although its content is insufficient to provide a complete tool for designing a Liquid Phase Methanol unit, it will serve as a reference for the design considerations involved in developing a commercial-scale facility. The report also discusses the advantages of this innovative technology relative to other commercial methanol processes.

1.2 BRIEF DESCRIPTION OF THE PROJECT

1.2.1 Background and History of the Project

The purpose of this project is to demonstrate the commercial viability of the Liquid Phase Methanol (LPMEOH™) process using coal-derived synthesis gas (syngas). This project is being conducted pursuant to the U. S. Department of Energy (DOE) Clean Coal Technology Program.

The United States needs future sources of alternative liquid fuels. With domestic oil production declining and imports increasing, the potential of producing affordable liquid fuels from non-petroleum sources could one day prove both strategically and economically important. The LPMEOH™ process offers an extremely attractive route to supplementing our liquid fuel supplies with methanol made from the abundant coal reserves of the United States.

Methanol also has a broad range of commercial applications. It can be substituted for or blended with gasoline to power vehicles. It is an excellent fuel for the rapid-start combustion turbines used by utilities to meet peak electricity demands. It contains no sulfur and has exceptionally low nitrogen oxide characteristics when burned. It can also be used as a chemical feedstock.

Air Products and Chemicals, Inc. (Air Products) and Eastman Chemical Company (Eastman) have entered into a partnership known as Air Products Liquid Phase Conversion Company, L.P. (The Partnership). The Partnership and the DOE are participating in the Clean Coal Technology Program demonstration of LPMEOH™ technology. The Partnership designed, built, and now owns and operates a nominal 80,000 gallon/day LPMEOH™ process demonstration unit at Eastman's chemicals-from-coal complex in Kingsport, Tennessee. The program objectives are to demonstrate the scale-up and operability (up to four years) of the LPMEOH™ process under various coal-based syngas feed compositions and to gain operating experience for future syngas conversion projects.

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 the Alternative Fuels Development Unit (AFDU), a small, DOE-owned experimental unit in LaPorte, Texas, the technology offers significant potential, compared to conventional gas phase methanol production technologies, to reduce electric power generation costs at coal-based IGCC facilities with the coproduction of chemical feedstocks and alternative liquid fuels. The domestically developed LPMEOH™ technology uses domestic coal reserves to produce clean, storable, liquid fuels and chemical feedstocks. Eventual commercialization of the LPMEOH™ process would provide chemical feedstock and electric power cost savings, lower sulfur dioxide (SO_2) and nitrogen oxide (NO_x) emissions, and could reduce the use of imported liquid fuels.

The LPMEOH™ technology to be demonstrated at the Eastman facility could someday be used as an adjunct to an IGCC power plant—one of the cleanest and most efficient of the 21st century power generating options. When the IGCC power plant is not generating electricity at its full capacity, excess syngas can be diverted to make methanol in a fast ramping LPMEOH™ plant. The methanol could be stored on-site and used in peaking turbines or sold as a commercial fuel or a chemical feedstock. In this configuration, the cost of making methanol from coal is likely to be competitive with stand alone natural gas-to-methanol facilities. A discussion of the process options and economics is provided in Section 7 of this report.

The LPMEOH™ demonstration unit is integrated with Eastman's chemicals-from-coal complex, accepting syngas and converting it to methanol, for use as a chemical feedstock within the Eastman facility. A portion of the methanol product will be evaluated as a zero sulfur, low-NO_x combustion fuel for testing in stationary power applications and mobile transportation use (buses, flexible-fuel vehicles, etc.).

These fuel-use tests will provide a basis for the comparison of the product methanol with conventionally accepted fuels with regard to emission levels and economic viability. The program goal of demonstrating methanol as a fuel can lead to the potential for greater use of oxygenated fuels, which burn cleaner than conventional fuels, thereby reducing air emissions from mobile (e.g., buses and vanpools) and stationary (e.g., engines, turbines, and fuel cells) sources.

The DOE, under the Clean Coal Technology Program, provided cost-shared financial assistance for the design, construction, and operation of the commercial-scale LPMEOH™ demonstration unit by The Partnership. Air Products designed and constructed the LPMEOH™ demonstration unit and Eastman is operating it. The demonstration unit is a nominal 260 ton-per-day (or 80,000 gallon/day) unit situated on a 0.6 acre plot within the existing Eastman chemicals-from-coal complex in Kingsport, Tennessee.

The Eastman coal gasification facility has operated commercially since 1983. Eastman currently both produces and purchases methanol for use at the site. At this site, it will be possible to ramp up and down to demonstrate the unique load-following flexibility of the LPMEOH™ demonstration unit for application to coal-based IGCC electric power generation facilities.

The operation at Eastman may also include the production of dimethyl ether (DME) as a mixed coproduct with methanol for demonstration as a potentially storable fuel pending preliminary laboratory and Design Verification Test results along with market analysis. The Demonstration Test Plan includes DME tests as provisional; the Demonstration Test Plan would be updated if the decision is made to proceed with the co-production of methanol and DME. DME tests would be performed during the last two months of the Demonstration.

1.2.2 Project Schedule

The project is divided into the following three phases:

Design
Construction
Operation

The design phase included all of the engineering needed to construct the demonstration unit. This activity started in October of 1993 and was completed in early 1996. The construction phase started in October of 1995 and included the fabrication of the equipment and the field construction tasks (foundations, steel erection, etc.). The demonstration unit was commissioned in March of 1997 and is presently in the third year of demonstration operations. The off-site fuel-use testing is included in the operation phase; this task was initiated in 1997 and is scheduled to be completed in 2000. The demonstration unit will operate for four years to prove the commercial viability of the LPMEOH™ process. The latest “Milestone Schedule Status Report” and the “Work Breakdown Structure” (from the “Statement of Work”) are included in Appendixes B and C.

1.2.3 Site and Facility Description

The 0.6 acre site proposed for the LPMEOH™ demonstration unit is located in Kingsport, Tennessee, at the Eastman chemicals-from-coal complex. The Eastman complex is on the western edge of Sullivan County and includes a small portion of Hawkins County. The world headquarters of Eastman Chemical Company is also located in Kingsport. The Eastman complex also includes the eastern half of Long Island, where the demonstration unit is located adjacent to existing process facilities.

The project includes five major process areas. The Feed Gas Purification area includes the feed gas supply tie-ins and the guard bed. The Compression area contains the recycle gas compressor and associated systems. The Reactor Loop and Catalyst Reduction area includes the reactor with its associated equipment and the slurry handling equipment and utility oil skid associated with catalyst preparation. The Distillation area includes two distillation columns and heat exchangers. The Storage and Miscellaneous area comprises oil storage and other process components. Block flow diagrams showing the process and its integration into the Eastman complex are shown in Section 4, Figures 4-1 and 4-2.

1.2.4 Summary of Testing Program

A Demonstration Test Plan comprising 37 specific tests was prepared from the Technical Objectives set for the project. These are given in Table 3-2. Copies of the Demonstration Test Plan are available from the DOE.

A portion of the methanol product is being tested off-site for its suitability as a stationary-use fuel (gas turbine and fuel cell) and as a vehicle fuel (buses and flexible-fuel vehicles). These fuel-use tests will provide a basis for the comparison of the as-produced methanol with conventionally accepted fuels with regard to emission levels and economic viability. The program goal of demonstrating methanol as a fuel could lead to the potential for greater use of oxygenated fuels, which burn cleaner than conventional fuels, thereby reducing air emissions from mobile and stationary sources.

1.3 OBJECTIVES OF THE PROJECT

1.3.1 Primary Objective

The LPMEOH™ process technology is expected to be commercialized as part of an IGCC electric power generation system. Therefore, the project incorporates the commercially important aspects of the operation of the LPMEOH™ process which would enhance IGCC power generation. These include the following:

- The coproduction of electric power and of high value liquid transportation fuels and/or chemical feedstocks from coal. This coproduction requires that the partial conversion of syngas to storable liquid products be demonstrated.

- Using an energy load-following operating concept which allows conversion of off-peak energy, at attendant low value, into peak energy commanding a higher value. The load-following concept makes use of gasifier capacity that is under utilized during low-demand periods by using the LPMEOH™ process to convert the excess syngas to a storable liquid fuel for use in electric power generation during the peak energy periods. This operating concept requires that on/off and syngas load following capabilities be demonstrated.

During operation, the instrumentation system will allow for the collection of engineering data, analysis, and reporting which will be done by technical personnel. Typical reporting will include on-stream factors, material and energy balances, reactor and equipment performance, comparison with laboratory and AFDU results, conversion efficiencies, and catalyst activity. The resulting data base will be used to quantitatively evaluate the LPMEOH™ process technology compared to other commercially available methanol- synthesis process technologies.

1.3.2 Secondary Objective

A secondary objective of the project is to demonstrate the production of DME (dimethyl ether) as a mixed coproduct with methanol.

Subject to Design Verification Testing (DVT), the Partnership proposes to enhance the project by including the demonstration of the slurry reactor's capability to produce DME as a mixed co-product with methanol.

DVT is required to address issues such as catalyst activity and stability and to provide data for engineering design and demonstration decision making.

At the conclusion of the DVT, a joint Partnership/DOE decision will be made regarding continuation of the methanol/DME demonstration. Timing of the final decision must ensure that the necessary design, procurement, construction, and commissioning can be completed to allow for operation at the end of the LPMEOH™ demonstration test period.

1.4 SIGNIFICANCE OF THE PROJECT

If successful this project will demonstrate the commercial viability of Air Products' LPMEOH™ process using coal-derived syngas, a mixture of hydrogen (H_2) and carbon monoxide (CO). The DOE's purpose for the demonstration of the proposed project is to help fulfill the goals and objectives of the Clean Coal Technology Program by demonstrating the potential of a more efficient, liquid phase reaction process as a preferred alternative to gas phase reactions for methanol production.

The United States needs future sources of alternative liquid fuels. With domestic oil production declining and imports increasing, the potential of producing affordable liquid fuels from non-petroleum sources could one day prove both strategically and economically important. The LPMEOH™ process offers an extremely attractive route to supplementing liquid fuel supplies with methanol made from abundant coal reserves in the United States.

Methanol has a broad range of commercial applications. It can be substituted for or blended with gasoline to power vehicles. It is an excellent fuel for the rapid-start combustion turbines used by utilities to meet peak electricity demands. It contains no sulfur and has exceptionally low nitrogen oxide characteristics when burned. It can also be used as a chemical feedstock.

The technology now in the demonstration phase is ideally suited as an adjunct to a coal-based IGCC power plant—one of the cleanest and most efficient of the 21st century power generating options. When the power plant is not generating at its full capacity, excess syngas can be diverted to make methanol. The methanol could be stored on-site and used in peaking turbines or sold as a commercial fuel or chemical feedstock. In this configuration, the cost of making methanol from coal is likely to be competitive with stand-alone natural gas-to-methanol facilities.

Air Products and Eastman entered into a partnership known as Air Products Liquid Phase Conversion Company, L.P. (the Partnership). The Partnership is participating with the DOE in the demonstration of Liquid Phase Methanol technology under the Clean Coal Technology Program. Air Products had primary responsibility for the design and construction of the facility and Eastman has primary responsibility for its operation. The facility production is rated at a nominal 80,000 gallons/day.

The program objectives are to demonstrate the scale-up and operability (up to four years) of the LPMEOH™ process under various coal-based feed gas compositions and to gain operating experience for future syngas conversion projects. If practical, the production of DME as a mixed co-product with methanol will also be demonstrated.

LPMEOH™ technology offers significant potential to economically produce chemical feedstocks (using a technology developed in the United States, LPMEOH™, over conventional gas phase methanol production technologies) and to reduce electric power generation costs with the production of alternative liquid fuels. The domestically developed LPMEOH™ technology can utilize the abundant coal reserves of the United States to produce clean, storable, liquid fuels and chemical feedstocks. Eventual commercialization of the LPMEOH™ process in IGCC power plants would provide low priced chemical feedstocks and fuel leading to electric power generation cost savings, lower sulfur dioxide (SO_2) and nitrogen oxide (NO_x) emissions, and the reduced use of imported liquid fuels.

1.5 DOE'S ROLE IN THE PROJECT

1.5.1 Clean Coal Technology Program

The DOE's Clean Coal Technology Program has been implemented through five competitive solicitations. Congress set the basic goals for the program and for each solicitation in the enabling legislation and accompanying report language. DOE subsequently translated the guidance into performance-oriented solicitations. For each solicitation, evaluation criteria were defined and weighted to reflect specific congressional guidance and the current program objectives. This process enabled industry to set the technical agenda by allowing companies to propose their own technologies as qualifying projects. This had the significant benefit of attracting higher levels of private-sector cost-sharing and increasing the likelihood of realizing commercialization objectives.

An important attribute to the solicitation approach used to implement the program was the use of multiple solicitations spread over a number of years. Allowing time between solicitations made it possible to adjust program implementation. At the end of each solicitation, Congress provided the flexibility as needed to effectively implement the program.

Each solicitation was issued as a Program Oppotunity Notice (PON). Proposals for demonstration projects consistent with the objectives of each PON were submitted to DOE by a specific deadline. DOE evaluated the proposals and announced those projects selected for negotiation.

The objective of the Round III was to solicit cost-shared clean coal technology projects to demonstrate innovative, energy-efficient technologies capable of being commercialized in the 1990's. These technologies were to be capable of (1) achieving significant reductions in emissions of SO_2 and/or NO_x from existing facilities to minimize environmental impacts, such as transboundary and interstate pollution, and/or (2) providing for future energy needs in an environmentally acceptable manner. DOE received 48 proposals and selected 13 projects as best furthering the goals and objectives of the PON.

The LPMEOH™ demonstration project was selected in December of 1989 under Round III of the Clean Coal Technology Program.

1.5.2 Management Plan

The DOE entered into a Cooperative Agreement with Air Products and Chemicals, Inc. which was later novated to the Partnership (between Air Products and Eastman) to conduct this project. The DOE is monitoring the project through the Contracting Officer and the Contracting Officer's Technical Representative (COTR). The Partnership is managing the project through an Air Products Program Manager, who is assisted by a team of technical and managerial personnel from Air Products and from Eastman.

The Air Products organizational chart for the design and construction phases is shown in Figure 1.5.2-1. The Air Products organizational chart for the operating phase is shown in Figure 1.5.2-2.

1.5.2.1 DOE

The DOE is responsible for monitoring all aspects of the project and for granting or denying approvals required by the Cooperative Agreement. The DOE Contracting Officer is the authorized representative of the DOE for all matters related to the Cooperative Agreement.

The DOE Contracting Officer will appoint a COTR, who is the authorized representative for all technical matters and will have the authority to issue "Technical Advice." The COTR also approves those reports, plans, and technical information required to be delivered by the Partnership to the DOE under the Cooperative Agreement. The DOE COTR does not have the authority to issue any technical advice that assigns additional work outside the Statement of Work, increases or decreases the total estimated cost or time required for performance of the Cooperative Agreement, changes any of the terms, conditions, or specifications of the Agreement, or interferes with The Partnership's right to perform the terms and conditions of the Agreement. All Technical Advice will be issued in writing by the DOE COTR.

1.5.2.2 The Partnership

The Air Products Program Manager will coordinate the overall project and will be responsible for all communication with the DOE and for interfacing with the DOE COTR.

TABLE 1.5.2-1

KINGSPORT LPMEOH™ PROJECT ORGANIZATION
DESIGN AND CONSTRUCTION PHASES

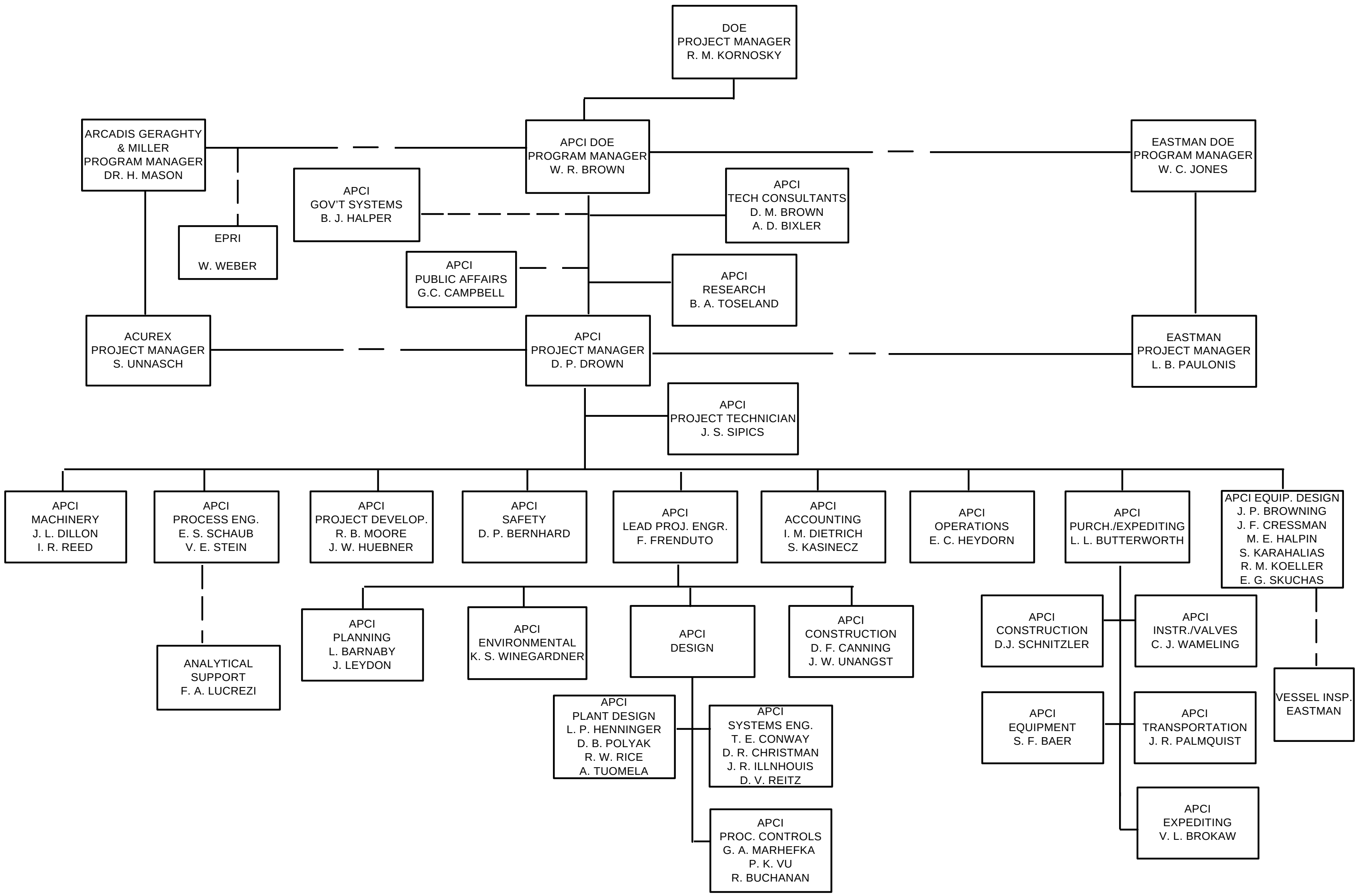
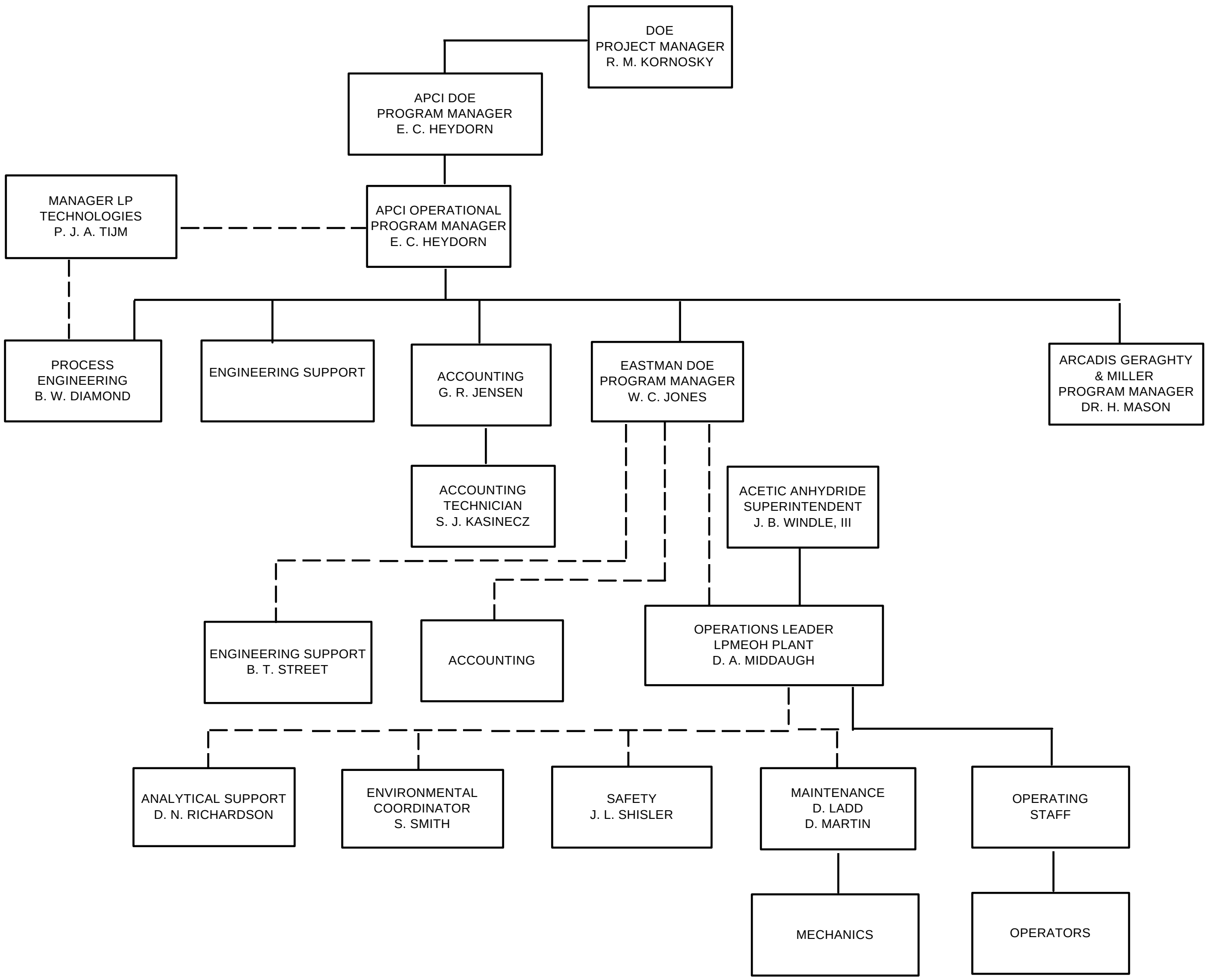


TABLE 1.5.2-2

KINGSPORT LPMEOH™ DEMONSTRATION OPERATING PHASE ORGANIZATION CHART



The Air Products Program Manager is the primary focal point for this project having responsibility and authority for direction of the project subsequent to the signing of the Cooperative Agreement. The Air Products Program Manager is the principal representative between the Partnership and the DOE.

