

9. Gas Use in Chemicals and Power Production -- Economic Analysis

The use of gas in chemical feedstock and in power generation applications is established. This study has analyzed the market for chemical products to assess demand growth and the potential for increased gas usage. Likewise, the issue of increased natural gas use in the power sector was addressed. In each of these applications use of low quality natural gas (LQNG) was compared with use of high quality natural gas.

The economics are summarized in Exhibit 9-1. As can be seen, using high quality natural gas, methanol can be produced at a price well below the current market price, but the market is currently very tight. More importantly, future prices are expected to support an acceptable return on investment. On the other hand ammonia prices, which are currently high, still do not cover the return to capital. Market growth and U.S. facility economics provide a consistently bleak outlook for increasing natural gas use for ammonia production.

Use of gas is increasing as a fuel for power generation and the favorable economics, a 3.7 cent per kilowatt hour cost, supports that growth.

In all of these cases, use of LQNG as a feed was compared with the use of high quality gas. The low quality gas considered contained 13% N₂. A high nitrogen gas was analyzed because it is more costly to remove N₂ than CO₂ when upgrading a low quality natural gas to a high quality gas. In the methanol case, assuming that the methanol product would sell at the same price when using the LQNG feed as when using high quality gas, the value of the LQNG as a feed would be 2.27 \$/MMBTU compared to 2.43 \$/MMBTU for the high quality gas. For the ammonia case, the value is 2.24 \$/MMBTU compared to the 2.43 \$/MMBTU, and in the combined-cycle case, the LQNG is valued at 2.32 \$/MMBTU. That is a range of 11-19 ¢/MMBTU lower value for the LQNG compared to high quality natural gas. Those numbers compare very favorably to upgrading the LQNG via nitrogen rejection which would place a value of only 1.71 \$/MMBTU on the LQNG. These results make direct use of LQNG appear to be the path to pursue. There is, however, an important element of the economics not considered -- the transportation of LQNG to the production facility, or transport of the product to customer. In the case of power, it is expensive to transport power, if it can not be marketed locally. In the case of ammonia, there is no need for new production facilities. For methanol, the product is needed on the Gulf Coast and little high N₂ gas production is located in the area.

Thus, while the economics for LQNG look favorable, there are limited opportunities because of market and production facilities locations.

Exhibit 9-1. Economic Comparison of Natural Gas To Chemicals and Power Processes

Process	Liquid Phase Methanol (LPMEOH)		Ammonia		Combined Cycle		Cryogenic Nitrogen Production
Product	Methanol		Ammonia		Power		High Quality Natural Gas
Type of Natural Gas Feed	High Quality	Low Quality	High Quality	Low Quality	High Quality	Low Quality	Low Quality
Costs (dollars per unit product)							
Name of Unit	Gallon	Gallon	Short Ton	Short Ton	kWh	kWh	10⁶ Btu
Feed	0.24	0.24	50	46	0.019	0.018	1.58
Other Operating	0.07	0.07	80	83	0.005	0.006	0.31
Capital	0.12	0.12	57	59	0.011	0.012	0.24
TOTAL	0.44	0.44	240	240	0.037	0.037	2.43
Price of Natural Gas Feed, \$/10⁶ Btu*	2.43	2.27	2.43	2.24	2.43	2.32	1.71
Market Price Range of Product, 10/93 - 10/94, \$/Unit	0.46 to 1.40		104 to 232		0.046 to 0.052		

* The price of the low quality natural gas is the value the LQNG would have in producing the same price product as determined for the high quality gas feed case.

9.1. Natural Gas to Methanol by the Liquid-Phase Methanol Process

The liquid-phase methanol (LPMEOH) process was invented by Chem Systems Inc. In a report by Chem Systems it is compared to a version of the more conventional gas-phase processes. The biggest difference between the two processes is in the method of removing the heat of reaction.

Due to its ability to absorb the methanol synthesis heat of reaction in an inert hydrocarbon liquid and transfer this heat to generate steam, the LPMEOH process is capable of achieving a high conversion per pass. Thus, it has economic advantages when combined with larger single-train natural gas-based plants that utilize advanced technology for generating synthesis gas.

The following section will give details of this process. The process description, major reactions and material balance are the same as previously shown in Section 8.1. They are repeated here to help introduce the comparison of the economics of methanol using high and low quality natural gas. Other methods of converting natural gas to methanol were reported in Section 8.1 and will not be repeated here.

LPMEOH Process Description

Exhibit 9-2 shows the LPMEOH process flow diagram. The LPMEOH process converts synthesis gas produced from natural gas to methanol. Alternatively, the process could use syngas from coal. The usual design incorporates generation of synthesis gas (H_2 , CO and CO_2) from natural gas by steam reforming.

An advanced synthesis gas generation process (not shown) includes both quench and convective trains to achieve maximum thermal efficiency in the methanol production. The synthesis gas is then fed to a three-stage LPMEOH system. The first reactor operates as 'once-through' and the last two reactors in series with converted syngas recycled to the second stage reactor. The methanol reactor operates at 1600 psia and a $482^{\circ}F$.

Major Reactions

The synthesis gas contains, H_2 , CO and CO_2 along with inert gases such as CH_4 and N_2 . The key reactions which take place in the reactor are:

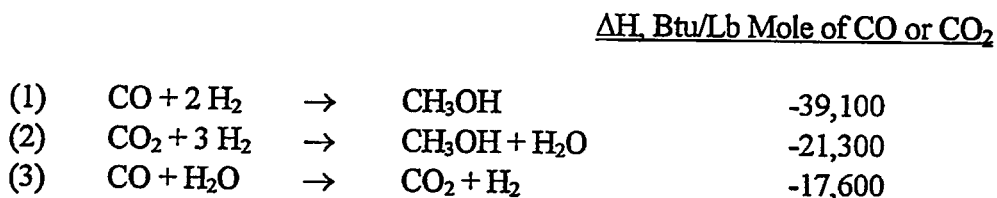
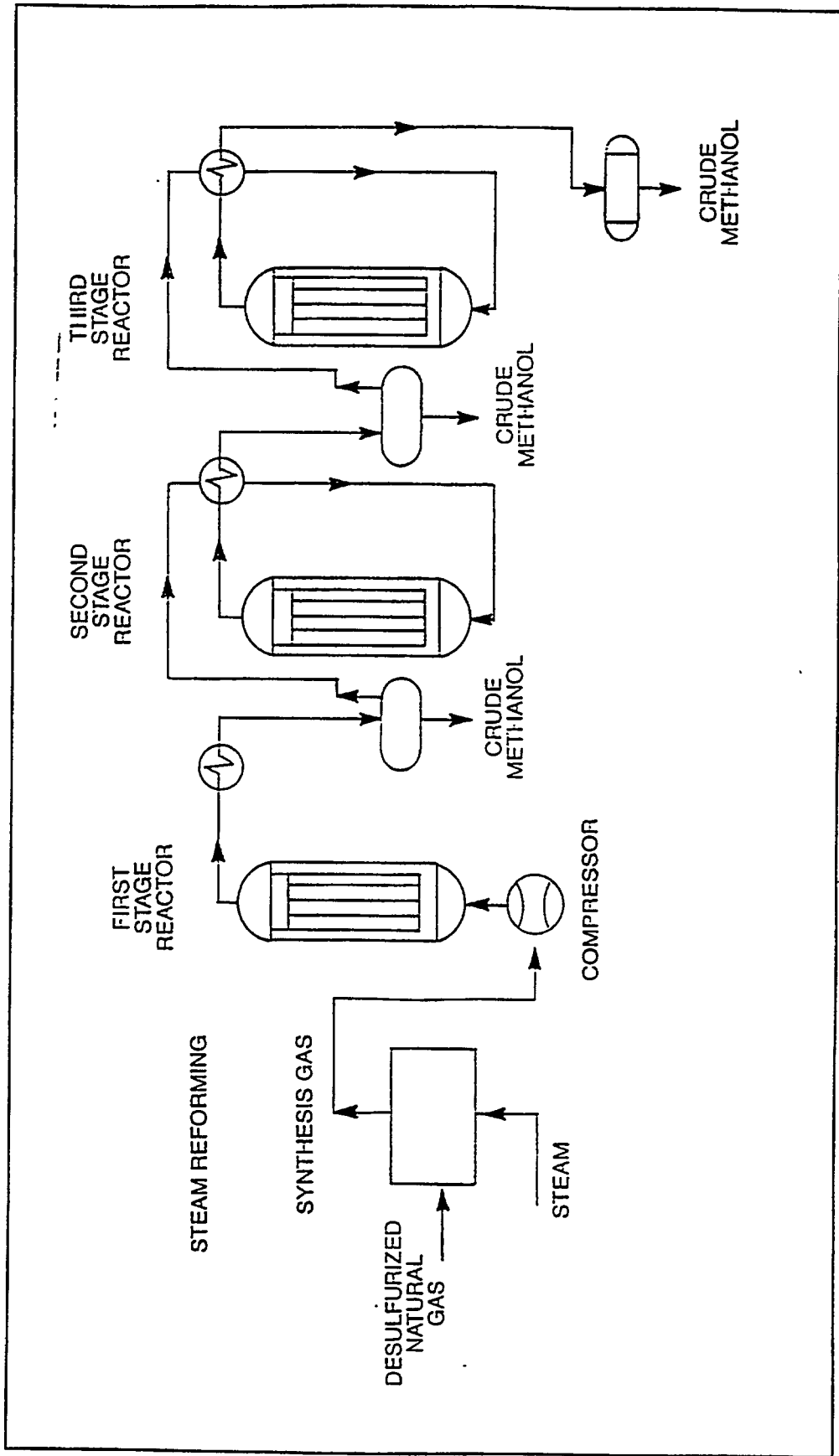