The Program Manager's responsibilities encompass both technical and fiscal considerations, including the following:

- Overall technical coordination of the program
- Monitoring of program cost
- Monitoring of program planning
- Monitoring of program schedule
- Commitment of resources to optimize performance under the Cooperative Agreement
- Reporting requirements
- Final review of all contract deliverables

Throughout the course of this project, reports dealing with technical, cost, and environmental aspects of the project will be prepared by the Partnership and provided to DOE.

2.0 TECHNOLOGY DESCRIPTION

2.1 BRIEF DESCRIPTION OF THE TECHNOLOGY BEING USED

Methanol is a versatile commodity chemical with applications ranging from synthetic fuels to feedstocks for higher valued chemicals such as methyl methacrylate and methyl tertiary-butyl ether (MTBE an octane booster). Many applications have evolved for methanol as a fuel including methanol as a gasoline extender, the methanol to gasoline process, and integrated gasification-combined cycle (IGCC) technology for the coproduction of methanol and electricity.

The Liquid Phase Methanol (LPMEOH™) process represents a major departure from traditional gas phase routes to methanol in the method of removing the heat of reaction. The reactions of hydrogen and carbon oxides to form methanol are highly exothermic.

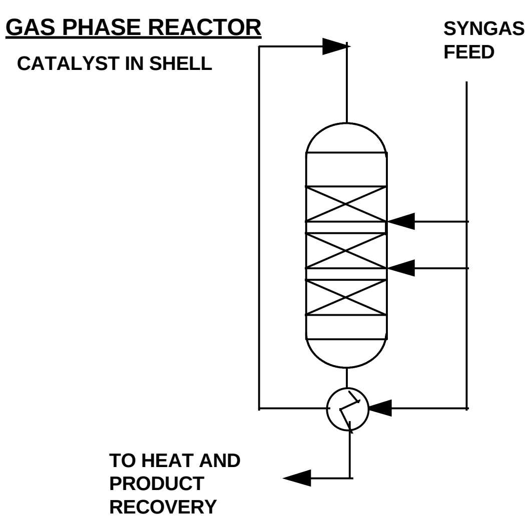
$$CO + 2H_2 \longrightarrow CH_3OH + 21.7\ KCAL/gmol$$

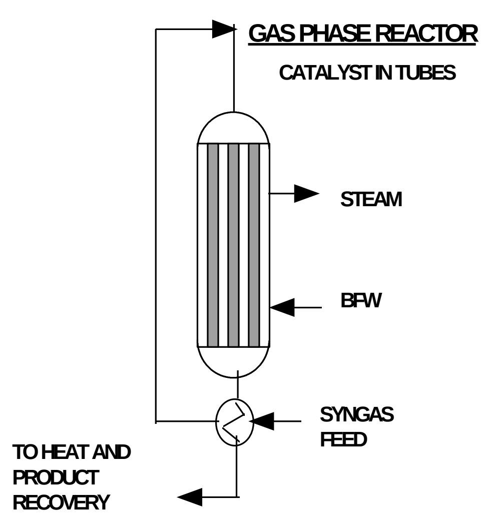
$$CO_2 + 3H_2 \longrightarrow CH_3OH + H_2O + 12.8\ KCAL/gmol$$

TYPICAL REACTION CONDITIONS:
1,000 psig
440°–520°F

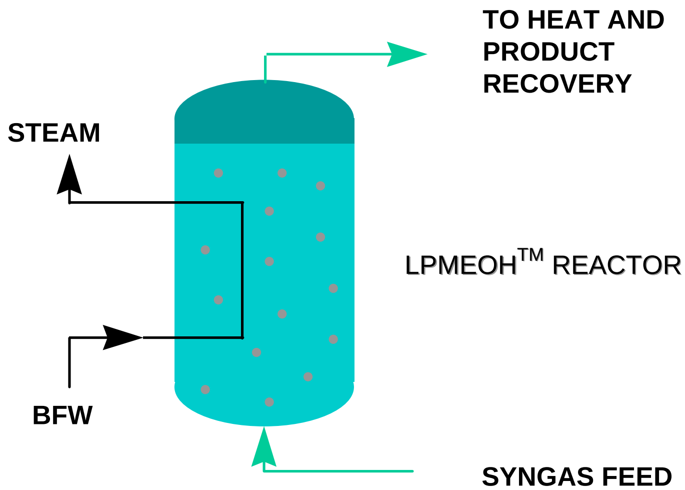
One of the most difficult design problems of the methanol synthesis process is removing the heat of reaction while maintaining precise temperature control to achieve optimum catalyst life and reaction rate. Catalyst life is seriously reduced by excessive temperatures.

Reactor schematics for conventional gas phase reactors are show below. The catalyst is in the form of pellets. In these conventional gas phase reactors, either cool unreacted gas is injected at stages in the catalyst bed, or internal cooling surfaces are used to provide temperature control.





These schemes, however, were developed for diluted syngas which yields low conversion per pass. The LPMEOH™ process uses fine catalyst particles entrained in an inert hydrocarbon liquid, usually a mineral oil. The mineral oil acts as a temperature moderator and a heat removal medium, transferring the heat of reaction from the catalyst surface via the liquid slurry to boiling water in an internal tubular heat exchanger. As a result of this capability to remove heat and maintain a constant, highly uniform temperature through the entire length of the reactor, the slurry reactor can achieve much higher syngas conversion per pass than its gas phase counterparts.



In addition, the LPMEOH™ process is particularly well suited to coal-derived syngas which is rich in carbon monoxide. These capabilities make the LPMEOH™ process a potentially lower-cost conversion route to methanol, especially when methanol coproduction is added to a coal-based IGCC power plant. For a modest increase in complexity of an IGCC power plant, the methanol coproduction scheme produces a storable liquid fuel in parallel with electric power production, providing a significant turndown and peak-load capability for the IGCC power plant. Information on the application of the LPMEOH™ process to the IGCC power plant is provided in Section 7 of this report.

3.0 PROCESS DESIGN CRITERIA

A definition of the abbreviations, units, and terms is provided in the beginning of this document.

3.1 PLANT CAPACITY

The design basis is set by the commitment of syngas as follows: Eastman will dedicate 990 KSCFH (based on 379.4 SCF/lb-mole at 14.7 psia and 60°F) of coal-derived syngas. Eastman will also make available additional quantities of syngas as available for special short-term tests.

Gas Flows (KSCFH)	DEDICATED RANGE			DESIGN
Balanced Gas	990	to	900	900
CO Gas	0	to	50	50
H_2 Gas	0	to	40	40
Total		990		990

Fresh Feed Compositions (vol%)	H_2	CO	Carbon Dioxide (CO_2)	Nitrogen (N_2)
Balanced Gas	67.8	27.7	4	0.5
CO Gas	2	97	0	1
H_2 Gas	74	11	3	12

The production capacity will be variable depending upon catalyst activity and operating mode. The normal operating mode will be at reduced (more efficient) feed gas rates of about 86%, utilizing a total of 765 KSCFH of Balanced Gas, 40 KSCFH of CO Gas and 45 KSCFH of H_2 Gas to produce 71,800 gallons/day of methanol.

For permitting purposes the demonstration unit was assumed to be capable of producing up to 90,000 gallons/day following future debottlenecking and operating at a 100% on-stream factor. Syngas is expected to be available 340 days per year, 93% of the time. The LPMEOH™ demonstration unit will operate on average 320 days/year (per the test plan).

	Methanol Product
Flow - Design, gallons/day	79,100
- Normal, gallons/day	71,800
- Min., gallons/day	39,400 @ 50% load following
Pressure	Atmospheric
Temperature, °F	100

The methanol produced will be a “refined grade” methanol suitable for use in one or more of the Eastman process units. The allowable composition of this stream is given in Table 3-1. About 25% of the crude methanol production, including the water and heavy components (higher alcohols, oil) will be directed to the distillation system of Eastman’s existing gas phase methanol unit for further refining to recover 99.7% of the contained methanol.

Table 3-1
Methanol Product Composition

PRODUCT PROPERTIES:	REFINED GRADE	CRUDE GRADE
methyl alcohol	99.8% min	80% min
color, pcs	5 max	
specific gravity @ 20/20 C	0.792 - 0.793	
water content	0.2% max	20% max
acidity, as acetic	0.003% max	
reducing substances, (potassium permanganate)	30 min. @ 59°F	
acetone	30 ppm max	30 ppm max
other alcohols	1,000 ppm max	1.0% max
ethanol	200 ppm max	1.0% max
other impurities	500 ppm max	0.1% max
methyl formate	50 ppm max	30 ppm max
alkalinity, as ammonia	3 ppm max	
oil		1.0% max
formic acid		30 ppm max
non volatiles		0.05% max

3.2 PROCESS DESIGN CONSIDERATIONS

A Demonstration Test Plan was prepared from the Technical Objectives set for the project. The demonstration test plan comprises 37 specific tests. These are given in Table 3-2. The conditions imposed by each of these tests is incorporated in the process design. For additional details on each of these tests please refer to the “Demonstration Test Plan (Final) September 1996.”

Test Run #	Test Run Description	Temp (Deg C)	Wt% Cat	H2/CO Ratio at Inlet	Space Velocity (Sl/hr-kg)	MeOH (tpd)	Fresh Feed Balanced (KSCFH)	Fresh Feed CO Gas (KSCFH)	Fresh Feed H2 Gas (KSCFH)	Recycle Gas (KSCFH)	Inlet Sup. Velocity (ft/sec)	Time Period (weeks)	Elapsed Time (incl. outages) (weeks)	Start of Test
Task 2.1.1 - Process Shakedown and Catalyst Aging:														
1.	Initial Shakedown; and	250	28	2.42	8,000	260	900	50	40	1,800	0.64	6	6	Feb-97
	Design Production Tests						(varies, to maintain syngas utilization.)							
2.	Gassed Slurry Level	Part of other tests										Ongoing		
3.	Reactor Feed: Texaco-	250	28	0.67	9,240	202	650	95 (*)	0	2,612 (*)	0.77	2	9	Mar-97
	Type Syngas													
4.	Early Testing @ High	250	28	2.54	10,300	TBD	1,200 (**)	50	40	2,520 (*)	0.88	2	12	Apr-97
	Superficial Velocity													
5.	Check @ Test 1	250	28	2.42	8,000	< 260	900	50	40	1,800	0.64	2	15	May-97
	Conditions													
6.	Catalyst Addition and	250	28 - 40	2.51	Dec. from	237	765	40	45	Max	0.79	18	41	May-97
	Aging	or less		2.30	8,000	260	900	50	40	(2,700)	TBD	6		to
	(Note: Kingsport Complex Outage during this test)													Nov-97
7.	Free-Drain Entrained/	250	28 - 40	2.51	Dec. from	237	765	40	45	Max	0.79	During		
	Condensed Oil to	or less			8,000							Test 6		
	Reactor													
8.	Operation @ Design	250	40	2.42	4,000	260	900	50	40	1,800	0.64	2	43	Nov-97
	Feed Gas Rates													
9.	Check for Limitation on	250	> 40	2.51	Varies	TBD	765	40	45	Max	0.79	6	50	Dec-97
	Catalyst Slurry Concentration									(2,700)				
10.	Catalyst Addition to	250	Target	2.49	3,320	256	765	40	45	2,605	0.79	12	68	Jan-98
	Reach Max Productivity	or less	45	2.29	3,500	293	900	50	40	2,520	0.81	2		
				TBD	TBD	TBD	1,110 (**)	50	40	2,520	0.86	2		

Test Run #	Test Run Description	Temp (Deg C)	Wt% Cat	H2/CO Ratio at Inlet	Space Velocity (Sl/hr-kg)	MeOH (tpd)	Fresh Feed Balanced (KSCFH)	Fresh Feed CO Gas (KSCFH)	Fresh Feed H2 Gas (KSCFH)	Recycle Gas (KSCFH)	Inlet Sup. Velocity (ft/sec)	Time Period (weeks)	Elapsed Time (incl. outages) (weeks)	Start of Test
Task 2.1.2 - Process Operational Test Phase:														
	Note: At this time, need to produce some "typical" product methanol for off-site fuel tests. Also need to reassess the optimum operating conditions for the remaining tests (e.g. feed gas allocation for commercial design/optimal performance).													
11.	Catalyst Addition/ Withdrawal Test	250	Target 45	2.49	3,320	256	765	40	45	2,605	0.79	6	74	May-98
12.	Test 11 Conditions with No CO Make-up	250	Target 45	4.97	3,282	229	765	0	45	2,605	0.78	2	76	Jul-98
13.	Test 11 Conditions with No H2 Make-up	250	Target 45	1.98	3,277	252	765	40	0	2,605	0.78	2	79	Jul-98
14.	Test 11 Conditions with No H2 or CO Make-up	250	Target 45	5.03	3,238	232	765	0	0	2,605	0.77	2	81	Aug-98
15.	Repeat of Test 11 Conditions	250	Target 45	2.49	3,320	256	765	40	45	2,605	0.79	2	83	Aug-98
16.	Design Fresh Feed Operation Test	250	Target 45	2.29	3,500	293	900	50	40	2,520	0.81	2	86	Sep-98
17.	Testing @ High Superficial Velocity	250	Target 45	TBD	TBD	TBD	1,110 (**)	50	40	2,520	0.86	2	88	Oct-98
18.	Turndown and Rampin(	250	Target 45	3.30	1,825	151	450	25	60	1,364	0.44	4	92	Oct-98
19.	Load-Following, Cyclone, & On/Off Tests		Target 45	Balanced, CO-Rich	To be Defined							6	99	Nov-98
20.	Reactor Feed: Texaco-Type Syngas	250	Target 45	0.69	2,870	207	650	85 (**)	0	2,195	0.67	4	103	Jan-99

Test Run #	Test Run Description	Temp (Deg C)	Wt% Cat	H2/CO Ratio at Inlet	Space Velocity (Sl/hr-kg)	MeOH (tpd)	Fresh Feed Balanced (KSCFH)	Fresh Feed CO Gas (KSCFH)	Fresh Feed H2 Gas (KSCFH)	Recycle Gas (KSCFH)	Inlet Sup. Velocity (ft/sec)	Time Period (weeks)	Elapsed Time (incl. outages) (weeks)	Start of Test
21.	Reactor Feed: Destec-Type Syngas	250	Target 45	1.01	2,770	215	670	65 (***)	0	2,147	0.67	3	106	Jan-99
22.	Reactor Feed: BGL-Type Syngas	250	Target 45	0.52	2,165	137	485	200 (***)	0	1,568	0.43	3	109	Feb-99
23.	Repeat of Test 15 Conditions	250	Target 45	2.49	3,320	256	765	40	45	2,605	0.79	2	112	Mar-99
24.	Reactor Feed: Nat. Ga Reformer-Type Syngas	250	Target 45	4.98	1,978	197	765	0	30	1,264	0.48	3	115	Apr-99
25.	Reactor Feed: Shell-Type Syngas with Steam Injection and 1:1 Recycle	250	Target 45	0.53	1,471	101	238	400 (***)	50	842	0.35	3	118	Apr-99
26.	Repeat of Test 15 Conditions	250	Target 45	2.49	3,320	256	765	40	45	2,605	0.79	2	121	May-99
27.	Repeat of Test 16 Conditions	250	Target 45	2.29	3,500	293	900	50	40	2,520	0.81	2	123	Jun-99
28.	Reactor Operation @ 260 deg C	260	Target 45	2.51	3,320	248	765	40	45	2,605	0.79	2	125	Jun-99
29.	Repeat of Test 26 Conditions	250	Target 45	2.49	3,320	256	765	40	45	2,605	0.79	2	127	Jul-99
30.	Reactor Inspection - Then, Continue Operational Tests - with Alternative Catalyst):											4	131	Jul-99
31.	Plant Shakedown	240	TBD	2.42	TBD	260	900	50	40	Max(TBD)	TBD	6	137	Aug-99
32.	Reactor Feed: Texaco-Type Syngas	240	TBD	0.67	TBD	202	650	95 (*)	0	2,612 (*)	0.77	2	140	Sep-99

Test Run #	Test Run Description	Temp (Deg C)	Wt% Cat	H2/CO Ratio at Inlet	Space Velocity (Sl/hr-kg)	MeOH (tpd)	Fresh Feed Balanced (KSCFH)	Fresh Feed CO Gas (KSCFH)	Fresh Feed H2 Gas (KSCFH)	Recycle Gas (KSCFH)	Inlet Sup. Velocity (ft/sec)	Time Period (weeks)	Elapsed Time (incl. outages) (weeks)	Start of Test
33.	Catalyst Aging	240	TBD	2.50	TBD	237	765	40	45	2,605		16	162	Oct-99
				2.31	TBD	260	900	50	40	2,520	TBD	4		
34.	Catalyst Addition/	240 -	Target	Balanced	To be					Max(TBD)		6	168	Mar-00
	Withdrawal to Achieve	250	45		defined									
	Target Slurry Concentration													
35.	Reactor Feed: Texaco-	250	Target	0.69	2,870	207	650	85 (**)	0	2,195	0.67	4	173	Apr-00
	Type Syngas		45											
Task 2.1.3 - Extended Optimum Operation:														
36.	Stable Operation	250	Target	2.49	3320	256	765	40	45	2605	0.79	16	200	Jun-00
		250	45	2.29	3500	293	900	50	40	2520	0.81	6		
		250		TBD	TBD	TBD	1,110 (**)	50	40	2520	0.86	2		
37.	DME Demo (Task 2.2)		Balanced and CO-Rich									TBD		Dec-00
	or													to
	Commercial Test Run (Task 2.1.3)		Target	2.49	3320	256	765	40	45	2605	0.79	8	208	Mar-01
			45											
		Syngas Outages (5%, including Kingsport complex outage during Test 6)										10		
		Planned LPMEOH Outages (including Reactor Inspection and Fresh Catalyst Charge in Test 28)										12		
		Unplanned LPMEOH Outages (10/yr @ 8 hrs.)										2		

Notes:

(*) - 700 HP motor on 29K-01 Compressor allows higher recycle gas flow than design basis.

(**) - 1200 KSCFH of Fresh Feed Syngas can be made available for testing (per Eastman debottlenecking of gasification area). Final decision on test will depend upon carbonyl concentrations in Balanced Gas and CO Gas, since 29C-40 Carbonyl Guard Bed will have to be bypassed. For this condition, test execution is subject to availability of CO Gas.

(***) - Subject to availability of CO Gas.

3.3 SITE-SPECIFIC DESIGN CONSIDERATIONS

The following utilities are supplied by Eastman at the site battery limits.

3.3.1 Cooling Water

A baseline normal flow of 1,300 gpm (2,000 gpm max.) of cooling water will be supplied by Eastman. The design supply pressure and temperature is 86 psig at 85°F. The allowable temperature rise is 20°F. The return header pressure is 43 psig.

The cooling water will have conductivity of 800–1,000 micro siemens and pH between 7.0 and 7.5. The system will be operated at 4-5 cycles of concentration. The make-up water will contain 90–120 mg/l alkalinity (essentially all is in bicarbonate form), 24–32 mg/l calcium, 8–10 mg/l magnesium, 13 mg/l sulfate, 6 mg/l chloride, and 5 mg/l sodium.

3.3.2 Boiler Feed Water

Medium pressure boiler feed water (BFW) make-up to the steam drum for the reactor heat exchanger will be supplied by Eastman. The boiler feed water is demineralized, deaerated, and treated with sodium diphosphate. Eastman uses the BFW for 1,500 psig steam. The chlorides are kept below the detection limit (<10 ppb); oxygen is below 10 ppb and sodium is 20 ppb after deionization.

The normal flow rate is 50 gpm. The maximum expected flow rate is 62 gpm. The boiler feed water is supplied at:

Source Normal	User Maximum	User Minimum
260 psig @ 325°F	300 psig	240 psig

3.3.3 Firewater

A connection to the existing fire water protection system is provided at the LPMEOH™ block. The existing firewater system supplies 5,000 gpm at a pressure of 125 psig. The firewater is supplied from a 1,000,000 gallon storage tank on Long Island.

The total firewater supply rate is 3,200 gpm.

3.3.4 City Water and River Water

3.3.4.1 Potable Water

Potable water for eye baths/emergency showers in the demonstration unit is supplied by Eastman at:

Source Normal	User Maximum	User Minimum
110 psig	120 psig @ 1,205 ft. elev.	65 psig at 1,205 ft. elev.

3.3.4.2 River Water

A 4-inch line provides river water to the facility. River water is piped to all utility stations and is also used to temper the boiler blowdown before it is sent to the process sewer. River water design flow rate is 30 gpm.

3.3.5 Steam

The demonstration unit imports a maximum of 38,300 lbs/hr (typical is 4,070 lbs/hr) of 100 psig steam from Eastman. Eastman also provides 600 psig start-up steam (approx. 6,000 lbs/hr) for cold starts about 6 times per year. Steam pressure levels are:

Nominal Pressure	Source Normal	User Maximum (Eastman relief device setting)	User Minimum
100 psig	100 psig at 375°F	125 psig at 450°F	95 psig at 335°F
600 psig	580 psig at 750°F	640 psig at 750°F	575 psig at 600°F

The demonstration unit will export steam to the Eastman header during the times when the rectifier column is not being used (i.e., when fuel methanol is being produced for off-site fuel-use demonstrations).

3.3.6 Nitrogen

A total of 65 KSCFH of nitrogen at 110 psig is supplied by Eastman. The baseline flow will be 5 KSCFH for purging and an additional 60 KSCFH is used for catalyst reduction and start-up. The purity required is 10 ppm oxygen maximum.

Source Normal	User Maximum	User Minimum
110 psig	150 psig	100 psig

3.3.7 Power

Electrical power is supplied by Eastman. The installed requirement is approximately 1,041 KW at 480 volts and 490 KW at 4,160 volts.

3.3.8 Instrument Air

A packaged air dryer utilizing plant air as the feed provides dry, oil-free air for instrument uses. The dryer is rated at 300 SCFM and produces a dry air stream with a dew point of -40°F at a line pressure of 80 psig.

3.3.9 Plant Air

Service air can contain oil-mist, oil-water emulsions, and chemicals. The design conditions are:

Source Normal	User Maximum	User Minimum
80 psig	90 psig	70 psig

3.3.10 Purge Gas

The Eastman purge gas header which is connected to the Eastman boiler system operates as follows:

Source Normal	User Maximum
15 psig	25 psig @ 482°F

3.3.11 Return Condensate

The steam condensate from the LPMEOH™ demonstration unit is returned to the Eastman condensate return system. Maximum flowrate 38,300 lbs/hr; typical 30,640 lbs/hr.

Steam Condensate Source
Steam Tracing on Slurry Lines
Stabilizer Reboiler (29E-10)*
Rectifier Reboiler (29E-20)*
Steam Piping Drip Leg Traps

* - Refer to Table 4-2 and Appendix A, Sheet P3 for information on these equipment items.

3.3.12 Special Engineering Considerations

There are no special tornado or hurricane considerations for this site.

Seismic Zone	2a
Flooding considerations.	None
Wind Design Criteria	
• maximum wind velocity, mph	70
Rainwater design criteria	
10 year, 30 minute rainfall, inches	1.7
annual rainfall, inches	41
Design Temperatures	
Summer high (dry bulb) °F	102
Average summer, °F	>91 for more than 30 hr/year
Summer high, °F (design dry bulb)	95 (10 year)
Summer high, °F (design wet bulb)	78
Average winter, °F	28, <9 for 22 hr/day
Design temperature for freeze protection, °F	-10
Snow accumulation, inches (record)	16

The frost penetration depth is 12–30 inches.

The demonstration unit is located approximately 1,200 feet above sea level.

The Tennessee State law requires use of the Standard Building Code (SBC) where appropriate.

4. DETAILED PROCESS DESIGN

4.1 PLOT PLAN AND EQUIPMENT ARRANGEMENT DRAWINGS

Please see Appendix D for these drawings.

4.2 GENERAL DESCRIPTION

The reactor used in the LPMEOH™ process is unlike conventional gas phase reactors that use fixed beds of catalyst pellets and largely depend upon recycle diluent gas to both dilute the CO concentration and control the temperature rise caused by the heat of reaction. The LPMEOH™ reactor is a slurry reactor with small, powder-size catalyst particles suspended in inert mineral oil. The syngas bubbles up through the slurry where the H_2 and CO dissolve in the oil and diffuse to the catalyst surface where the methanol reaction occurs. The product methanol diffuses out of the slurry and exits as a vapor with the unreacted syngas. The inert oil acts as a heat sink and permits isothermal operation. The net heat of reaction is removed via an internal heat exchanger which produces steam. Unlike gas phase reactors that limit per-pass conversion of syngas to methanol to accommodate the reaction exotherm, the LPMEOH™ reactor maintains isothermal operation. The methanol vapor leaves the reactor and is condensed to a liquid, sent to distillation columns for removal of higher alcohols, water, and other impurities, and is then stored in lot tanks for sampling prior to being sent to Eastman's methanol storage. A portion of the unreacted syngas is sent back to the reactor with the recycle compressor, improving cycle efficiency. The methanol is used for downstream feedstocks and for off-site fuel-use testing.

Unlike gas phase reactors, the LPMEOH™ reactor is tolerant to CO-rich gas. Shift and CO_2 removal are not required. Low H_2-to-CO ratios are acceptable as is any CO_2 content. Finally, in contrast to the gas phase reactor in which the catalyst is sensitive to flow variations and changes from steady-state, the LPMEOH™ reactor is eminently suited for load-following and on/off operation.

The LPMEOH™ demonstration unit is integrated with Eastman's coal gasification facility and inserted in parallel with an existing gas phase methanol unit.

4.3 DETAILED DESCRIPTION

The LPMEOH™ demonstration unit consists of five main process sections: Feed Gas Purification, Compression, Reactor Loop and Catalyst Reduction, Distillation, and Storage and Miscellaneous. The process flow diagrams for the various sections are shown in Appendix A (Sheets 1 through 7); and an equipment list is provided in Table 4-2. (Block diagrams for the Kingsport complex and the LPMEOH™ demonstration unit are also provided in Figures 4-1 and 4-2). A glossary of syngas terminology is provided in the beginning of this document. A discussion of each major process section, with reference to the specific process flow diagram sheets in Appendix A, follows.

INTEGRATION OF EXISTING FACILITIES WITH LPMEOH™ FACILITY

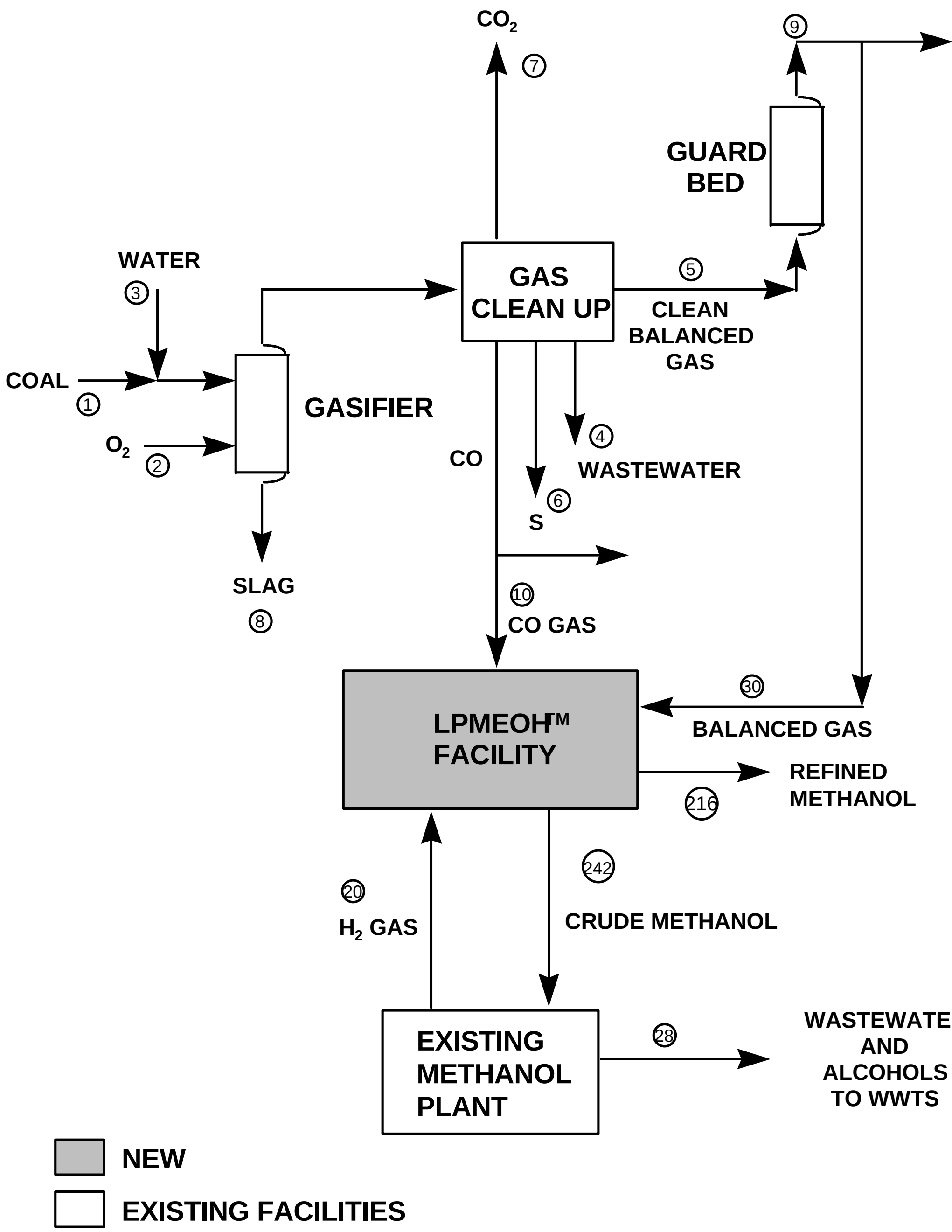


FIGURE 4-1

FSF July 1997

LPMeOH1

"LPMEOH^{TM} FACILITY"
SIMPLIFIED PROCESS FLOW DIAGRAM

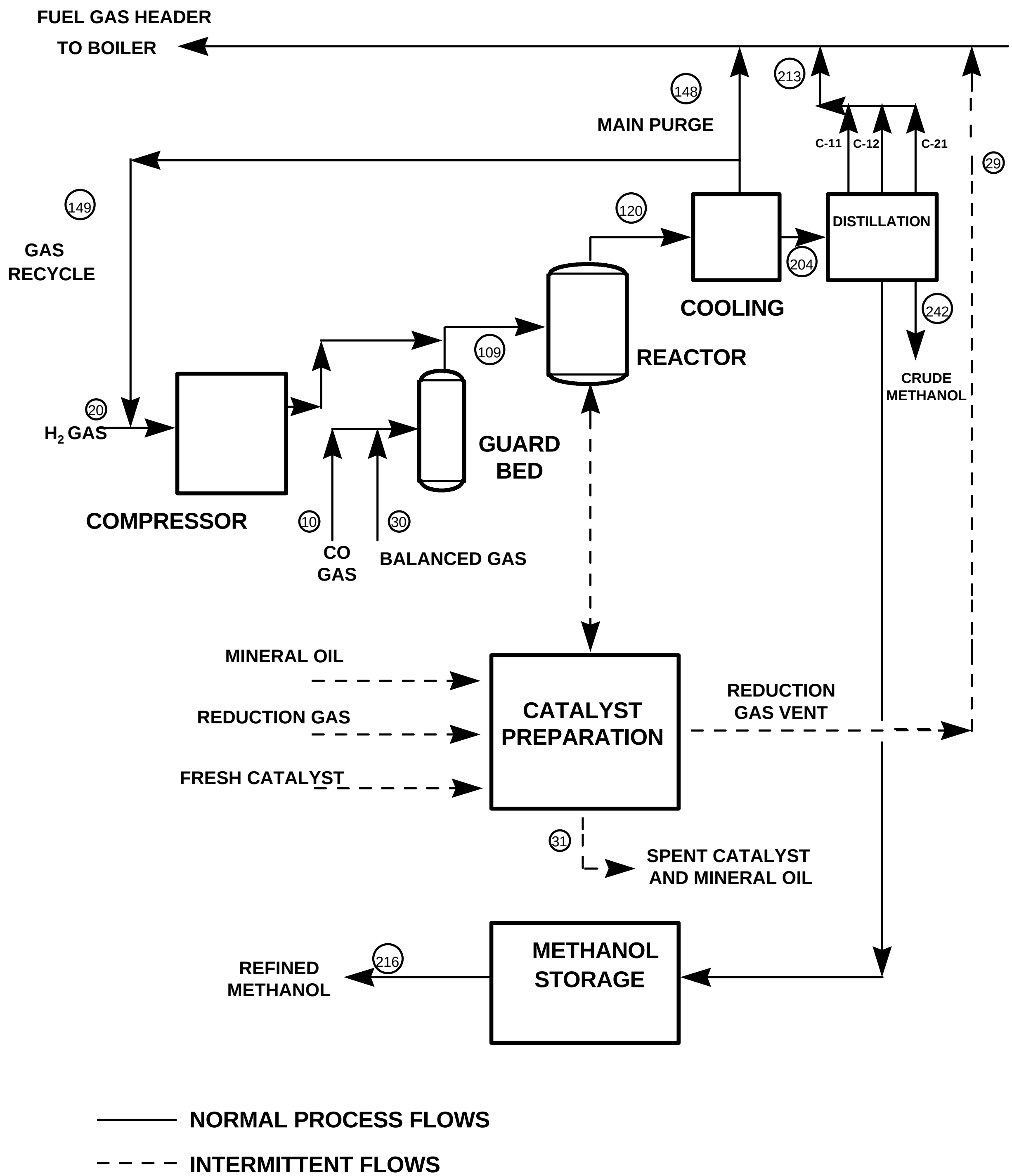


FIGURE 4-2

FSF July 1997

4.3.1 Feed Gas Purification (Sheet 1; also see Figures 4-1 and 4-2)

Three sources of syngas from the Kingsport facility are available to form the LPMEOH™ reactor feed stream. Approximately half of the Balanced Gas fresh feed to the gas phase methanol unit is diverted to the LPMEOH™ demonstration unit (Stream 30). A high purity carbon monoxide (CO) gas stream is also available from the Kingsport cold boxes (Stream 10). The third feed stream is the hydrogen (H_2 Gas) exiting the gas phase unit (Stream 20). The CO Gas and Balanced Gas streams are combined and passed through the (29C-40) carbonyl guard bed. This bed, packed with activated carbon, protects the methanol catalyst against possible upsets of iron and nickel carbonyl contaminants from the upstream gas cleanup units.

4.3.2 Compression (Sheet 1)

Since the H_2 Gas stream is at a lower pressure than the other two feed streams, it is combined with the Recycle Gas stream (Stream 149), made up of unconverted syngas from the LPMEOH™ reactor, and compressed in the (29K-01) recycle compressor. A dry gas seal is used on the compressor to reduce the emissions of syngas to the atmosphere.

4.3.3 Reactor Loop and Catalyst Reduction (Sheets 1, 2, 4 and 3; also see Figures 4-1 and 4-2)

The combined Reactor Feed gas composition is typically by volume 60.9% H_2, 25.1% CO, 4.1% N_2, and 9.0% CO_2 (Stream 109). This high pressure Reactor Feed gas stream is heated to approximately 402°F in the (29E-02) feed/product heat exchanger against the reactor effluent. The feed is then sparged into the (29C-01) LPMEOH™ reactor, mixes with the catalyst slurry and is partially converted to methanol vapor, releasing the heat of reaction to be absorbed by the slurry. The slurry temperature is controlled by varying the steam temperature within the heat exchanger tubes, which is accomplished by adjusting the steam pressure. A Topical Report on the design and fabrication of the LPMEOH™ reactor has been published (Ref. g).

Disengagement of the product gas (methanol vapor and unreacted syngas) from the catalyst/oil slurry occurs in the freeboard region of the LPMEOH™ reactor. Any entrained slurry droplets leaving the top of the reactor are collected in the (29C-06) cyclone separator. An oil flush is maintained to this vessel to assist in the knockout of slurry. The product gas passes through the tubeside of the (29E-02) exchanger, where it is cooled to 250°F by heat exchanging the effluent gas stream against the reactor inlet feed gas stream. The condensed liquid oil droplets are collected in the (29C-05) high pressure oil separator and then pumped back with the entrained slurry from the (29C-06) separator to the LPMEOH™ reactor by the (29G-01A/B) oil circulation pumps. These streams can also be gravity-drained to the LPMEOH™ reactor. To make up for oil losses into the product recovery train, fresh oil is added into the (29C-05) separator via the (29G-03 A/B) pumps. Bypasses have been installed to allow both the (29C-06) and (29C-05) separators to free drain back to the reactor without the use of the (29G-01) pumps. In this mode, the fresh makeup oil would be added as a flush to the (29C-06) separator.

The product gas (Stream 120) is further cooled to 105°F in an air-cooled exchanger (29E-03) and a cooling water exchanger (29E-04). The liquid methanol which is condensed is collected in the (29C-03) product separator. The overhead stream from the (29C-03) product separator contains unreacted syngas, 0.9 vol% uncondensed methanol, and 2 ppmw uncondensed oil. Approximately 91% of this unreacted syngas stream is recycled back to the LPMEOH™ reactor after undergoing compression in the (29K-01) recycle compressor. The balance of the unreacted gas returns to the Kingsport facility at 100°F and is sent to the boilers.

The catalyst slurry is activated in the (29C-30) reduction vessel which is an agitated, 304 stainless steel vessel equipped with a heating/cooling jacket. This vessel has three purposes:

1. Fresh Slurry Mix Tank
2. Catalyst Reduction Vessel
3. Spent Slurry Receiver.

Any reclaimed oil stored in the (29C-31) accumulator is first gravity drained into the top oil fill nozzle of the reduction vessel. The balance of 740 gallons of mineral oil is added using the (29G-34) oil feed pump. The oil is heated to approximately 200°F using the jacketed utility oil skid. Once the oil is at temperature, 2,250 pounds of catalyst oxide is added to form a 30 wt% catalyst slurry mixture. The agitator is used during this time to ensure adequate suspension and more uniform concentration of the slurry.

Reduction gas, consisting of a blend of 96 vol% N_2 and 4 vol% CO, is introduced into the reduction vessel via a gas sparger. The agitator is not required once the reduction gas is introduced. Over the course of the reduction, the slurry temperature is carefully increased while the consumption of CO is monitored to determine when the catalyst is completely reduced. The loss of oil to the vapor phase results in an increase to the catalyst concentration in the slurry from 30 wt% to approximately 40 wt%. The gas stream exiting the reduction vessel is cooled in the (29E-31) condenser, to condense any oil vapors leaving the reduction vessel against the reduction feed. The condensed oil is collected in the (29C-31) separator over the course of the reduction. This oil is reclaimed at the beginning of the next catalyst reduction batch. The temperature in the (29C-31) separator condensate accumulator is controlled by bypassing the reduction feed to minimize the amount of water condensed and collected with the oil.

The catalyst reduction procedure is completed in approximately 20 hours. At the end of reduction, the catalyst is fully active and can be transferred directly to the LPMEOH™ reactor via the (29G-30) slurry transfer pump.

As new catalyst slurry is added to the LPMEOH™ reactor, the catalyst inventory is maintained by withdrawing an equivalent amount of partially deactivated or spent slurry from the reactor. Prior to transferring the slurry from the reactor, the (29C-30) catalyst reduction vessel is pre-warmed using the utility oil skid. The spent slurry is pressure transferred back to the (29C-30) catalyst reduction vessel via the recycle control valve around the (29G-30) slurry transfer pump. Once there, the slurry is purged of the dissolved gases and cooled to a safe handling level at a rate of 60°F/hour using the utility oil system. After cooling, the spent slurry is transferred to the drums or tote bins. The containers are typically shipped off-site to a processor for metals recovery.

4.3.4 Distillation (Sheets 3 and 7)

The condensed methanol (Stream 204) contains 6 volume % dissolved gases, methyl formate, water, and some higher alcohols. These impurities are removed in a two column distillation train which will produce a methyl acetate feed-grade methanol product. The liquid (Stream 204) from the (29C-03) product separator is flashed into the (29C-12) methanol stabilizer feed drum at approximately 70 psig. This vessel has one hour of holdup time to allow for some lag time due to rate and composition changes between the reactor train and the distillation system. Flashed gas from this separator is combined with the overheads of the two columns and sent to the Eastman boilers.

The first distillation column (29C-10) removes the dissolved gases and lighter boiling impurities, such as methyl formate, in the overhead (Stream 211). The bottoms from this column are fed to the second train (29C-20) where the purified methanol product is removed as a top stage distillate product. Any non-condensables are combined with the overhead stream from the (29C-10) and (29C-12). The bottom draw from the (29C-20) is a crude methanol stream containing 25% (by weight) of the raw methanol, and the higher alcohols, water and any of the oil which was carried over from the reactor. This stream is sent to the existing gas phase distillation system for production of additional refined methanol and a stream of the oil, higher alcohols and water, the latter of which is sent to Eastman’s wastewater treatment system.

The methanol product produced from the (29C-20) distillation column is pumped by the (29G-21) methanol rectifier reflux pump to either the (29D-20) or (29D-21) lot tanks. After the appropriate purity checks are completed, the contents of the lot tanks are transferred via the (29G-23) methanol transfer pump to Eastman bulk storage. In some off-design cases where impurities are greater than normal, the lot tanks would be rejected to the distillation system within Eastman's existing methanol plant for recovery.

Product methanol for off-site fuel testing is produced at limited times during the demonstration period by using only the first distillation column. In these circumstances, the bottoms product is cooled in the (29E-23) heat exchanger before transferring to the lot tanks.

4.3.5 Storage and Miscellaneous (Sheets 5 and 6)

Other utility vessels (such as the (29D-30) fresh oil storage tank and the (29D-02) slurry tank) are provided. A (29D-01) vapor-liquid separator is located upstream of the (29C-120) vent stack to limit the volume of entrained liquid entering the sump of the stack. Other utilities include the instrument air skid to reduce the dew point of the plant air and an in-ground separator to collect any spills of process oil within the plant battery limits.

4.4 MATERIAL BALANCE

Table 4-1 shows the material balance for the points shown on Figures 4-1 and 4-2. These points are also shown on the Process Flow Diagrams included in Appendix A.