EXHIBIT 9-2
LPMEOH PROCESS FLOW DIAGRAM



Material Balance

According to the report on LPMEOH [29] (page 69, Table III - B-3), when charging a feed consisting of 582,000 lb/hr of syngas to the LPMEOH section, the products consist of 525,000 lb/hr of crude methanol and 58,000 lb/hr of fuel gas. The syngas has the following composition in mol %: 30% carbon monoxide, 3% carbon dioxide, 65% hydrogen and 2% other gases. The fuel gas produced consists of 5% carbon monoxide, 11% carbon dioxide, 17% argon, 14% nitrogen, 50% hydrogen and 3% other gases.

Economics

The following economics table for the liquid phase methanol process is based on the process proposed by Chem Systems when charging high quality natural gas.

Plant Capacity, BPSD of Methanol	39,572
Total Capital, 10 ⁶ US\$ (USGC, December 1993)	451.4
	<u>\$/B Methanol Product</u>
Natural Gas Feed, \$2.43/10 ⁶ Btu	10.08
Operating & Maintenance Costs	2.73
Capital Charges, 14.2 % of Total Capital	4.91
Product Price, \$/Barrel	18.65
\$/Gallon	0.44

Economics were also obtained for methanol production charging low quality gas. This gas contained 13% nitrogen. The previous set of economics obtained when charging high quality natural gas was used as a starting point. Economics in a U.S. Department of Energy report [113] shows economics for methanol production using high quality natural gas and low quality natural gas containing 30% nitrogen. These economics were interpolated to get the effect on the economics of going from high quality natural gas feedstock to 13% nitrogen content low quality natural gas.

The economics table for methanol production when charging low quality natural gas is as follows. The feedstock price is adjusted to give the same methanol product price as for the economics using high quality natural gas as shown above.

Plant Capacity, BPSD of Methanol	39,572
Total Capital, 10 ⁶ US\$ (USGC, December 1993)	469.4
	<u>\$/B Methanol Product</u>
Natural Gas Feed, \$2.27/10 ⁶ Btu	9.87
Operating & Maintenance Costs	2.86
Capital Charges, 14.2 % of Total Capital	5.12
Product Price, \$/Barrel	18.65
\$/Gallon	0.44

The resulting value of calculated for the low quality gas in this process is \$2.27/MMBtu as compared to the \$2.43/MMBtu for the high quality natural gas.

Other Methanol Process Routes and Effect of Cost Factors on Product Costs

Information on "Other Methanol Process Routes" and "Effect of Cost Factors on Product Costs" are covered as part of Section 8.1.

Summary

The process description, major reactions, material balance and economics have been shown for the Liquid Phase Methanol (LPMEOH) process. The LPMEOH process has never been experimentally tested and there is a need for a demonstration plant similar to the tests of coal-based power plants carried out under the Clean Coal Program. The LPMEOH process is economically competitive or somewhat better than commercial methanol plants. Commercial methanol plants are in a high state of development. Nevertheless, there is still room for small improvements in the methanol processes and a potential advantage for building larger, more economical plants.

Economics have been shown for producing methanol when charging both high and low quality natural gas. The use of low quality gas requires an expansion of process equipment dimensions and the use of additional feedstock compared to the use of high quality natural gas. Thus methanol can be made economically if the price of low quality feedstock is greater than 16 cents per million Btu less than the price of high quality natural gas. In this case the low quality gas contains 13% nitrogen. The problem with use of this low quality gas for methanol production is that it is available in the parts of western U.S. where methanol needs are minimal.