4.5 WASTE STREAMS

Figure 4-3 shows the major process blocks and the waste streams that are generated by these areas. For a more detailed discussion of waste streams, refer to the Environmental Information Volume (EIV), the Environmental Monitoring Plan (EMP), or the Environmental Assessment (EA) (DOE/EA-1029).

4.5.1 Vapor Waste Streams

The primary waste stream from the process is the unreacted syngas (Stream 148) which is sent to the Eastman boilers for heat recovery. The second largest stream are vapors that are flashed or stripped from the product methanol (Stream 19) together these comprise the block originating in Product Purification and designated as “Vapors to existing boilers”.

The EA identifies the following vapor emissions from all sources. These are primarily fugitive emissions from the pumps, valves, connectors, compressor seals, and pressure relief devices within the LPMEOH™ process.

	Tons/Year
VOC (Volatile Organic Compound)	7.3
CO (Carbon Monoxide)	2.1
H_2S/SO_2 (Hydrogen sulfide/sulfur dioxide)	0
H_2 (Hydrogen)	0.42
NH_3 (Ammonia)	0
TSP (Particulate Matter)	1.0

4.5.2 Liquid Waste Streams

The process generates an additional load of approximately 1,150 gallons/day of waste containing 4,180 lbs/day BOD at the Eastman Waste Water Treatment Facility.

4.5.3 Solid Waste Streams

The Spent Catalysts/Oil mix from the demonstration unit is drummed and sent to a metals reclaimer. Approximately 140,000 lbs/yr of mixture containing approximately 50% by weight of solid catalyst is expected.

4.6 EQUIPMENT LIST

Table 4-2 contains the Major Process Equipment List organized by Major Processing Area.

Table 4-1 Heat and Material Balance Summary

CASE:	L:\ kingsprt\ hmb\ 1feb95\ design							
Stream No.		10	109	120	148	149	20	204
Press	PSI	875.0	785.0	735.0	722.0	722.0	765.0	85.0
Temp	C	37.8	169.5	121.1	40.5	40.5	37.8	40.1
Temp	F	100.0	337.1	249.9	105.0	105.0	100.0	104.2
Lbmol/hr								
H2		2.636	4402.484	2967.899	252.163	2712.126	77.996	3.610
CO		127.846	1818.957	1121.754	95.167	1023.566	11.594	3.021
N2		1.318	293.947	294.054	24.922	268.047	12.648	1.084
CO2		0.000	654.744	635.137	51.831	557.470	3.162	25.836
MEOH		0.000	42.496	744.457	3.952	42.504	0.000	698.001
DME		0.000	3.790	5.057	0.353	3.796	0.000	0.909
H2O		0.000	0.448	21.642	0.042	0.448	0.000	21.152
ETOH		0.000	0.016	0.447	0.002	0.016	0.000	0.429
C3OH		0.000	0.001	0.065	0.000	0.001	0.000	0.063
C4OH		0.000	0.000	0.026	0.000	0.000	0.000	0.026
IBOH		0.000	0.001	0.087	0.000	0.001	0.000	0.086
C5OH		0.000	0.000	0.018	0.000	0.000	0.000	0.018
MEAC		0.000	0.065	0.242	0.006	0.065	0.000	0.171
MEFM		0.000	1.256	2.790	0.117	1.259	0.000	1.415
C1		0.000	12.228	13.449	1.137	12.234	0.000	0.077
C2		0.000	3.549	3.982	0.331	3.557	0.000	0.095
C3		0.000	1.007	1.171	0.094	1.010	0.000	0.067
OIL		0.000	0.000	0.032	0.000	0.000	0.000	0.032
Total molar flow		131.8	7235.0	5812.3	430.1	4626.1	105.4	756.1
Total mass flow	lb/hr	3,623	98,846	98,686	6,337	68,154	975	24,195
Enthalpy	MMBtu/hr	-6.083	-188.288	-221.879	-13.674	-147.071	-1.066	-79.011
mol%								
H2		2.000	60.850	51.062	58.627	58.627	74.000	0.477
CO		97.000	25.141	19.300	22.126	22.126	11.000	0.400
N2		1.000	4.063	5.059	5.794	5.794	12.000	0.143
CO2		0.000	9.050	10.927	12.051	12.051	3.000	3.417
MEOH		0.000	0.587	12.808	0.919	0.919	0.000	92.317
DME		0.000	0.052	0.087	0.082	0.082	0.000	0.120
H2O		0.000	0.006	0.372	0.010	0.010	0.000	2.797
ETOH		0.000	0.000	0.008	0.000	0.000	0.000	0.057
C3OH		0.000	0.000	0.001	0.000	0.000	0.000	0.008
C4OH		0.000	0.000	0.000	0.000	0.000	0.000	0.003
IBOH		0.000	0.000	0.001	0.000	0.000	0.000	0.011
C5OH		0.000	0.000	0.000	0.000	0.000	0.000	0.002
MEAC		0.000	0.001	0.004	0.001	0.001	0.000	0.023
MEFM		0.000	0.017	0.048	0.027	0.027	0.000	0.187
C1		0.000	0.169	0.231	0.264	0.264	0.000	0.010
C2		0.000	0.049	0.069	0.077	0.077	0.000	0.013
C3		0.000	0.014	0.020	0.022	0.022	0.000	0.009
OIL		0.000	0.000	0.001	0.000	0.000	0.000	0.004
Total		100	100	100	100	100	100	100
Vapor Fraction		1.000	1.000	1.000	1.000	1.000	1.000	0.025
Liquid Fraction		0.000	0.000	0.000	0.000	0.000	0.000	0.975
Mol. Wt. Mix		27.490	13.662	16.979	14.733	14.733	9.255	32.000
Mol. Wt. Vapor		27.490	13.662	16.979	14.733	14.733	9.255	32.286
Mol. Wt. Liquid								31.992
Density	LB/CUFT	4.092	1.206	1.621	1.701	1.701	1.114	13.386
Vapor Density	LB/CUFT	4.092	1.206	1.621	1.701	1.701	1.114	0.461
Liquid Density	LB/CUFT							48.981

Table 4-1 Heat and Material Balance Summary

CASE:	L:\kingsprt\hmb\1feb95\design				
Stream No.		213	214	242	30
Press	PSI	45.0	14.8	80.0	790.0
Temp	C	40.7	40.6	100.8	37.8
Temp	F	105.3	105.1	213.4	100.0
Lbmol/hr					
H2		3.610	0.000	0.000	1608.013
CO		3.021	0.000	0.000	656.961
N2		1.084	0.000	0.000	11.859
CO2		25.836	0.000	0.000	94.868
MEOH		3.601	520.901	173.499	0.000
DME		0.909	0.000	0.000	0.000
H2O		0.013	0.054	21.084	0.000
ETOH		0.000	0.043	0.386	0.000
C3OH		0.000	0.000	0.063	0.000
C4OH		0.000	0.000	0.026	0.000
IBOH		0.000	0.000	0.085	0.000
C5OH		0.000	0.000	0.018	0.000
MEAC		0.170	0.002	0.000	0.000
MEFM		1.415	0.000	0.000	0.000
C1		0.077	0.000	0.000	0.000
C2		0.095	0.000	0.000	0.000
C3		0.067	0.000	0.000	0.000
OIL		0.000	0.000	0.032	0.000
Total molar flow		39.9	521.0	195.2	2371.7
Total mass flow	lb/hr	1,521	16,694	5,980	26,150
Enthalpy	MMBtu/hr	-5.147	-53.303	-19.938	-46.877
mol%					
H2		9.048	0.000	0.000	67.800
CO		7.572	0.000	0.000	27.700
N2		2.718	0.000	0.000	0.500
CO2		64.755	0.000	0.000	4.000
MEOH		9.026	99.981	88.885	0.000
DME		2.278	0.000	0.000	0.000
H2O		0.033	0.010	10.802	0.000
ETOH		0.001	0.008	0.198	0.000
C3OH		0.000	0.000	0.033	0.000
C4OH		0.000	0.000	0.013	0.000
IBOH		0.000	0.000	0.044	0.000
C5OH		0.000	0.000	0.009	0.000
MEAC		0.425	0.000	0.000	0.000
MEFM		3.546	0.000	0.000	0.000
C1		0.193	0.000	0.000	0.000
C2		0.237	0.000	0.000	0.000
C3		0.168	0.000	0.000	0.000
OIL		0.000	0.000	0.017	0.000
Total		100	100	100	100
Vapor Fraction		1.000	0.000	0.000	1.000
Liquid Fraction		0.000	1.000	1.000	0.000
Mol. Wt. Mix		38.132	32.042	30.634	11.026
Mol. Wt. Vapor		-0.003			11.026
Mol. Wt. Liquid			32.042	30.634	
Density	LB/CUFT		48.307	45.200	1.377
Vapor Density	LB/CUFT				1.377
Liquid Density	LB/CUFT		48.307	45.200	

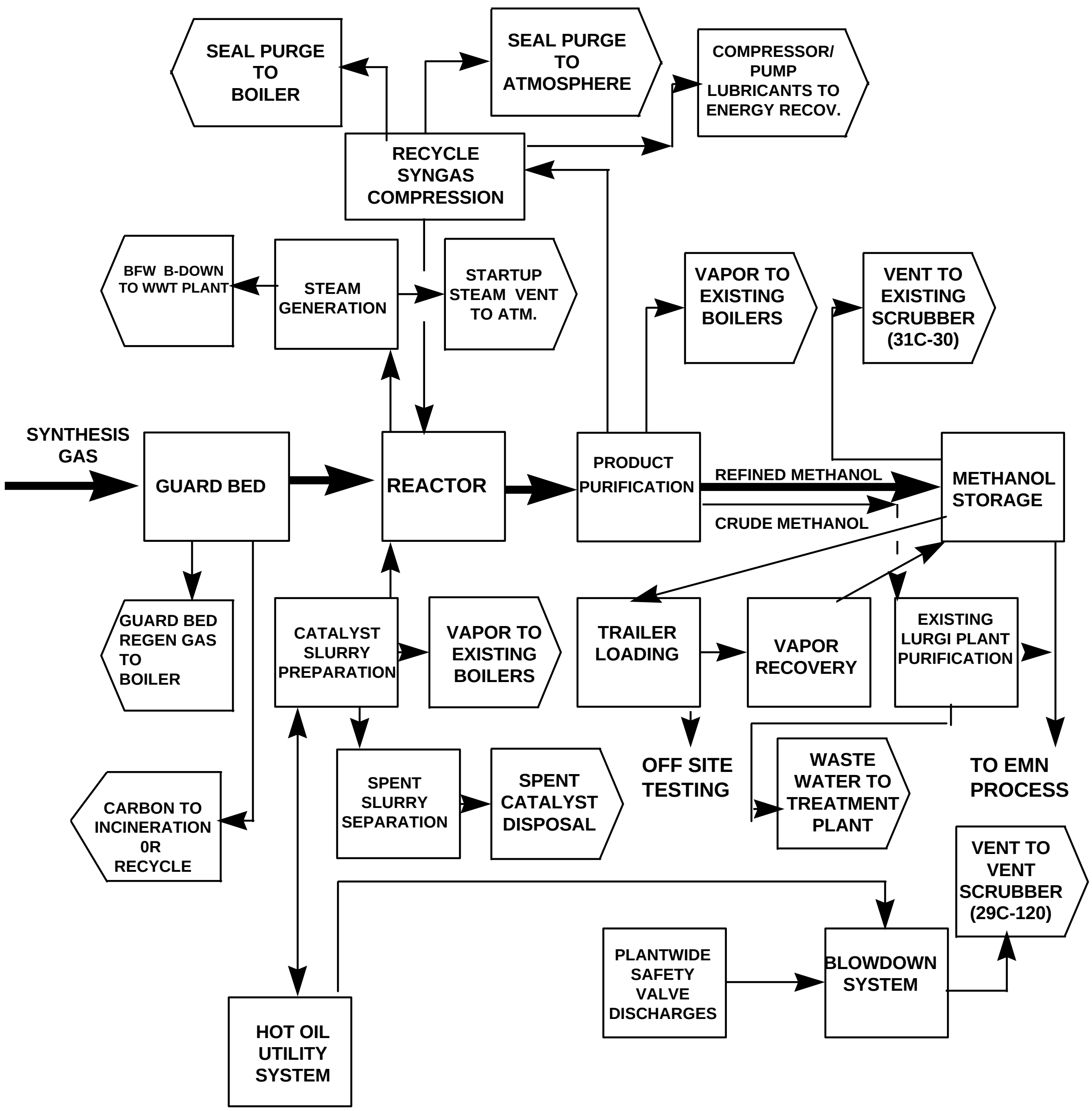


FIGURE 4-3 LPMEOH^{TM} DEMONSTRATION UNIT PROCESS AND WASTE STREAMS

FSF July 1997

Item #	Item Description / Type	Number In Use	Number Spare	Capacity / Size	Design Characteristics	Construction Materials	Vendor
				FEED GAS PURIFICATION			
29C-40	Carbonyl Guard Bed Vertical Vessel	1		Diameter/ft: 4.0 T-T Height/ft: 15.0 Weight/lbs: 24,250 Bed Volume/cu ft: 151	Operating Temperature: 40-350 Deg F Operating Pressure: 25-775 PSIG Design Temperature: 450 Deg F Design Pressure: 1000 PSIG	CS Screens: 304L SS / 316SS Bed: Activated Carbon Supports: Ceramic Balls	Arrow Tank Cambridge, MN
29E-40	Guard Bed Regeneration Heater Electric	1		Diameter/ft: 0.90 Shell Length/ft: 11.0 Bundle Length/ft: 9.3 Duty: 169 KW Weight/lbs: 875	Operating Temperature: (-2)-385 Deg F Operating Pressure: 80 PSIG Design Temperature: 1000 Deg F Design Pressure: 600 PSIG	CS Heater Elements are Incaloy 800	Watlow Process Sys Troy, MO
29Y-01A/B	Fresh Feed Syngas Filters Cartridge	2		Diameter/ft: 1.0 Length/ft: 7.5 Weight/lbs: 1,400 Flow Rate: 31992 lb/hr	Operating Temperature: 100-150 Deg F Operating Pressure: 775 PSIG Design Temperature: 250 Deg F Design Pressure: 1000 PSIG	CS Internals: 304L SS Filter Media: Polyester	Consler Honeoya Falls, NY
				COMPRESSION			
29C-07	Compressor Knockout Separator Vertical K. O. Drum	1		Diameter/ft: 4.5 T-T Height/ft: 8.5 Weight/lbs: 21,300 Volume/gal: 1,190	Operating Temperature: 100 Deg F Operating Pressure: 700-900 PSIG Design Temperature: 250 Deg F Design Pressure: 1000 PSIG	CS Demister: 304 SS	Arrow Tank Cambridge, MN
29E-01	Compressor Recycle Cooler - Shell Horizontal Shell & Tube	1		Diameter/ft: 2.0 Length/ft: 23.6 Weight/lbs: 9,680	Operating Temperature: 105-141 Deg F Operating Pressure: 800 PSIG Design Temperature: 250 Deg F Design Pressure: 1000 PSIG	CS	Atlas Industrial Clifton, NJ
	Compressor Recycle Cooler - Tubes	1		Surface Area/sq ft: 935 Duty: 1.92 mm BTU/Hr	Operating Temperature: 85 Deg F Operating Pressure: 75 PSIG Design Temperature: 250 Deg F Design Pressure: 150 PSIG	304L SS	
29K-01	Syngas Recycle Compressor Axial-Inlet, Single-Stage, Centrifugal	1		Flow Rate: 4731 Mole/Hr Power: 520 BHP Motor: 700 HP	Operating Temperature: 143 Deg F Inlet Pressure: 722 PSIA Outlet Pressure: 790 PSIA Design Temperature: 250 Deg F Design Pressure: 1000 PSIG	CS	Mannesmann-Demag Germany

Item #	Item Description / Type	Number In Use	Number Spare	Capacity / Size	Design Characteristics	Construction Materials	Vendor
	REACTOR LOOP AND CATALYST REDUCTION						
29C-01	LPMEOHTM Reactor - Shell Gas Sparged Reactor with Internal Heat Exchanger	1		Diameter/ft: 7.5 T-T Height/ft: 68.25 Weight/lbs: 268,000 Volume/gal: 23,383	Operating Temperature: 482 Deg F Operating Pressure: 743 PSIA Design Temperature: 600 Deg F Design Pressure: 1000 PSIG	304L SS Clad on CS	Joseph Oat Camden, NJ
	LPMEOHTM Reactor - Heat Exchanger	1		Surface Area/sq ft: 2053 Pipe Length/ft: 57.7 Pipe OD/In: 1.9	Operating Temperature: 376 Deg F Operating Pressure: 180-335 PSIG Design Temperature: 600 Deg F Design Pressure: 640 PSIG	Duplex 2205	
29C-02	Steam Drum Horizontal Vessel	1		Diameter/ft: 4.5 Length/ft: 12.0 Weight/lbs: 25,000 Volume/gal: 1,605	Operating Temperature: 375-430 Deg F Operating Pressure: 180-335 PSIG Design Temperature: 750 Deg F Design Pressure: 640 PSIG	CS	ABCO Abilene, TX
29C-03	H. P. Methanol Separator Vertical Vessel	1		Diameter/ft: 4.0 T-T Height/ft: 9.42 Weight/lbs: 17,640 Volume/gal: 1,128	Operating Temperature: 85-105 Deg F Operating Pressure: 750 PSIG Design Temperature: 600 Deg F Design Pressure: 1000 PSIG	304L SS Clad on CS Demister: 304L SS	Struthers Gulfport, MS
29C-05	Secondary Oil Knockoout Vessel Vertical Vessel	1		Diameter/ft: 6.0 T-T Height/ft: 11.42 Weight/lbs: 40,600 Volume/gal: 2,834	Operating Temperature: 300 Deg F Operating Pressure: 750 PSIG Design Temperature: 600 Deg F Design Pressure: 1000 PSIG	304L SS Clad on CS Demister: 304L SS	Struthers Gulfport, MS
29C-06	Reactor Cyclone Vertical Whirl-A-Way Separator	1		Diameter/ft: 1.2 T-T Height/ft: 17.00 Weight/lbs: 3,000 Volume/gal: 179	Operating Temperature: 100-550 Deg F Operating Pressure: 700-900 PSIG Design Temperature: 600 Deg F Design Pressure: 1000 PSIG	304L SS	Dynatherm Houston, TX
29C-30	Catalyst Reduction Vessel Jacketed Vertical Tank w/Agitator	1		Diameter/ft: 4.0 T-T Height/ft: 19.8 Weight/lbs: 10,000 Volume/cu ft: 265	Operating Temperature: 60-464 Deg F Operating Pressure: 20-100 PSIG Design Temperature: 600 Deg F Design Pressure: 150 PSIG	304L SS	Four Corporation Green Bay, WI

Item #	Item Description / Type	Number In Use	Number Spare	Capacity / Size	Design Characteristics	Construction Materials	Vendor
29C-31	Reduction Condensate Accumulator Vertical Vessel	1		Diameter/ft: 3.5 T-T Height/ft: 9.5 Weight/lbs: 3,350 Volume/gal: 763	Operating Temperature: 60-464 Deg F Operating Pressure: 20-100 PSIG Design Temperature: 600 Deg F Design Pressure: 150 PSIG	CS (Packing: 304L SS)	Arrow Tank Cambridge, MN
29C-32	Utility Oil Surge Tank (V-01 Skid) Horizontal Vessel	1		Diameter/ft: 2.5 T-T Height/ft: 5.5 Weight/lbs: 650 Volume/gal: 234	Operating Temperature: (-10)-550 Deg F Operating Pressure: 25-80 PSIG Design Temperature: 600 Deg F Design Pressure: 100 PSIG	CS	HEAT, Inc. Carnegie, PA
29E-02	Syngas Feed/Product Economizer - Shell Horizontal Shell & Tube	1		Diameter/ft: 2.5 Length/ft: 27 Weight/lbs: 30,000	Operating Temperature: 116-402 Deg F Operating Pressure: 775 PSIG Design Temperature: 600 Deg F Design Pressure: 1000 PSIG	304L SS	Southern Heat Exch Tuscaloosa, AL
	Syngas Feed/Product Economizer - Tubes			Surface Area/sq ft: 2245 Duty: 13.13 mm BTU/Hr	Operating Temperature: 482-250 Deg F Operating Pressure: 728 PSIG Design Temperature: 600 Deg F Design Pressure: 1000 PSIG	304L SS	
29E-03	MEOH Product Air Cooled Condenser Fan-Cooled	1		Overall Dim/ft: 42 x 16 x 13 Weight/lbs: 51,630 Finned Area/sq ft: 65,594 Bare Area/sq ft: 3,049 Duty: 15.82 mm BTU/Hr Motor Size/HP: 60	Operating Temperature: 250-140 Deg F Operating Pressure: 720 PSIG Design Temperature: 600 Deg F Design Pressure: 1000 PSIG	304L SS w/aluminum extruded fins	Cooling Products Tulsa, OK
29E-04	MEOH Product C.W. Condenser - Shell Shell & Tube	1		Diameter/ft: 1.8 Length/ft: 29.5 Weight/lbs: 8,730	Operating Temperature: 85-105 Deg F Operating Pressure: 75 PSIG Design Temperature: 250 Deg F Design Pressure: 150 PSIG	CS Head: 304L SS	Atlas Industrial Clifton, NJ
	MEOH Product C.W. Condenser - Tube	1		Surface Area/sq ft: 971 Duty: 2.952 mm BTU/Hr	Operating Temperature: 140-105 Deg F Operating Pressure: 715 PSIG Design Temperature: 600 Deg F Design Pressure: 1000 PSIG	304L SS	

Item #	Item Description / Type	Number In Use	Number Spare	Capacity / Size	Design Characteristics	Construction Materials	Vendor
29E-31	Reduction Vessel Ovhd Condenser - Shell Shell & Tube	1		Diameter/ft: 0.6 Length/ft: 10.7 Weight/lbs: 750	Operating Temperature: 70-250 Deg F Operating Pressure: 95 PSIG Design Temperature: 600 Deg F Design Pressure: 150 PSIG	304L SS	Atlas Industrial Clifton, NJ
	Reduction Vessel Ovhd Condenser - Tube	1		Surface Area/sq ft: 76 Duty: .222 mm BTU/Hr	Operating Temperature: 85-105 Deg F Operating Pressure: 75 PSIG Design Temperature: 250 Deg F Design Pressure: 150 PSIG	304L SS	
:9E-32 A/B/C	Utility Oil Heater (V-01 Skid) Electric	3		Diameter/ft: 1.0 Shell Length/ft: 5.5 Duty: 360 KW	Operating Temperature: 502-524 Deg F Operating Pressure: 110 PSIG Design Temperature: 600 Deg F Design Pressure: 150 PSIG	CS Heater Elements are SS	HEAT, Inc. Carnegie, PA
29E-33	Utility Oil Cooler (V-01 Skid) - Shell Shell & Tube	1		Diameter/ft: 1.0 Length/ft: 6.2	Operating Temperature: 180-170 Deg F Operating Pressure: 85 PSIG Design Temperature: 600 Deg F Design Pressure: 150 PSIG	CS	HEAT, Inc. Carnegie, PA
	Utility Oil Cooler (V-01 Skid) - Tube	1		Surface Area/sq ft: 30.5 Duty: .310 mm BTU/Hr	Operating Temperature: 85-105 Deg F Operating Pressure: 85 PSIG Design Temperature: 250 Deg F Design Pressure: 150 PSIG	CS	
29E-34	Seal Oil Cooler Horizontal Shell & Tube	1		Diameter/ft: 0.75 Length/ft: 19.0 Weight/lbs: 2,000	Operating Temperature: 100-121 Deg F Operating Pressure: 6 PSIG Design Temperature: 250 Deg F Design Pressure: 150 PSIG	CS	Aurora Industrial Orchard Park, NY
	Seal Oil Cooler - Tube	1		Surface Area/sq ft: 529 Duty: 0.32 mm BTU/Hr	Operating Temperature: 85-98 Deg F Operating Pressure: 86 PSIG Design Temperature: 250 Deg F Design Pressure: 150 PSIG	304L SS	

Item #	Item Description / Type	Number In Use	Number Spare	Capacity / Size	Design Characteristics	Construction Materials	Vendor
29G-01A/B	Condensed Oil Circulation Pump Twin Screw Positive Displacement	1	1	Flow Rate: 7 GPM Power: 7.1 BHP Motor: 10 HP	Operating Temperature: 220-285 Deg F Inlet Pressure: 744 PSIA Outlet Pressure: 988.5 PSIA Design Temperature: 600 Deg F Design Pressure: 1095 PSIG	CS	Ingersoll-Dresser Brampton, Ontario
29G-02	Slurry Return Pump Twin Screw Positive Displacement	1		Flow Rate: 36 GPM Power: 7 BHP Motor: 15 HP	Operating Temperature: 150-500 Deg F Inlet Pressure: 22 PSIA Outlet Pressure: 161.7 PSIA Design Temperature: 600 Deg F Design Pressure: 200 PSIG	CS	Ingersoll-Dresser Brampton, Ontario
29G-03A/B	Oil Makeup Pump Twin Screw Positive Displacement	1	1	Flow Rate: 45 GPM Power: 63.7 BHP Motor: 125 HP	Operating Temperature: 50-100 Deg F Inlet Pressure: 16 PSIA Outlet Pressure: 1112 PSIA Design Temperature: 150 Deg F Design Pressure: 1095 PSIG	CS	Ingersoll-Dresser Brampton, Ontario
29G-04A/B	Boiler Feed Water Pump Centrifugal	1	1	Flow Rate: 84 GPM Power: 41.4 BHP Motor: 60 HP	Operating Temperature: 260 Deg F Inlet Pressure: 275 PSIA Outlet Pressure: 765.4 PSIA Design Temperature: 260 Deg F Design Pressure: 1000 PSIG	CS	Sulzer Bingham Portland, OR
29G-30	Slurry Transfer Pump Reciprocating	1		Flow Rate: 30 GPM Power: 17.5 BHP Motor: 25 HP	Operating Temperature: 464 Deg F Inlet Pressure: 76 PSIA Outlet Pressure: 1076.5 PSIA Design Temperature: 600 Deg F Design Pressure: 1095 PSIG	CS	Ingersoll-Dresser Phillipsburg, NJ
29G-32	Utility Oil Circulating Pump (V-01 Skid) Magnetic Drive Centrifugal	1		Flow Rate: 150 GPM Power: 13 BHP Motor: 30 HP	Operating Temperature: (-10)-550 Deg F Inlet Pressure: 59 PSIA Outlet Pressure: 146.5 PSIA Design Temperature: 600 Deg F Design Pressure: 150 PSIG	CS	Dickow Pump Marietta, GA

Item #	Item Description / Type	Number In Use	Number Spare	Capacity / Size	Design Characteristics	Construction Materials	Vendor
29G-34	Oil Feed Pump Durco Centrifugal	1		Flow Rate: 30 GPM Power: 15 BHP Motor: 30 HP	Operating Temperature: 50-100 Deg F Inlet Pressure: 16.4 PSIA Outlet Pressure: 154.7 PSIA Design Temperature: 150 Deg F Design Pressure: 200 PSIG	CS	ABE Berkleigh Dayton, OH
29SP-001	BFW Blowdown Sample Cooler Heliflow Heat Exchanger	1		Flow Rate: 1 GPM	Operating Temperature: 381-105 Deg F Operating Pressure: 182 PSIG Design Temperature: 600 Deg F Design Pressure: 640 PSIG	304L SS	Graham Mfg Batavia, NY
29Y-30	Catalyst Reduction Agitator	1		Motor: 5 HP	Operating Temperature: 60-464 Deg F Operating Pressure: 150 PSIG Design Temperature: 600 Deg F Design Pressure: 150 PSIG	304 SS	Chemineer Dayton, OH
29Y-35A/B	Seal Oil Filters Cartridge	2		Diameter/ft: 0.5 Length/ft: 5.2 Weight/lbs: 375 Flow Rate: 25 GPM	Operating Temperature: 50-100 Deg F Operating Pressure: 1100 PSIG Design Temperature: 150 Deg F Design Pressure: 1095 PSIG	CS Filter Media: Polyester	Consler Honeoya Falls, NY
				DISTILLATION			
29C-10	Methanol Stabilizer Column Distillation Column	1		Diameter/ft: 3.5 Height/ft: 65 Weight/lbs: 34,000 Sump Volume/gal: 741.4	Operating Temperature: (-10)-210 Deg F Operating Pressure: 0-35 PSIG Design Temperature: 400 Deg F Design Pressure: 60 PSIG	CS Packing: 304 SS	Modern Welding Owensboro, KY
29C-11	Methanol Stabilizer Reflux Drum Vertical Vessel	1		Diameter/ft: 3.5 T-T Height/ft: 6.0 Weight/lbs: 2,000 Volume/gal: 514	Operating Temperature: (-10)-105 Deg F Operating Pressure: 0-25 PSIG Design Temperature: 250 Deg F Design Pressure: 60 PSIG	CS	Arrow Tank Cambridge, MN
29C-12	Methanol Stabilizer Feed Drum Vertical Vessel	1		Diameter/ft: 8.0 T-T Height/ft: 20.0 Weight/lbs: 23,450 Volume/gal: 8,525	Operating Temperature: 105 Deg F Operating Pressure: 70 PSIG Design Temperature: 250 Deg F Design Pressure: 120 PSIG	CS	Arrow Tank Cambridge, MN

Item #	Item Description / Type	Number In Use	Number Spare	Capacity / Size	Design Characteristics	Construction Materials	Vendor
29C-13	Stabilizer Condensate Pot Vertical Vessel	1		Diameter/ft: 1.5 T-T Height/ft: 2.5 Weight/lbs: 600 Volume/gal: 39	Operating Temperature: (-10)-338 Operating Pressure: ATM-100 Design Temperature: 450 Deg F Design Pressure: 125 PSIG	CS	Arrow Tank Cambridge, MN
29C-20	Methanol Rectifier Column Distillation Column	1		Diameter/ft: 5.0 Height/ft: 80.0 Weight/lbs: 65,000 Sump Volume/gal: 1577	Operating Temperature: (-10)-215 Deg F Operating Pressure: 0-35 PSIG Design Temperature: 400 Deg F Design Pressure: 60 PSIG	CS Packing: 304 SS	Modern Welding Owensboro, KY
29C-21	Methanol Rectifier Reflux Drum Vertical Vessel			Diameter/ft: 4.0 T-T Height/ft: 7.0 Weight/lbs: 2,930 Volume/gal: 784	Operating Temperature: (-10)-105 Deg F Operating Pressure: 0-20 PSIG Design Temperature: 250 Deg F Design Pressure: 60 PSIG	CS	Arrow Tank Cambridge, MN
29C-23	Rectifier Condensate Pot Vertical Vessel	1		Diameter/ft: 2.0 T-T Height/ft: 3.0 Weight/lbs: 700 Volume/gal: 86	Operating Temperature: (-10)-338 Operating Pressure: ATM-100 Design Temperature: 450 Deg F Design Pressure: 125 PSIG	CS	Arrow Tank Cambridge, MN
29D-20	Methanol Lot Tank API Tank	1		Diameter/ft: 16.0 Height/ft: 25.0 Weight/lbs: 20,000 Volume/gal: 30,000	Operating Temperature: (-10)-105 Deg F Operating Pressure: ATM Design Temperature: 250 Deg F Design Pressure: 2 PSIG	CS	Brown-Minneapolis Birmingham, AL
29D-21	Methanol Lot Tank API Tank	1		Diameter/ft: 16.0 Height/ft: 25.0 Weight/lbs: 20,000 Volume/gal: 30,000	Operating Temperature: (-10)-105 Deg F Operating Pressure: ATM Design Temperature: 250 Deg F Design Pressure: 2 PSIG	CS	Brown-Minneapolis Birmingham, AL
29D-60	Caustic Mix Tank Vertical Vessel	1		Diameter/ft: 4.0 T-T Height/ft: 4.0 Weight/lbs: 797 Volume/gal: 376	Operating Temperature: (-10)-105 Deg F Operating Pressure: ATM Design Temperature: 180 Deg F Design Pressure: 0 PSIG	CS	Industrial Piping Pineville, NC

Item #	Item Description / Type	Number In Use	Number Spare	Capacity / Size	Design Characteristics	Construction Materials	Vendor
29E-10	Methanol Stabilizer Reboiler - Shell Thermosiphon	1		Diameter/ft: 2.2 Length/ft: 20.0 Weight/lbs: 6,900	Operating Temperature: 338 Deg F Operating Pressure: 100 PSIG Design Temperature: 450 Deg F Design Pressure: 125 PSIG	CS 304L SS Band	Southern Heat Exch Tuscaloosa, AL
	Methanol Stabilizer Reboiler - Tube	1		Surface Area/sq ft: 664 Duty: 10.70 mm BTU/Hr	Operating Temperature: 215 Deg F Operating Pressure: 40 Design Temperature: 450 Deg F Design Pressure: 60 PSIG	CS	
29E-11	Methanol Stabilizer Condenser - Shell Shell & Tube	1		Diameter/ft: 1.8 Length/ft: 23.3 Weight/lbs: 7,240	Operating Temperature: 199-105 Deg F Operating Pressure: 30 PSIG Design Temperature: 400 Deg F Design Pressure: 60 PSIG	CS	Atlas Industrial Clifton, NJ
	Methanol Stabilizer Condenser - Tube	1		Surface Area/sq ft: 1351 Duty: 8.35 mm BTU/Hr	Operating Temperature: 85-105 Deg F Operating Pressure: 75 PSIG Design Temperature: 250 Deg F Design Pressure: 150 PSIG	304L SS	
29E-20	Methanol Rectifier Reboiler - Shell Thermosiphon	1		Diameter/ft: 2.8 Length/ft: 21.3 Weight/lbs: 12,900	Operating Temperature: 338 Deg F Operating Pressure: 100 PSIG Design Temperature: 450 Deg F Design Pressure: 125 PSIG	CS 304L SS Band	Southern Heat Exch Tuscaloosa, AL
	Methanol Rectifier Reboiler - Tube	1		Surface Area/sq ft: 1279 Duty: 23.04 mm BTU/Hr	Operating Temperature: 220 Deg F Operating Pressure: 41 Design Temperature: 450 Deg F Design Pressure: 60 PSIG	CS	
29E-21	Methanol Rectifier Air Cooler Fan-Cooled	1		Overall Dim/ft: 45 x 14 x 12 Weight/lbs: 41,500 Finned Area/sq ft: 94,429 Bare Area/sq ft: 4,384 Duty: 21.86 mm BTU/Hr Motor Size/HP: 50	Operating Temperature: 205-140 Deg F Operating Pressure: 30 PSIG Design Temperature: 400 Deg F Design Pressure: 60 PSIG	CS	Cooling Products Tulsa, OK

Item #	Item Description / Type	Number In Use	Number Spare	Capacity / Size	Design Characteristics	Construction Materials	Vendor
29E-22	MEOH Rectifier C.W. Condenser - Shell Shell & Tube	1		Diameter/ft: 1.0 Length/ft: 21.0 Weight/lbs: 2,710	Operating Temperature: 140-105 Deg F Operating Pressure: 25 PSIG Design Temperature: 250 Deg F Design Pressure: 60 PSIG	CS	Atlas Industrial Clifton, NJ
	MEOH Rectifier C.W. Condenser - Tube	1		Surface Area/sq ft: 368 Duty: 1.01 mm BTU/Hr	Operating Temperature: 85-105 Deg F Operating Pressure: 75 PSIG Design Temperature: 250 Deg F Design Pressure: 150 PSIG	304L SS	
29E-23	Crude Methanol Cooler - Shell Shell & Tube	1		Diameter/ft: 1.0 Length/ft: 25.0 Weight/lbs: 3,350	Operating Temperature: 210-105 Deg F Operating Pressure: 75 PSIG Design Temperature: 250 Deg F Design Pressure: 200 PSIG	CS	Atlas Industrial Clifton, NJ
	Crude Methanol Cooler - Tube	1		Surface Area/sq ft: 432 Duty: 1.92 mm BTU/Hr	Operating Temperature: 85-105 Deg F Operating Pressure: 75 PSIG Design Temperature: 250 Deg F Design Pressure: 150 PSIG	CS	
29G-10A/B	Methanol Stabilizer Underflow Pump Durco Centrifugal	1	1	Flow Rate: 77 GPM Power: 4.5 BHP Motor: 10 HP	Operating Temperature: 210 Deg F Inlet Pressure: 51 PSIA Outlet Pressure: 91 PSIA Design Temperature: 250 Deg F Design Pressure: 200 PSIG	CS	ABE Berkleigh Dayton, OH
29G-11A/B	Methanol Stabilizer Reflux Pump Durco Centrifugal	1	1	Flow Rate: 55 GPM Power: 5 BHP Motor: 10 HP	Operating Temperature: 105 Deg F Inlet Pressure: 43 PSIA Outlet Pressure: 120.6 PSIA Design Temperature: 250 Deg F Design Pressure: 150 PSIG	CS	ABE Berkleigh Dayton, OH
29G-20A/B	Methanol Rectifier Underflow Pump Durco Centrifugal	1	1	Flow Rate: 24 GPM Power: 12 BHP Motor: 20 HP	Operating Temperature: 216 Deg F Inlet Pressure: 55 PSIA Outlet Pressure: 185.9 PSIA Design Temperature: 250 Deg F Design Pressure: 150 PSIG	CS	ABE Berkleigh Dayton, OH

Item #	Item Description / Type	Number In Use	Number Spare	Capacity / Size	Design Characteristics	Construction Materials	Vendor
29G-21A/B	Methanol Rectifier Reflux Pump Durco Centrifugal	1	1	Flow Rate: 122 GPM Power: 14 BHP Motor: 25 HP	Operating Temperature: 105 Deg F Inlet Pressure: 36 PSIA Outlet Pressure: 145.7 PSIA Design Temperature: 250 Deg F Design Pressure: 200 PSIG	CS	ABE Berkleigh Dayton, OH
29G-23A/B	Methanol Transfer Pump Durco Centrifugal	1	1	Flow Rate: 180 GPM Power: 6.5 BHP Motor: 7.5 HP	Operating Temperature: 105 Deg F Inlet Pressure: 14.1 PSIA Outlet Pressure: 53.3 PSIA Design Temperature: 250 Deg F Design Pressure: 200 PSIG	CS	ABE Berkleigh Dayton, OH
29G-60A/B	Caustic Metering Pumps Diaphragm	2		Flow Rate: 7 GPH Power: 1.0 BHP Motor: 1.0 HP	Operating Temperature: 105 Deg F Inlet Pressure: 14.1 PSIA Outlet Pressure: 90.1 PSIA Design Temperature: 250 Deg F Design Pressure: 200 PSIG	PVC	Neptune Chem Pump Lansdale, PA
29SP-002	Stabilizer Col Underflow Sample Cooler Heliflow Heat Exchanger	1		Flow Rate: 1 GPM	Operating Temperature: 212-105 Deg F Operating Pressure: 60 PSIG Design Temperature: 250 Deg F Design Pressure: 200 PSIG	304L SS	Graham Mfg Batavia, NY
29SP-003	Rectifier Col Underflow Sample Cooler Heliflow Heat Exchanger	1		Flow Rate: 1 GPM	Operating Temperature: 216-105 Deg F Operating Pressure: 171 PSIG Design Temperature: 250 Deg F Design Pressure: 200 PSIG	304L SS	Graham Mfg Batavia, NY
29Y-10	Methanol Product Filter Cartridge	1		Diameter/ft: 0.7 Length/ft: 5 Weight/lbs: 300 Flow Rate: 22673 lb/hr	Operating Temperature: 209-214 Deg F Operating Pressure: 80-88 PSIG Design Temperature: 250 Deg F Design Pressure: 200 PSIG	CS Filter Media: Polyester	Consler Honeoya Falls, NY
29Y-60	Caustic Tank Agitator	1		Motor: 1 HP	Operating Temperature: 105 Deg F Operating Pressure: ATM Design Temperature: 250 Deg F Design Pressure: 2 PSIG	CS	Chemineer Dayton, OH

Item #	Item Description / Type	Number In Use	Number Spare	Capacity / Size	Design Characteristics	Construction Materials	Vendor
	STORAGE AND MISCELLANEOUS						
29C-50	Oil / Water Separator Coalescing Tank	1		Dimensions: 11.5' x 9' x 5' Weight/lbs: 8420 Design Capacity: 355 gpm Hold Tank Vol: 1000 Gal	Operating Temperature: 60 Deg F Design Temperature: 150 Deg F	CS Plates: Non-Metallic	Facet International Tulsa, OK
29C-60	Air Receiver (V-02 Skid) Vertical Vessel	1		Diameter/ft: 2.5 T-T Height/ft: 7.0 Volume/gal: 240	Design Temperature: 450 Deg F Design Pressure: 200 PSIG	CS	Henderson Engin Mt. Laurel, NJ
29C-120	Vent Stack Vertical Stack	1		Diameter/ft: Bottom Stack: 11 Middle Stack: 8 Upper Stack: 4 Vent Tip: 2 Weight/lbs: 104,000	Operating Temperature: 482 Deg F Operating Pressure: 30 PSIG Design Temperature: 600 Deg F Design Pressure: 50 PSIG	CS	John Zink Tulsa, OK
29D-01	Safety Relief Knockout Drum Vertical Vessel	1		Diameter/ft: 6.0 T-T Height/ft: 17.0 Weight/lbs: 9,500 Volume/gal: 4,019	Operating Temperature: (-10)-482 Deg F Operating Pressure: 0-45 PSIG Design Temperature: 600 Deg F Design Pressure: 55 PSIG	CS	Arrow Tank Cambridge, MN
29D-02	Slurry Tank Vertical Vessel	1		Diameter/ft: 12.0 T-T Height/ft: 21.8 Weight/lbs: 28,402 Vessel Volume/gal: 21,856 Slurry Volume/gal: 9,555	Operating Temperature: (-10)-482 Deg F Operating Pressure: 0-45 PSIG Design Temperature: 600 Deg F Design Pressure: 55 PSIG	CS	Industrial Piping Pineville, NC
29D-25	Methanol Area Drain Tank Vertical Vessel	1		Diameter/ft: 5.0 T-T Height/ft: 8.0 Weight/lbs: 3,120 Volume/gal: 1,175	Operating Temperature: (-10)-200 Deg F Operating Pressure: ATM Design Temperature: 400 Deg F Design Pressure: 2 PSIG	CS	Modern Welding Owensboro, KY
29D-30	Fresh Oil Storage Tank Vertical Vessel	1		Diameter/ft: 10.0 T-T Height/ft: 14.9 Weight/lbs: 10,500 Volume/gal: 10,686	Operating Temperature: (-10)-200 Deg F Operating Pressure: ATM Design Temperature: 400 Deg F Design Pressure: 2 PSIG	CS	Industrial Piping Pineville, NC

Item #	Item Description / Type	Number In Use	Number Spare	Capacity / Size	Design Characteristics	Construction Materials	Vendor
29G-25	Methanol Drain Tank Lift Pump Durco Centrifugal	1		Flow Rate: 40 GPM Power: 20.4 BHP Motor: 40 HP	Operating Temperature: 105 Deg F Inlet Pressure: 14.1 PSIA Outlet Pressure: 160.6 PSIA Design Temperature: 250 Deg F Design Pressure: 220 PSIG	CS	Duririon Dayton, OH
29G-26	Methanol Drain Tank Sump Pump Wilden Diaphragm	1		Flow Rate: 20 GPM Power: Air-Driven	Operating Temperature: 60 Deg F Inlet Pressure: (-10.7) PSIA Outlet Pressure: 5.3 PSIA Design Temperature: 125 Deg F Design Pressure: 150 PSIG	Polypropylene	Quaker Pump Lansdale, PA
29V-02	Instrument Air Drier Skid Exhaust Purge Heat Reactivated	1		Max inlet flow: 347 SCFM Net out flow: 317 SCFM Overall Dim/ft: 8 x 4 x 6 Weight/lbs: 1450	Inlet Pressure: 80 PSIG Inlet Temperature: 100 Deg F Overall Pressure Drop: 5 PSI Product Air Temp: < 120 Deg F Product Dewpoint: (-40) Deg F	CS Dessicant: ALCOA activated alumina	Henderson Engin Mt. Laurel, NJ
29Y-02	Slurry Tank Agitator	1		Motor: 50 HP	Operating Temperature: 60-500 Deg F Operating Pressure: 55 PSIG Design Temperature: 600 Deg F Design Pressure: 55 PSIG	304 SS	Chemineer Dayton, OH

4.7 PHOTOGRAPHS OF THE FACILITY

Please see Appendix E.

5. PROCESS CAPITAL COST

The Liquid Phase Methanol Demonstration Project is being performed under the DOE Cooperative Agreement No. DE-FC22-92PC90543. The project cost summary is provided in Table 5-1.

Table 5-1
Kingsport LPMEOH™ Project Cost Summary
(Dollars in Thousands)

Pre-Award and Previous Site Development	$16,304
Project Definition	$1,051
Design Engineering	$11,576
Procurement	$9,783
Construction	$11,550
Training and Commissioning	$1,115
Start-up	$680
LPMEOH™ Process Demonstration Facility Operation	$148,279
Dismantlement	$364
On-site Testing (Product Use Demonstration)	$4
Off-site Testing (Product Use Demonstration)	$3,982
Data Analysis and Reports	$2,670
Planning and Administration (including DME DVT)	$6,343
Total	$213,700

Table 5.2 contains the Capital Cost Summary for the LPMEOH™ demonstration unit according to the area breakdown in Section 4. Unit costs are expected to decrease as the maximum production capability of the equipment within the LPMEOH™ demonstration unit is determined. Also, design and construction costs will likely be reduced as future plants will not require the instrumentation and analytical equipment which is included in a first-of-a-kind facility.

Table 5-2
Kingsport LPMEOH™ Capital Cost Summary
(Dollars in Thousands)

Area	Item Description	Total	Subtotal
A	**Feed Gas Purification**	**$77.0**	$77.0
B	**Compression**	**$776.8**	
	Compression		$699.6
	Exchangers		$37.6
	Separators		$39.6
C	**Reactor Loop and Catalyst Reduction**	**$2,194.2**	
	Equipment		$1,487.8
	Exchangers		$332.3
	Pumps		$374.1
D	**Distillation**	**$703.4**	
	Equipment		$309.5
	Exchangers		$214.7
	Tanks		$131.3
	Pumps and Misc.		$47.9
E	**Storage and Miscellaneous**	**$648.6**	$648.6
	Total	**$4,400.0**	$4,400.0

6. ESTIMATED OPERATING COSTS

The summary of the estimated costs for startup of the LPMEOH™ demonstration unit is provided in Table 6-1. A report on the startup of the LPMEOH™ demonstration unit has been published (Ref. h).

Table 6-1
Summary of Estimated Startup Costs

Base Year 1997

Startup Cost Element	Cost, $
Operating Labor Cost	174,000
Maintenance and Materials Cost	100,000
Administrative and Support Cost	268,000
Commodity Cost	
1. Syngas	38,000
2. Electric Power	13,000
3. Steam	21,000
4. Cooling Water	8,000
5. Waste Water	10,000
6. Miscellaneous	48,000
Length of Startup Period, months	1

The estimated operating cost for the four-year test program are shown in Table 6-2. The unit costs reflect the costs incurred by the Partnership and are typical of published utility costs.

Table 6-2
Summary of Estimated Operating Costs

Base Year 1997

ANNUAL FIXED OPERATING COST

Number of Operators per Shift	1.5
Number of Shifts per Week	4.2
Operating and Technical Support Labor Rate, $/hr	71.96

	Cost, $/yr
Total Annual Operating and Technical Support Labor Cost	630,367
Total Annual Maintenance Labor Cost	478,837
Total Annual Maintenance Material Cost	234,521
Total Annual Administration and Support Labor Cost	1,806,445
TOTAL ANNUAL FIXED O&M COST	3,150,170

VARIABLE OPERATING COST

Commodity	Unit	$/Unit[1]	Quantity/hr	Cost, $/hr
Compressed Air	1,000 SCF	0.21	18.00	3.71
Demineralized Water	1,000 Lbs	0.62	2,500.00	1.55
Electricity	kWh	0.041	770.00	31.72
Filtered Water	1,000 Gal	0.46	0.30	0.14
High Pressure Steam	1,000 Lbs	5.15	6.00	30.90
Low Pressure Steam	1,000 Lbs	4.64	-1.00	-4.64
Medium Pressure Steam	1,000 Lbs	4.12	12.50	51.50
Nitrogen	1,000 SCF	0.62	4.00	2.47
Syngas	1,000 SCF	4.20	850.00	3,570.00
Catalyst	Lbs	8.5	6.67	56.67
Mineral Oil	Lbs	0.43	20.99	9.03
Waste Water	1 year	14,214		1.85
Slurry Disposal	1 year	26,780		3.49
Distillation of 29C-20 Underflow	1 Gal	0.052	781.25	40.23
Area Services (laboratory, general services)	1 month	81,600		127.50
TOTAL VARIABLE OPERATING COST				3,926.12
TOTAL PLANNED OPERATING HOURS FOR DEMONSTRATION (at 320 days/year)				30,720

[1] - These unit costs reflect the costs incurred by the Air Products Liquid Phase Conversion Company, L.P. (the Partnership) and are typical of published utility costs (for example, Process Economics Program Report 136A, “Plant Utility Costs”, published by SRI International, Menlo Park, CA).

7. COMMERCIAL APPLICATIONS

The design of the LPMEOH™ Process Demonstration Unit at Kingsport recognized the commercial application requirements. These are discussed in the following Sections 7.1 through 7.5. The Demonstration Test Plan and the Environmental Monitoring Plan, were prepared to fully satisfy these commercial needs. The IGCC integration for methanol coproduction will be simulated at Kingsport, over a wide range of operating conditions, so that engineering data needed for future commercial designs will be obtained.

7.1 METHANOL COPRODUCTION WITH IGCC - DESIGN OPTIONS

The LPMEOH™ process is a very effective technology for converting a portion of the H_2 and CO in an IGCC electric power plant's coal-derived syngas to methanol. The process is very flexible in being able to process many variations in syngas composition. The LPMEOH™ process can be used with an IGCC power plant (Ref. a), to provide the once-through methanol production as depicted in Figure 7-1. The process can be designed to operate in a continuous, baseload manner, converting syngas from oversized gasifiers or from a spare gasifier. The process can also be designed to operate only during periods of off-peak electric power demand to consume a portion of the excess syngas and allow the electricity output from the combined-cycle power unit to be reduced. In this latter circumstance, the gasifiers continue to operate at full baseload capacity, so the IGCC facility's major capital asset is fully utilized. In either baseload or cycling operation, partial conversion of between 20% and 40% of the volume of H_2 and CO in the IGCC power plant's syngas is optimal on an economic basis, and conversion of up to 50% is feasible.

A simplified process flow diagram for the LPMEOH™ process design options, is shown in Figure 7-2. This shows several once-through LPMEOH™ process design options, as described more fully in Ref. b, and summarized in the following.

In its simplest configuration, part or all of the syngas (feed gas) at its maximum available pressure from the IGCC power plant (Stream 1) is passed once, without recycle through the LPMEOH™ unit. The unreacted gas (Stream 3) is returned to the IGCC power plant's combustion turbines.

If greater amounts of syngas conversion are required, different once-through design options (Figure 7-2) are available. These are discussed later in this section. With any of these options, there is still no need for upstream stoichiometric adjustment of the feed gas by the water-gas shift reaction and CO_2 removal; so the simplicity of once-through CO-rich gas processing is retained.

The design configuration for the LPMEOH™ process depends upon the degree of conversion of syngas (or the quantity of methanol produced relative to the power plant size). The gas pressure has a strong effect on the degree of syngas conversion, as shown in Figure 7-3. The various LPMEOH™ process design options (shown also on Figure 7-2) for greater syngas conversion are:

“ONCE-THROUGH” COPRODUCTION

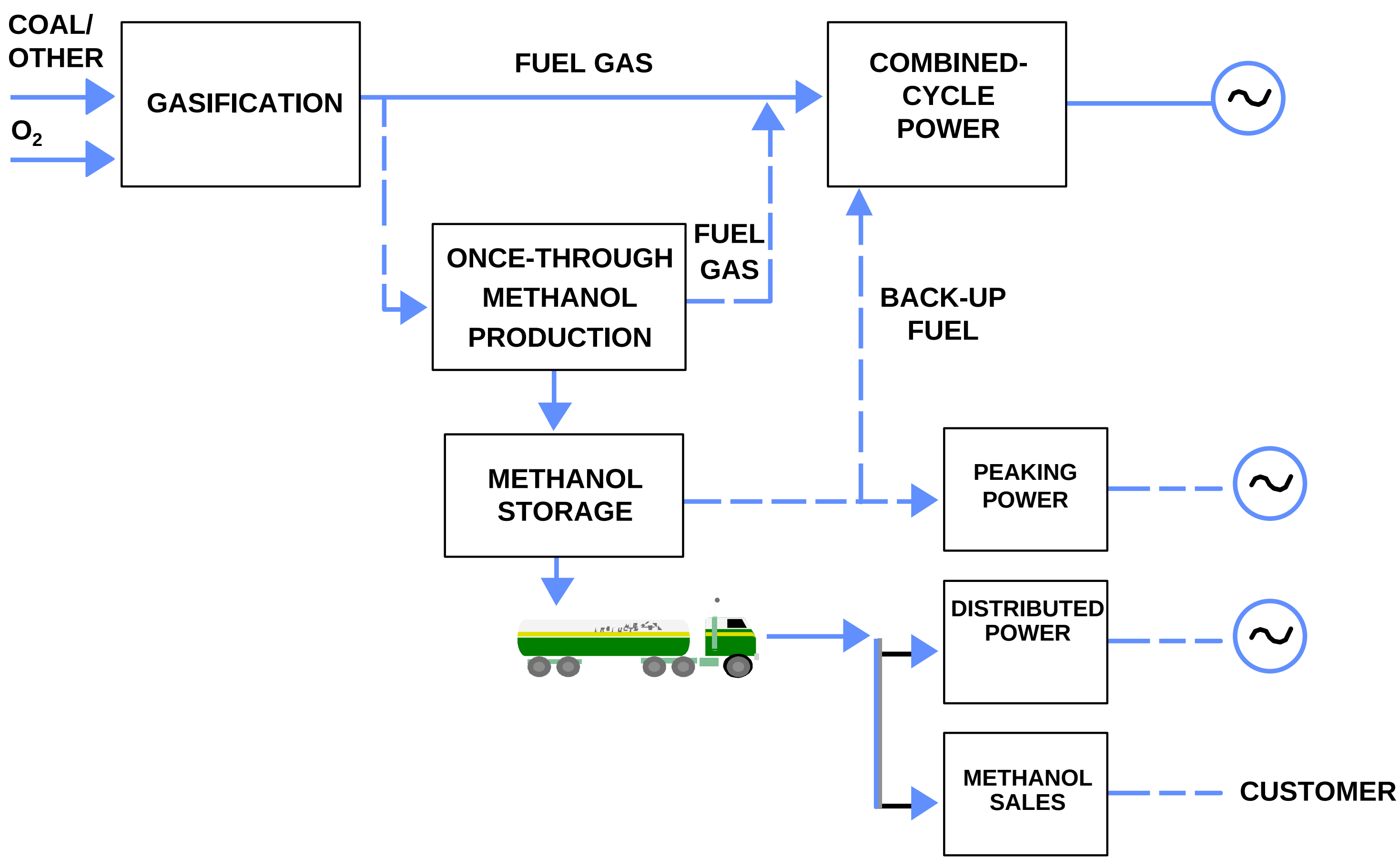


Figure 7-1 Once-Through Coproduction

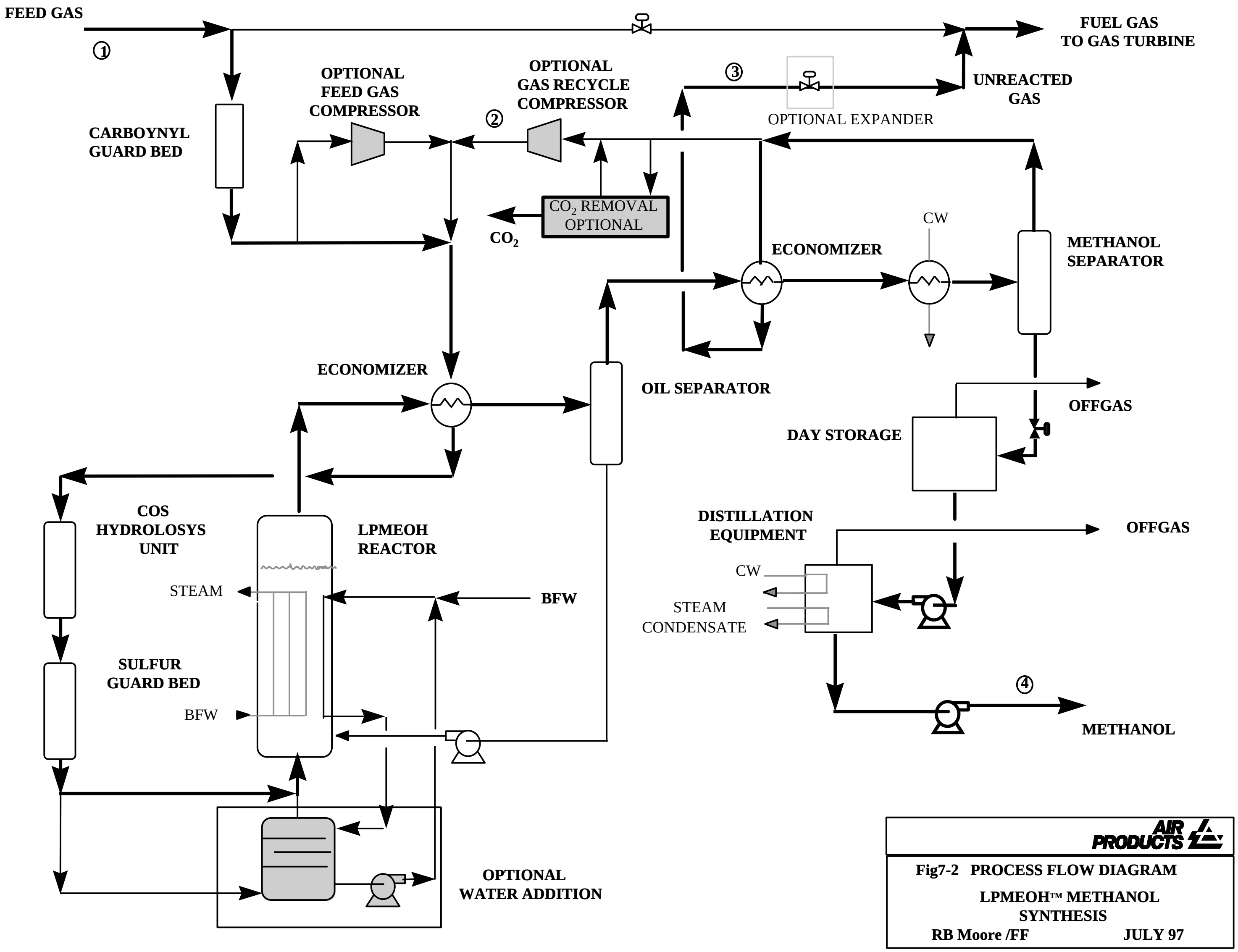


Figure 7-2 Process Flow Diagram

- Once-Through, with Gas Recycle
  One design technique to increase the degree of syngas conversion is to condense out methanol from the reactor effluent and to recycle part of the unreacted feed gas back to the reactor inlet. With the LPMEOH™ process, this simple recycle refers to recycle of CO-rich gas. The recycle ratio required for the LPMEOH™ process is moderate, for example, one part unreacted syngas to one part fresh feed gas. This 1 to 1 recycle ratio (Figure 7-3 - 1:1 recycle) is usually quite effective in optimizing the methanol production. At higher recycle ratios, little is gained, since most of the available H_2 has already been converted to methanol.

- Once-Through, with Feed Gas Compression
  When the feed gas pressure from the IGCC electric power plant is low (e.g. below 750 psia), feed gas compression may be added to the LPMEOH™ process design, to increase reactor productivity and overall conversion of syngas to methanol.

- Once-Through, with Water Addition
  If additional conversion is desired, the LPMEOH™ process design can be altered to generate additional H_2. The inherent shift activity of the methanol catalyst can be utilized to accommodate a modest amount of shift activity within the reactor. This is done by the addition of water, as steam, to the syngas before it passes through the liquid phase methanol reactor. Within the reactor, the additional steam is converted to H_2, which is, in turn, converted to methanol (Figure 7-3 - 1:1 Recycle with 5% water). In the water addition case, the increase in conversion is accompanied by a modest increase of water in the crude methanol product and of CO_2 in the reactor effluent gas.

- Once-Through, with Water Addition and CO_2 Removal
  If syngas conversion greater than about 50% is required, then a CO_2 removal unit, combined with the water addition option, is effective.

- Staged Liquid Phase Reactors
  Although not shown in Figure 7-2, Staged Liquid Phase Reactors are also a design option alternative. For designs requiring higher syngas conversions, two liquid phase reactors, staged in series, are an alternative to the Gas Recycle Compression design option. The staged reactor design option should be considered when compression could be eliminated, and/or for cases where reactor shipping sizes are a constraint. A partial listing of the potential advantages and considerations for staged reactor designs is given in Table 7-1.

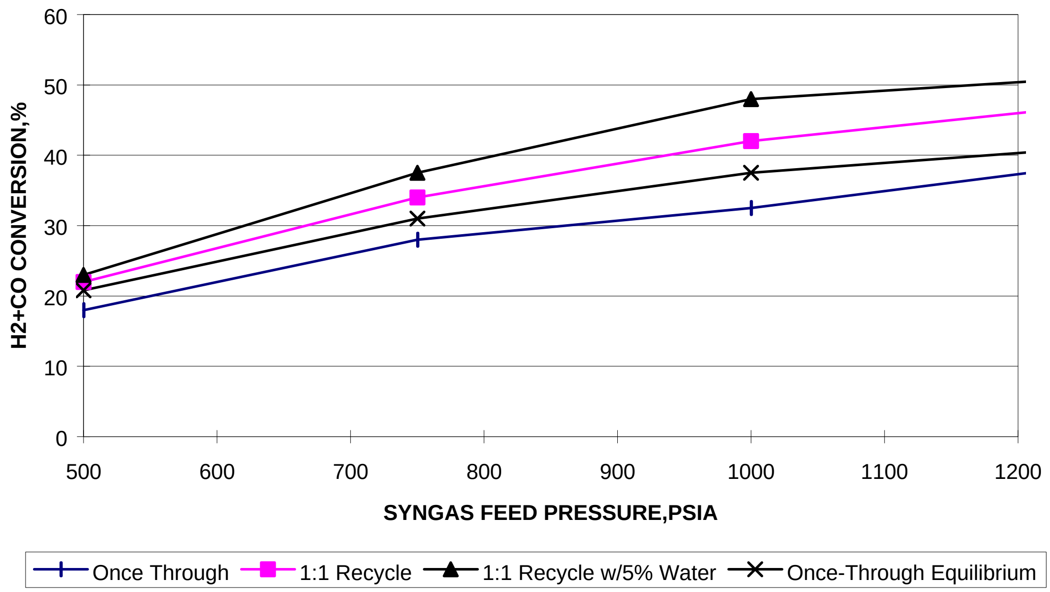


Figure 7-3 Effect of Pressure on Syngas Conversion to Methanol

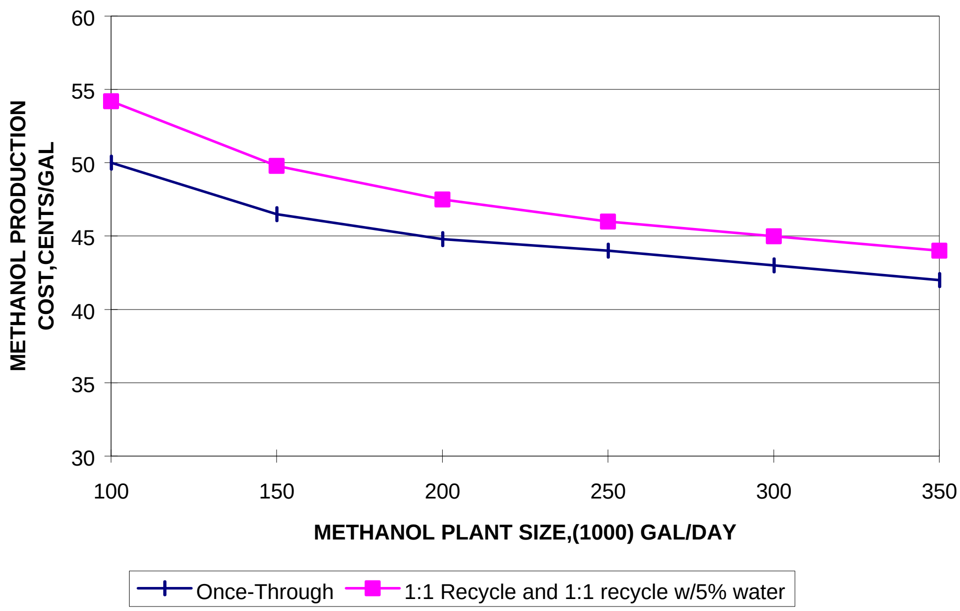


Figure 7-4 IGCC Coproduct Methanol Cost versus Methanol Plant Size

Table 7-1
Staged Reactors - Design Considerations

Design Consideration	Advantage
Reactor design (SV's) optimized for each stage.	Allows each reactor to be the same (volumetric and length/diameter) size; each much smaller than a single stage reactor.
In-situ catalyst reduction, with temperature ramping/catalyst aging.	Use one reactor for initial operation, until second is needed.
Staged catalyst addition, with fresh catalyst in reactor #2, old catalyst in #1.	Enhance catalyst life and reduce risk of losses due to feed gas poisons or upsets.
Water addition/ CO_2 removal Optimization	Could be optimized (for example: to stage #1 only, with partial interstage removal; or to both stages).
Reactor temperature optimization.	Separately optimized and controlled for each reactor.
On-off and load-following operation.	No compressor to start/stop and control. Part load operation could be with one reactor shut down.

Summary of Design Options

The higher the CO content of the once-through syngas, the more the production of methanol is limited by the availability of H_2. Normally the least expensive methanol conversion cost comes from converting as much hydrogen as practical without feed gas compression, unreacted gas recycle, or further processing of the feed gas. The higher the pressure at which the syngas is available, the greater is the degree of conversion and the lower the conversion cost. If additional conversion is required, then the feed syngas (Stream 1 in Figure 7-2) may be compressed (gas feed compressor option) and a portion of the unreacted syngas (Stream 2) may also be compressed and recycled back to the reactor inlet (gas recycle compressor option). Reaction pressure for methanol synthesis design is usually 750 psia or higher, so that if the available feed pressure is below this, then these two compression options are normally combined, utilizing a single compressor for the best economy. Current economic analysis indicates that the feed compression of up to 1,250 psig, and a recycle gas ratio of 1:1 (Stream 2 : Stream 1) is about optimum. If additional conversion is required, then the water addition design option, without or with CO_2 removal, may be combined with the compression design options.

7.2 SYNTHESIS GAS COMPOSITION VARIATIONS

Coal composition affects syngas composition from an IGCC power plant only slightly. Sulfur is effectively removed and recovered as a by-product with conventional (low temperature) syngas cleanup technology, to a generally low (5 to 20 ppm) level as set by economic and environmental design considerations. Sulfur content in the available feed gas has a significant design and cost impact on LPMEOH™ unit design (Ref. a). The inert content of the syngas will vary with coal composition, but variations above or below the normal 1% design basis do not significantly affect LPMEOH™ unit design (Ref. a). Syngas compositions (H_2, CO, and CO_2) do vary significantly with the type of gasifier, and the LPMEOH™ unit design (Ref. a) must take these into consideration.

7.3 BASELOAD COPRODUCTION OF METHANOL AND POWER - PROCESS DESIGN STUDY

Process design study work for the LPMEOH™ process has been directed towards converting a portion of coal-derived syngas produced in an IGCC electric power plant to methanol. A feed gas containing 35% H_2, 51% CO, 13% CO_2, and 1% inerts (N_2) was used for preparing the baseload methanol coproduction economics.

With a given gasification plant size, the IGCC coproduction plant can be designed to accommodate a range of methanol to power output ratio's. For example (Ref. c, d), a gasification plant, with two gasifiers of 1,735 million Btu (HHV) per hour output each, could be sized for baseload power output of 426 megawatts of electricity (MWe) and for baseload methanol coproduction of 152,000 U.S. gallons per day. Other plant design size options are shown in Table 7-2.

Table 7-2
Methanol Plant to Power Plant Size Ratio

% of Syngas Converted to Methanol (%)	Baseload Power Plant Size (MWe)	Baseload Methanol Plant Size (G/D)	Methanol Plant to Power Plant Size Ratio (G/D per MWe)
0	500	0	0
13.8	**426**	**152,000**	**357**
20.0	394	210,000	533
30.0	342	330,000	965

The IGCC coproduction plant with 426 MWe of power and 152,000 gallon/day of methanol is used for the baseload production cost estimate for coproduced methanol, shown in Table 7-3. If the baseload fuel gas value is $4.00 per million Btu, then 152,000 gallons/day of methanol can be coproduced from coal for under 50 cents per gallon.

As one would expect, the methanol production cost is lower at larger methanol plant sizes. Figure 7-4, shows the effect of plant size for once-through methanol coproduction. Methanol production costs for two of the LPMEOH™ plant design options for higher syngas conversion: 1 to 1 gas recycle, and 1 to 1 gas recycle with water addition, are also shown.

Today, new methanol plants are being built where natural gas is inexpensive (Chile, Saudi Arabia). These new world-scale plants range in size from 700,000 to 900,000 gallons/day (2,000 to 2,700 metric tons per day) in size. The economy of scale savings in natural gas gathering, syngas manufacturing, and in methanol storage and ocean transport facilities, drive these plants to their large size. Estimates (Ref. e, f) show that an 836,000 gallon/day off-shore methanol plant (with the same, 20% per year capital charge as in Table 7-3 and Figure 7-4), with natural gas at $0.50 to $1.00 per million Btu, has a total ex-plant methanol production cost of 46 to 50 cents per gallon. Adding ocean freight, duty, and receiving terminal storage typically adds 8 to 10 cents per gallon; giving a total delivered U.S. Gulf Coast methanol cost (Chemical Grade) of 55 to 60 cents per gallon.

Figure 7-4 is interesting, because it provides an unexpected result. Methanol coproduction with IGCC and the once-through LPMEOH™ process, does not need large methanol plant sizes to achieve good economies of scale. The gasification plant is already at a large economical scale for power generation; so the syngas manufacturing economies are already achieved. Methanol storage and transport economies are also achieved by serving local markets, and achieving freight savings over the competing methanol, which is usually shipped from the U. S. Gulf coast.

The 50 cents per gallon coproduction cost for a 152,000 gallon/day once-through LPMEOH™ unit size is competitive in local markets with new world-scale, off-shore methanol plants. Figure 7-4 shows an additional 3 to 4 cent per gallon savings for a 350,000 gallon/day LPMEOH™ unit size. These additional savings might be used to off-set higher freight costs to more distant local customers; while still maintaining a freight and cost advantage over the imported methanol from the Gulf Coast.

7.4 TECHNOLOGY DESCRIPTION

The heart of the LPMEOH™ process is the slurry bubble column reactor (Figure 7-5). The liquid medium is the feature that differentiates the LPMEOH™ process from conventional technology. Conventional methanol reactors use fixed beds of catalyst pellets and operate in the gas phase. The LPMEOH™ reactor uses catalyst in powder form, slurried in an inert mineral oil. The mineral oil acts as a temperature moderator and a heat removal medium, transferring the heat of reaction from the catalyst surface via the liquid slurry to boiling water in an internal tubular heat exchanger. Since the heat transfer coefficient on the slurry side of the heat exchanger is relatively large, the heat exchanger occupies only a small fraction of the cross-sectional area of the reactor. The slurry reactor can thus achieve high syngas conversion per pass, due to its capability to remove heat and maintain a constant, highly uniform temperature through the entire length of the reactor.

Because of the LPMEOH™ reactor's unique temperature control capabilities, it is able to directly process syngas which is rich in carbon oxides (CO and CO_2). Gas phase methanol technology would require such a feedstock to undergo stoichiometry adjustment by the water gas shift reaction (to increase the H_2 content) and CO_2 removal (to reduce the excess carbon oxides). In a gas phase reactor, temperature moderation is only achieved by recycling large amounts of H_2-rich gas, utilizing the higher heat capacity of H_2 gas as compared to CO gas. Typically a gas phase reactor is limited to about 16% CO gas in the inlet to the reactor, in order to limit the conversion per pass to avoid excess heating. In contrast, with the LPMEOH™ reactor, syngas having CO

TABLE 7-3

Production Cost Estimate for Coproduced Methanol

LPMEOH™ Plant Capacity: 152,000 gallons per day (500 sT/D)

Capital Investment: Circa $30 million

		Baseload
Methanol Plant Operation:	*(Hours/year)*	7884 hr/yr.
Methanol Production	*(million Gal./year)*	49.9

Methanol Production Cost		**cents/gallon**
Syngas cost		
Feed Gas @ fuel value ($4.00/mmBtu)		98.7
Unreacted (CO-rich) gas @ fuel value ($4.00/mmBtu)		(68.4)
Subtotal; net cost of syngas converted:		**30.3**
Operating cost		
Catalyst and chemicals		2.6
Export steam		(2.9)
Utilities		0.9
Other (fixed) costs		4.0
Subtotal; Operating Costs:		**4.6**
Capital charge	@ 20% of investment per year	**11.6**
Total Methanol Production Cost:		**46.5**

Basis:

U.S. Gulf Coast Construction, 4thQ 1996 $

Includes owner costs and 30 days of Product Storage

CO-rich feed gas from IGCC electric power plant at 1,000 psia, with 5ppm (max.) sulfur.

Once-through LPMEOH™ process design with 1,562 mmBtu/hr in, 1,082 mmBtu/hr out (HHV)

Excludes License and Royalty fee. Air Products is the LPMEOH™ process technology licensor.

Product methanol with 1 wt% water; Chemical Grade would add 4 to 5 cents per gallon.

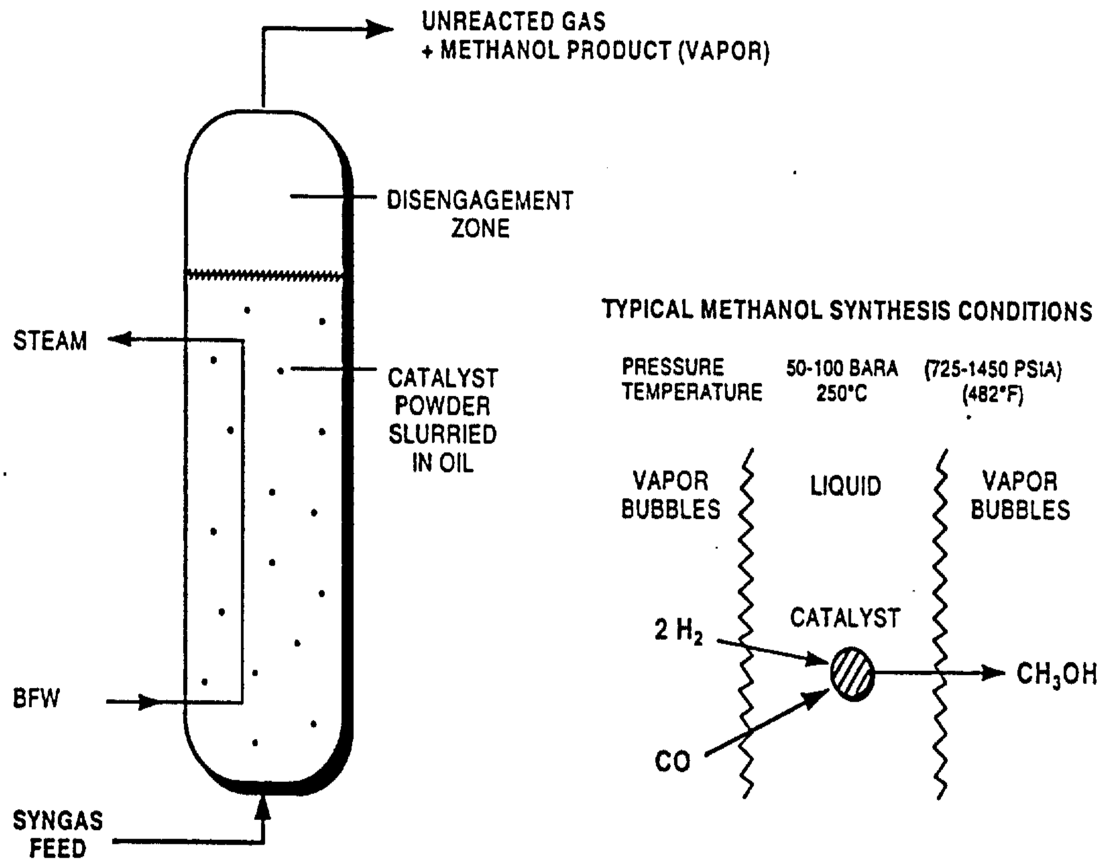


Figure 7-5 LPMEOH™ Reactor and Reaction Schematics

gas concentrations in excess of 50% have been routinely tested without any adverse effect on the catalyst activity.

A second differentiating feature of the LPMEOH™ reactor is its robust character. The slurry reactor is suitable for rapid ramping, idling, and even extreme stop/start actions. The thermal moderation provided by the liquid inventory in the reactor acts to buffer sharp transient operations that would not normally be tolerable in a gas phase methanol synthesis reactor.

A third differentiating feature of the LPMEOH™ process is that a high quality methanol product is produced directly from syngas which is rich in carbon oxides. Gas phase methanol synthesis, which relies on H_2-rich syngas, results in a crude methanol product with a water content ranging from 4 to 20% water by weight. The product from the LPMEOH™ process typically contains only 1% water by weight. This methanol product, coproduced with IGCC, is therefore suitable for many applications, and at a substantial savings in purification costs. These savings are a significant part of the LPMEOH™ process technology's advantage, and the applications for this methanol product (produced directly from syngas which is rich in carbon oxides) are discussed in the next section. These applications are to be tested as part of the fuel-use test plan.

7.5 APPLICATIONS FOR THE COPRODUCED METHANOL PRODUCT

The methanol coproduction process studies show that the LPMEOH™ process can produce a clean high quality methanol product at less than 50 cents per gallon; from an abundant, non-inflationary local fuel source (coal). Serving local markets, the methanol coproduced at central IGCC electric power plants, can be a valuable premium fuel or fuel feedstock for many applications, such as:

1. An economical hydrogen source for small fuel cells, which are being developed for transportation applications. Methanol is a storable, and transportable, liquid fuel which can be reformed under mild conditions to provide an economical source of H_2 for fuel cells.

2. When reformed under mild conditions, may be an economical source of H_2 or CO for industrial applications.

3. A substitute for chemical grade methanol being used for MTBE manufacture. (MTBE is added to gasoline to boost octane and to meet environmental clean air mandates. MTBE is one of the major current markets for methanol.)

4. An environmentally advantaged fuel for dispersed electric power stations. Small packaged power plants (combustion turbine, internal combustion engine, or fuel cell) provide power and heat locally, at the use point, eliminating the need for natural gas pipelines and high voltage power lines.

5. Finally, the coproduced methanol may be used by the utility owning the IGCC facility (see Figure 7-1). Potential uses are: a) as a backup fuel for the IGCC plant's main gas turbines; b.) as a fuel for a separate, dedicated cycling combined-cycle unit at the same site; c.) as the fuel exported to the utility's distributed power generation system(s); or d) as the transportation fuel for the utility's bus or van pool. Since methanol is an ultra-clean (zero sulfur) fuel which burns with very low (better than natural gas) emissions of nitrogen oxides, the incremental power is very clean. Since the methanol is derived from the coal pile, the IGCC facility can be truly independent and self-sufficient for fuel needs. In addition, should the external prices for methanol command higher value to the IGCC plant's owner, the methanol can be exported for additional revenues.

Many of the applications listed above, are embryonic developments. Their ultimate market size potential; for transportation applications, for industrial applications and for distributed power generation; could become large. The methanol product specification for the five applications is not adequately known. Therefore, part of the LPMEOH™ demonstration project's program is to confirm the suitability of the methanol product for these (and other) uses. Fuel-use tests will be used to develop final methanol product specifications. During the demonstration, a maximum of 400,000 gallons of the "as-produced from CO-rich syngas" methanol will be available for off-site, product-use testing. The off-site fuel-use testing in both stationary and mobile applications is underway.

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LISTING OF MEETINGS, PAPERS AND PUBLICATIONS

DATE	MEETING/PUBLICATION	TITLE	AUTHORS
22-24 September 1992	DOE Contractors' Conference, Pittsburgh, PA	"Liquid Phase Fischer-Tropsch Synthesis in a Bubble Column"	B. L. Bhatt* - APCI E. S. Schaub - APCI E. C. Heydorn - APCI D. W. Studer - APCI D. M. Brown - APCI
22-24 September 1992	DOE Contractors' Conference, Pittsburgh, PA	"Demonstration of a Slurry Phase Shift Process in the Alternative Fuels Development Unit"	T. H. Hsiung* - APCI D. M. Herron - APCI E. C. Heydorn - APCI E. S. Schaub - APCI
22-24 September 1992	First Annual Clean Coal Technology Conference, Cleveland, OH	"Fuel and Power Coproduction: The Integrated Gasification/Liquid Phase Methanol (LPMEOH™) Demonstration Project"	W. R. Brown* - APCI F. S. Frenduto - APCI
27-29 March 1993	2nd International Conference on Gas-Liquid-Solid Reaction Engineering	"Flow Patterns in a High Pressure Slurry-Bubble-Column Reactor Under Reaction Conditions"	B. A. Toseland* - APCI D. M. Brown - APCI M. Dudukovic - APCI B. S. Zou
26-29 April 1993	18th International Tech. Conference on Coal Utilization and Fuel Systems	"Recent Developments in Slurry Reactor Technology at the LaPorte Alternative Fuels Development Unit"	B. L. Bhatt* - APCI E. S. Schaub - APCI E. C. Heydorn - APCI
2-7 May 1993	13th North American Meeting of Catalysis Society, Pittsburgh, PA	"A C_1 Route to MTBE" Catalysis for Reformulated Fuels" (invited contribution)	D. M. Brown* - APCI T. H. Hsiung - APCI F. J. Waller - APCI R. P. Underwood - APCI
20-24 September 1993	10th Annual International Pittsburgh Coal Conference	"Dehydration of Isobutanol in a Slurry-Phase Reaction"	P. A. Armstrong* - APCI B. L. Bhatt - APCI B. E. Latshaw - APCI B. A. Toseland - APCI R. P. Underwood - APCI

LIQUID PHASE METHANOL AND ALTERNATIVE FUELS
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DATE	MEETING/PUBLICATION	TITLE	AUTHORS
27-29 September 1993	DOE Contractors' Conference Coal Liquifaction and Gas Conversion Contractors Review	"Isobutanol Dehydration: A Key Step in Producing MTBE from Syngas"	P. A. Armstrong _ APCI B. L. Bhatt - APCI E. C. Heydorn - APCI B. A. Toseland* - APCI
17-22 October 1993	Joint ASME/IEEE Power Generation Conference, Kansas City, KS	"Liquid Phase Methanol Energy Storage with Gasification Combined-Cycle Power Generation"	F. S. Frenduto - APCI J. Klosek - APCI E. R. Osterstock - APCI
7-10 November 1993	Tenth International Symposium on Alcohol Fuels, Colorado Springs, CO	"Synthesis Gas Conversion to Isobutanol: A Key Step in a C_1 Route to MTBE"	R. P. Underwood* - APCI E. S. Schaub - APCI
17-19 November 1993	Power-Gen '93, Dallas, TX	"Cost-Effective Dispatchable Power from a Gasification Combined-Cycle System: Liquid Phase Methanol Energy Storage"	F. S. Frenduto - APCI E. R. Osterstock - APCI G. D. Snyder - APCI
5-6 January 1994	1994 $IChem^E$ Research Event	"Oxygenated Fuels and Chemicals Using Slurry Reactor Technology"	B. A. Toseland* - APCI D. M. Brown - APCI
March 1994	OilGas - European Magazine	"Synthesis Gas Conversion to Isobutanol: A Key Step in a C_1 Route to MTBE"	R. P. Underwood* - APCI E. S. Schaub - APCI
15-17 March 1994	1994 ACS Spring Meeting San Diego, CA	"Catalyst and Process Scale-up for Fischer Tropsch Synthesis"	B. L. Bhatt* - APCI R. Frame - UOP A. Hoek - Shell, Amsterdam K. Kinnari - Statoil, Norway V. U.S. Rao - DOE/PETC - PA F. L. Tungate - UCI
17-21 April 1994	1994 AIChE Spring Meeting, Atlanta, GA	"Dehydration of Isobutanol in a Slurry-Phase Reactor"	P. A. Armstrong* - APCI B. L. Bhatt - APCI B. E. Latshaw - APCI B. A. Toseland - APCI R. P. Underwood - APCI

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DATE	MEETING/PUBLICATION	TITLE	AUTHORS
9-11 May 1994	1994 International Symposium on Gas Conversion and Utilization Exxon, NJ	"Government and Industry Partnership in Developing New Technologies for Fuels and Chemicals"	D. M. Brown* - APCI
Summer 1994	Hydrocarbon Technology International (Magazine)	"Gasification of Residues: Implications for Plant Economics"	W. F. Baade - APCI H. H. Gunardson - APCI
6-8 September 1994	Third Annual Clean Coal Technology Conference, Chicago, IL	"Flexible Electric Power Generation: The Integrated Gasification/Liquid Phase Methanol (LPMEOH™) Demonstration Project"	W. R. Brown* - APCI R. B. Moore - APCI
7-8 September 1994	Coal Liquefaction and Gas Conversion DOE Contractor's Review Conference	"New Technologies for Fuels and Chemicals - the Alternative Fuels II Program"	B. A. Toseland* - APCI R. P. Underwood - APCI F. J. Waller - APCI
7-8 September 1994	Coal Liquefaction and Gas Conversion DOE Contractor's Review Conference	"Recent Progress on Syngas Conversion to Isobutanol"	E. C. Heydorn - APCI E. S. Schaub - APCI R. P. Underwood*- APCI V. E. Stein - APCI F. J. Waller - APCI
27-29 March 1995	2nd International Conference On Gas-Liquid-Solid Reactor Engineering - IChem^{E} Cambridge, UK	"Flow Patterns In High Pressure, Slurry - Bubble-Column Reactors Under Reaction Conditions"	*B. A. Toseland - APCI D. M. Brown - APCI B. S. Zou - Washington Univ. M.P. Dudukovic - Wash. Univ.
3-5 April 1995	209th American Chemical Society National Meeting - Anaheim, CA	"Commercial-Scale Demonstration of a Liquid-Phase Methanol Process"	*Steven L. Cook - Eastman Chemical Company
May 1995	Topics In Catalysis, Vol.2, 1995, No. 1-4 (See also 15-17 March, 1994).	"Catalyst and Process Scale-up for Fischer-Tropsch Synthesis"	B.L.Bhatt - APCI R. Frame - UOP A. Hoek - Shell, Amsterdam K. Kinnari - Statoil, Norway U. Rao - DOE/PETC - PA F. Tungate - United Catalyst

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DATE	MEETING/PUBLICATION	TITLE	AUTHORS
11-16 June 1995	14th North American Meeting of the Catalysis Society, Snowbird, Utah	"Productivity Improvements for Fischer-Tropsch Synthesis"	B. L. Bhatt* - APCI D. M. Brown - APCI E. C. Heydorn - APCI A. Hoek - Shell, Amsterdam G. Stiegel - DOE/PETC - PA
30 August 1995	Coal Liquefaction and Gas Conversion DOE Contractors Review Conference	"Catalyst Activity Maintenance Study for the Liquid Phase Dimethyl Ether Process"	X. D. Peng - APCI B. A. Toseland - APCI R. P. Underwood - APCI
30 August 1995	Coal Liquefaction and Gas Conversion DOE Contractors Review Conference	"Flow Patterns in a Slurry Bubble Column Reactor Under Reaction Condition"	D. M. Brown - APCI B. A. Toseland - APCI M. P. Dudukovic - Wash. Univ. St. Louis, MO
5-8 September 1995	4th Annual DOE Clean Coal Technology Conference, Denver, Colorado	"An Update on Liquid Phase Methanol (LPMEOH™) Technology and the Kingsport Demonstration Project"	E. S. Schaub*- APCI V. E. Stein - APCI E. C. Heydorn - APCI E. R. Osterstock - APCI
24-29 March 1996	American Chemical Society, Spring Meeting, New Orleans, La.	“Conversion of Syngas to Chemicals"	F. J. Waller* - APCI
9-11 July 1996	First Joint Power & Fuel Systems - Contractors Conference (DOE) - Pittsburgh, Pa. - July 1996	“Progress in Understanding The Fluid Dynamics of Bubble Column Reactors”	S. Degaleesan - Wash. Univ. St. Louis, MO S. Kumar - Wash. Univ. St. Louis, MO M.P. Dudukovic - Wash. Univ. St. Louis, MO B.A. Toseland - APCI B.L. Bhatt - APCI

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DATE	MEETING/PUBLICATION	TITLE	AUTHORS
9-11 July 1996	First Joint Power & Fuel Systems - Contractors Conference (DOE) - Pittsburgh, Pa. - July 1996	"Recent Results From The LaPorte Alternative Fuels Development Unit"	B.L. Bhatt - APCI E.C. Heydorn - APCI B.A. Toseland - APCI T.J. O'Hern - Sandia K.A. Shollenberger - Sandia
9-11 July 1996	First Joint Power & Fuel Systems - Contractors Conference (DOE) - Pittsburgh, Pa. - July 1996	"Some Advances In Catalysis For Alternate Fuels"	X.D. Peng - APCI G.E. Parris - APCI B.A. Toseland - APCI R.P. Underwood - APCI K. Verkerk - ITCP, Aachen B. Jaeger - ITCP, Aachen C.H. Finkeldei - ITCP, Aachen W. Keim - ITCP, Aachen
11 November 1996	American Institute of Chemical Engineers, Annual Meeting, Chicago, Illinois - Nov. 1996	"An Investigation of the Cause of Catalyst Deactivation Under Liquid Phase Syngas-to-DME Reaction Conditions"	G.E. Parris - APCI X.D. Peng* - APCI B.A. Toseland - APCI R.P. Underwood - APCI
13 November 1996	Amercian Institute of Chemical Engineers, Annual Meeting, Chicago, Illinois - Nov. 1996	"Two Dimensional Model For Liquid Mixing In Bubble Columns"	S. Degaleesan - Wash. Univ., St. Louis, MO M. Dudukovic - Wash. Univ., St. Louis, MO B.A. Toseland - APCI B.L. Bhatt - APCI
7-10 January 1997	5th Annual DOE Clean Coal Technology Conference, Tampa, FL	"Fuel and Power Coproduction - The Liquid Phase Methanol (LPMEOH™) Process Demonstration at Kingsport"	D.P. Drown* - APCI W.R. Brown - APCI R.B. Moore - APCI W.C. Jones - Eastman Chemical Co. R.M. Kornosky - DOE

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DATE	MEETING/PUBLICATION	TITLE	AUTHORS
9-13 March 1997	American Institute of Chemical Engineers, Spring 1997 Meeting, Houston, TX	"Further Development of the Slurry Phase Fischer-Tropsch Process at LaPorte, Texas"	D.M. Brown* - APCI B.L. Bhatt - APCI E.C. Heydorn - APCI G. Stiegel - DOE U. Mahagaoker - Shell A. Hoek - Shell
9-13 March 1997	American Institute of Chemical Engineers, Spring 1997 Meeting	"Overview of the Liquid Phase Methanol (LPMEOH™) Process Demonstration at Kingsport"	P.J.A. Tijm* - APCI E.C. Heydorn - APCI W.R. Brown - APCI R.B. Moore - APCI
15 April 1997	American Chemical Society Meeting, April 1997; San Francisco, CA	"Advances in Liquid Phase Technology"	P.J.A. Tijm* - APCI E.C. Heydorn - APCI W.R. Brown - APCI R.B. Moore - APCI
April 1997	British Sulphur Publishing Ltd.; in Magazine: "Nitrogen"	"The Liquid Phase Methanol Process Demonstration at Kingsport"	(Acknowledges Jan 1997 CCT Conference Paper)
18 June 1997	Power - Gen Europe '97 Meeting, Madrid, Spain, June 1997	"Dispatchable IGCC Facilities: Flexibility Through Coproduction"	E.R. Osterstock - APCI
July 1997	Energy Frontiers International Conference, Alaska	"Liquid Phase Di-Methyl Ether" A Promising New Diesel Fuel	P.J.A. Tijm* - APCI F.J. Waller - APCI B.A.Toseland - APCI X.D. Peng - APCI
5-8 October 1997	EPRI Gasification Technologies Conference	"Liquid Phase Methanol (LPMEOH) Project Start-Up"	E.C. Heydorn* - APCI V.E. Stein - APCI B.T. Street - EMN R.M. Kornosky - DOE
5-8 October 1997	7th Int'l Symposium on Catalyst Deactivation; Cancun, Mexico	"A Novel Mechanism of Catalyst Deactivation in Liquid Phase Synthesis Gas-to-DME Reactions"	X.D. Peng* - APCI B.A. Toseland - APCI R.P. Underwood - APCI

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DATE	MEETING/PUBLICATION	TITLE	AUTHORS
27-29 October 1997	1997 Fall Mtg. Eastern States Section of the Combustion Institute; Hartford, CT	"A Comparison of Carbon Monoxide Emissions from Butane, Propane, and Dimethyl Ether Flames"	C.A. Frye* - PSU A.L. Boehman - PSU P.J.A. Tijm - APCI
23-27 September 1997	14th Annual International Pittsburgh Coal Conference, Taiyuan, P.R. China	"Advances In Liquid Phase Technology"	P. Tijm* - APCI X.D. Peng - APCI B.A. Toseland - APCI
5-8 October, 1997	EFI Gasification Technologies Conference	"Liquid Phase Methanol (LPMEOH™) Project Start-Up"	E.C. Heydorn* - APCI V.E. Stein* - APCI B.T. Street - EMN R.M. Kornosky - DOE
5-8 October, 1997	7th International Symposium on Catalyst Deactivation, Cancun, Mexico	"A Novel Mechanism of Catalyst Deactivation in Liquid Phase Synthesis Gas-to-DME Reactions"	X.D. Peng* - APCI B.A. Toseland - APCI R.P. Underwood - APCI
27-29 October, 1997	1997 Technical Fall Mtg. Of the Eastern States Section of the Combustion Institute, Hartford, CT	"A Comparison of Carbon Monoxide Emissions from Butane, Propane, & Dimethyl Ether Flames	C.A. Frye* - PSU A.L. Boehman* - PSU P.J.A. Tijm - APCI
27-30 October 1997	5th Annual Technical Seminar of the Asian-Pacific Economic Cooperation (APEC) Experts' Group on Clean Fossil Energy, Reno, NV	"Alternate Fuels from Coal Through Application of Liquid Phase Conversion Technology"	P.J.A. Tijm* - APCI F.A. Waller - APCI E.C. Heydorn - APCI X.D. Peng - APCI B.A. Toseland - APCI R.B. Moore - APCI B.L. Bhatt - APCI
13 November 1997	Cancun, Mexico	"Chemicals, Transportation Fuels and Transportation Fuel Additives from Synthesis Gas"	F.J. Waller* - APCI P.J.A. Tijm - APCI B.A. Toseland - APCI
16-21 November 1997	AIChE Meeting, Los Angeles, CA	(Slides Only)	B.A. Toseland* - APCI

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DATE	MEETING/PUBLICATION	TITLE	AUTHORS
8-9 December 1997	GURF Meeting, Houston, TX	"Rapid Phase Transitions Prediction Tools Applied to Large Scale FPSO Structures"	
8-10 December 1997	1997 World Methanol Conference, Tampa, FL	"A New Premium Diesel Fuel From Methanol"	P.J.A. Tijm* - APCI R.J.Waller - APCI E.C. Heydorn - APCI X.D. Peng - APCI B.A. Toseland - APCI
10-12 December 1997	Monitizing Stranded Gas Reserves Conference, Houston, TX	"Gas To Liquids Processes"	P.J.A. Tijm* - APCI F.J. Waller - APCI E.C. Heydorn - APCI X.D. Peng - APCI B.A. Toseland - APCI R.B. Moore - APCI B.L. Bhatt - APCI
January 1998	In Industrial Catalysis News, January 1998 Issue	LPM Process Demo: 1997 Was An Exciting Year	E.C. Heydorn* - APCI V.E. Stein - APCI P.J.A. Tijm - APCI B.T. Street - EMN R.M. Kornosky - DOE
13-14 January 1998	EFI Meeting, San Antonio, TX	"Transportation Fuel Additives from Synthesis Gas"	R. Quinn* - APCI F.J. Waller - APCI
9-13 March 1998	23rd International Technical Conference on Coal Utilization & Fuel Systems, Clearwater, FL	"Advances in Liquid Phase™ Technology"	W.R. Miller* - APCI E.C. Heydorn - APCI R.B. Moore - APCI P.J.A. Tijm - APCI

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DATE	MEETING/PUBLICATION	TITLE	AUTHORS
27 April - 1 May 1998	6th Clean Coal Technology Conference, Reno, NV	"Commercial-Scale Demonstration of the Liquid Phase Methanol (LPMEOH™) Process: Initial Operating Experience"	E.C. Heydorn* - APCI V.E. Stein - APCI P.J.A. Tijm - APCI B.T. Street - EMN R.M. Kornosky - DOE
8-10 June 1998	Windsor Workshop, Toronto, Canada	"Overview of CETANER™ for Diesel Fuel and AET Study of CETANER™ Blended with Low Cetane Diesel Fuel"	P.J.A. Tijm* - APCI
August 1998	American Chemical Society National Meeting in Boston, MA	"The Use of Oxygenated Diesel Fuels for Reduction of Particulate Emissions from a Single-Cylinder Indirect Injection Engine"	H.S. Hess* - PSU M.A. Roan - PSU S. Bhalla - PSU T. Butnark - PSU V. Zarnescu - PSU A.L. Boehman - PSU P.J.A. Tijm - APCI F.J. Waller - APCI
4 August 1998	White Paper	"Oxygenated DME Derivatives as Diesel Fuel Additives	
23-27 August 1998	ACS Symposium, Boston, MA	"Chemistry of Diesel Fuel"	H.S. Hess - PSU M.A. Roan - PSU S. Bhalla - PSU S. Butnark - PSU V. Zarnescu - PSU A.L. Boehman - PSU P.J.A. Tijm - APCI F.J. Waller - APCI
21-25 September 1998	5th Natural Gas Conversion Symposium, Taormina, Sicily, Italy	"Generation of Synthesis Gas Off-Shore: Oxygen Supply and Opportunities for Integration with GTL Technologies"	D.M. Brown* - AP-PLC D. Miller - AP-PLC R.J. Allam - AP-PLC P.J.A.Tijm - APCI

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DATE	MEETING/PUBLICATION	TITLE	AUTHORS
13-16 September 1998	15th ISCRE, Newport Beach, CA	"Kinetic Understanding of Chemical Synergy in LPDME™ Process"	X.D. Peng* - APCI B.A. Toseland - APCI P.J.A. Tijm - APCI
		"A Comparison of CO and NOx Emissions from Propane, n-Butane, and Dimethyl Ether Premixed Flames"	C.A. Frye* - PSU A.L. Boehman - PSU P.J.A. Tijm - APCI
7-9 October 1998	EFI Gas Conversion Conference. San Francisco, CA	"Oxygen Production for Floating Production, Storage and Off-Loading Mounted GTL Projects"	P.J.A. Tijm* - APCI D. Miller - APCI D.M. Brown - AP-PLC R.J. Allam - AP-PLC P.G. Goldstone - AP-PLC
20-25 September 1998	5th Natural Gas Conversion Symposium, Taormina, Sicily, Italy	"Advanced Gas-To-Liquid Processes For Syngas and Liquid Phase Conversion"	E.P. Foster* - APCI P.J.A. Tijm - APCI D.L. Bennett - APCI
6-9 October 1998	Sixth Annual Technical Seminar Promotion of Clean Fossil Energy Technologies in the APEC Region	"Liquid Phase Technology Developments for The New Millenium"	W.R. Miller* - APCI P.J.A. Tijm - APCI B.A. Toseland - APCI F.J. Waller - APCI
7-9 October 1998	EFI Gas Conversion Conference. San Francisco, CA	"Liquid Phase Technology Developments for The New Millenium"	E. C. Heydorn* - APCI P.J.A. Tijm - APCI
4-7 October 1998	EPRI/GTC Gasification Technologies Conference, San Francisco, CA	"Liquid Phase Methanol (LPMEOH™) Project Operational Experience"	E.C. Heydorn* - APCI V.E. Stein - APCI P.J.A.Tijm - APCI B.T. Street - EMN R.M. Kornosky - DOE

LIQUID PHASE METHANOL AND ALTERNATIVE FUELS
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DATE	MEETING/PUBLICATION	TITLE	AUTHORS
7 January 1999	Paper from Penn State University	"The Use of Oxygenated Diesel Fuels for Reduction of Particulate Emissions from a Single-Cylinder Indirect Injection Engine"	H.S. Hess - PSU M.A. Roan - PSU S. Bhalla - PSU T. Butnark - PSU V. Zarnescu - PSU A. Boehman - PSU P.J.A. Tijm - APCI F.J. Waller - APCI
21-25 March 1999	Spring ACS Meeting, Anaheim, CA	"Syngas Conversion to Fuels and Chemicals"	B.L. Bhatt* - APCI E.C. Heydorn - APCI P.J.A. Tijm - APCI B.T. Street - EMN R.M. Kornosky - DOE
	Industrial Engineering and Chemistry Publication	"Single Step Syngas-to-DME Processes for Optimal Productivity, Minimal Emissions and Natural Gas-Derived Syngas"	X.D.Peng* - APCI A.W. Wang - APCI B.A. Toseland - APCI P.J.A. Tijm - APCI
26-28 May 1999	Combustion and Global Climate Change Conference, Calgary, Alberta, Canada	"Particulate Matter Emissions Reduction with CETANER™ as a Blend Component for Diesel Fuels with Low Cetane Levels"	F.J. Waller* - APCI P.J.A. Tijm - APCI A.S. Wilcox - AET C.D. Gray - AET G.D. Webster - AET et al
June 1999	Chemical Engineering Science	"Kinetic Understanding of the Chemical Synergy under LPDME™ Conditions - Once-through Applications"	X.D.Peng* - APCI P.J.A. Tijm - APCI B.A. Toseland - APCI
21-24 June 1999	7th Clean Coal Technology Conference, Knoxville, TN	"Commercial-Scale Demonstration of the Liquid Phase Methanol (LPMEOH™) Process: Operating Experience Update"	B.W. Diamond* - APCI E.C. Heydorn - APCI P.J.A. Tijm - APCI B.T. Street - EMN R.M. Kornosky - DOE

APPENDIX A

PROCESS FLOW DIAGRAMS

APPENDIX B

MILESTONE SCHEDULE

MILESTONE SCHEDULE STATUS REPORT
LIQUID PHASE METHANOL DEMONSTRATION
DE-FC22-92PC90543

Task Name	Duration in Months	Start Date	End Date	% Com	% Sched	9 3	9 4	9 5	9 6	9 7	9 8	9 9	0 0	0 1
PHASE 1: DESIGN	**78.00 mon**	**10/1/1993**	**3/31/2000**	**98**	**98**									
PROJECT DEFINITION (TASK 1)	16.78 mon	10/1/1993	9/30/1994	100	100									
CONTINUATION APPLICATION (B.P. #2)	0.41 mon	8/2/1994	8/10/1994	100	100		X							
PERMITTING (TASK 2)	47.31 mon	11/17/1993	9/10/1996	100	100									
NEPA FONSI APPROVAL		6/30/1995	6/30/1995	100	100			X						
DESIGN ENGINEERING (TASK 3)	38.62 mon	4/15/1994	8/1/1996	100	100									
VENDOR ENGINEERING	33.15 mon	8/10/1994	7/30/1996	100	100									
OFF-SITE TESTING (TASK 4)	73.13 mon	2/25/1994	3/31/2000	94	94									
UPDATED FUEL TEST PLAN APPROVAL		8/29/1997	8/29/1997	100	100					X				
DECISION TO CONTINUE DME TESTING		12/4/1996	12/4/1996	100	100				X					
PLANNING, ADMIN & DME DVT (TASK 5)	55.54 mon	10/1/1993	1/20/1997	100	100									
PHASE 2: CONSTRUCTION	**65.44 mon**	**10/17/1994**	**3/31/2000**	**99**	**99**									
PROCUREMENT (TASK 1)	30.02 mon	10/17/1994	7/30/1996	100	100									
CONSTRUCTION (TASK 2)	22.44 mon	10/2/1995	1/31/1997	100	100									
TRAINING AND COMMISSIONING (TASK 3)	24.92 mon	9/5/1995	2/27/1997	100	100									
OFF-SITE TESTING (TASK 4)	32.00 mon	8/1/1997	3/31/2000	96	96									
PLANNING AND ADMINISTRATION (TASK 5)	34.02 mon	6/1/1995	5/1/1998	100	100									
CONTINUATION APPLICATION (B.P. #3)	4.18 mon	5/31/1996	10/8/1996	100	100									
PHASE 3: OPERATION	**59.24 mon**	**1/20/1997**	**12/28/2001**	**66**	**66**									
START-UP (TASK 1)	3.22 mon	1/23/1997	4/2/1997	100	100									
METHANOL OPERATION (TASK 2.1)	47.84 mon	4/2/1997	3/28/2001	69	69									
DISMANTLE PLANT (TASK 2.3)	6.90 mon	6/1/2001	12/28/2001	0	0									
ON-SITE PRODUCT USE DEMO (TASK 3)	2.14 mon	11/3/1997	1/7/1998	100	100									
OFF-SITE PRODUCT USE DEMO (TASK 4)	31.97 mon	12/1/1997	7/31/2000	87	87									
DATA ANALYSIS/REPORTS (TASK 5)	56.02 mon	1/20/1997	9/21/2001	63	63									
PLANNING AND ADMINISTRATION (TASK 6)	59.24 mon	1/20/1997	12/28/2001	60	60									
PROVISIONAL DME IMPLEMENTATION	**51.12 mon**	**4/1/1997**	**7/5/2001**	**47**	**47**									
DME DVT (PDU TESTS) (TASK 3.6)	31.61 mon	4/1/1997	11/19/1999	100	100									
DECISION TO IMPLEMENT	22.47 mon	3/1/1998	1/14/2000	0	0									
DESIGN, MODIFY & OPERATE (TASK 3.2.2)	36.14 mon	7/1/1998	7/5/2001	0	0									

APPENDIX C

WORK BREAKDOWN STRUCTURE

Work Breakdown Summary			
Phase I	Task 1	1.1	Project Definition
	Task 2	1.2	Permitting
	Task 3	1.3	Design Engineering
	Task 4	1.4	Off-site Testing (Definition and Design)
	Task 5	1.5	Planning, Administration and DME Verification Testing
Phase II	Task 1	2.1	Procurement
	Task 2	2.2	Construction
	Task 3	2.3	Training and Commissioning
	Task 4	2.4	Off-site Testing (Procurement and Construction)
	Task 5	2.5	Planning and Administration
Phase III	Task 1	3.1	Startup
	Task 2	3.2	LPMEOH™ Process Demonstration Facility Operation
	Task 2.1	3.2.1	Methanol Operation
	Task 2.2	3.2.2	DME Design, Modification and Operation
	Task 2.3	3.2.3	LPMEOH™ Process Demonstration Facility Dismantlement
	Task 3	3.3	On-site Testing (Product Use Demonstration)
	Task 4	3.4	Off-site Testing (Fuel Use Demonstration)
	Task 5	3.5	Data Collection and Monitoring
	Task 6	3.6	Planning and Administration

APPENDIX D

PLOT PLAN AND EQUIPMENT ARRANGEMENT DRAWINGS

APPENDIX E

PHOTOGRAPHS

KINGSPORT - ERECTING THE LPMEOH™ REACTOR VESSEL - JULY 1996

KINGSPORT LPMEOH™ FACILITY - MAY 1997