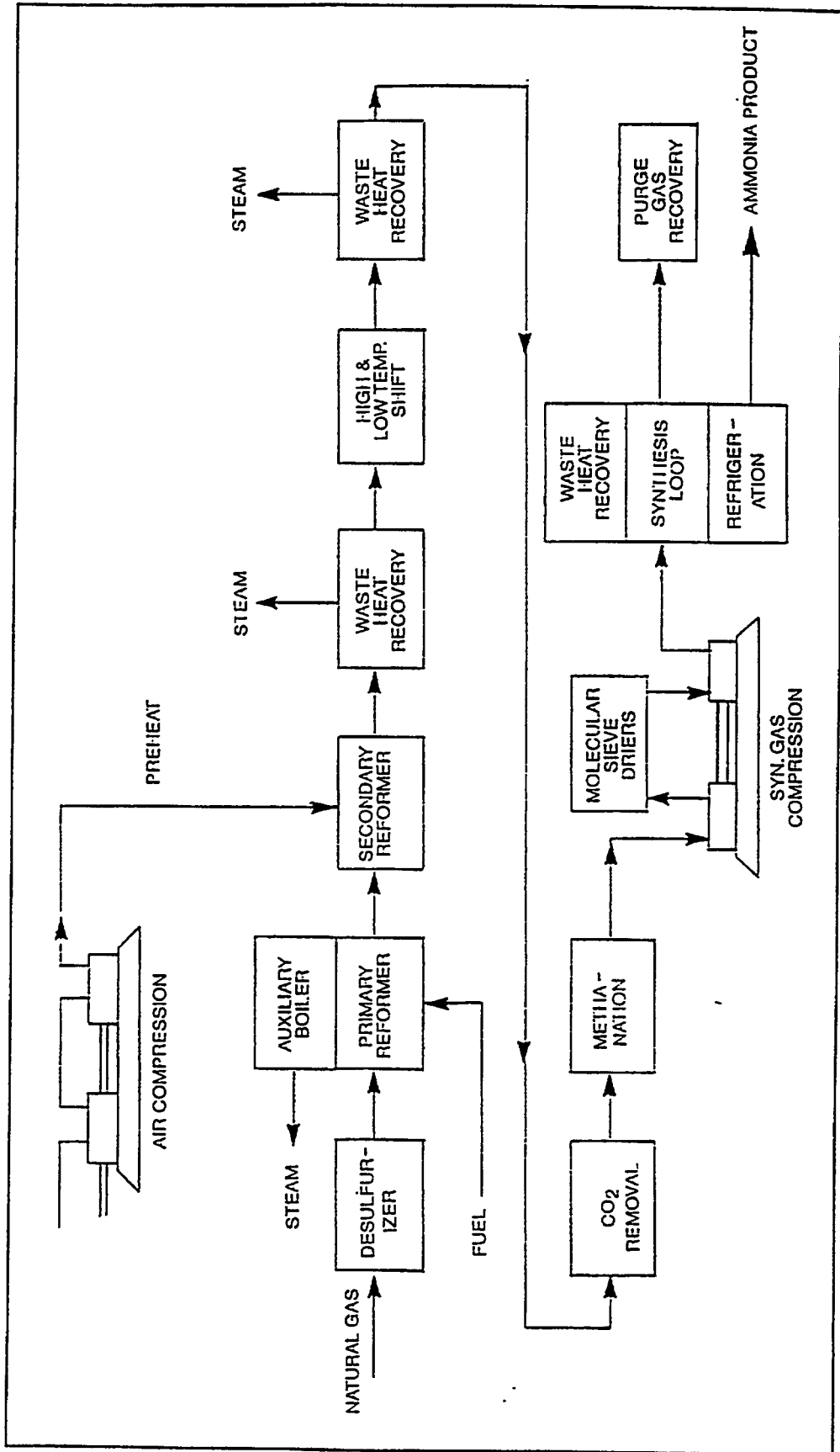
9.2. Natural Gas to Ammonia

The M. W. Kellogg Company is the world's leader in terms of the number of ammonia plants built. These plants have been built principally to supply fertilizer for use in agriculture. Like the production of methanol, ammonia production requires the production of synthesis gas as a first step. The following section will give details of this process as developed by M. W. Kellogg. Variations in ammonia production technology will also be discussed.

Process Description

Natural gas is desulfurized and is then fed to a primary steam reformer which converts 70% of the natural gas into a raw synthesis gas. In the secondary reformer, air is introduced to supply the nitrogen. The heat of combustion of the partially reformed gas supplies the energy to reform the remainder of the natural gas feed.

**EXHIBIT 9-3
AMMONIA PROCESS**



Economics were also obtained for ammonia production when charging low quality gas. This gas contained 13 volume percent nitrogen. The previous set of economics obtained when charging high quality natural gas was used as a starting point. In calculating the investment, it was assumed that only the steam reforming and Rankine cycle investment and attendant offsites and Exhibit 9-3 utilities were increased by the greater flow rate of the low quality gas compared to high quality gas. A DOE report [113] was used to determine the capacity exponent used in the investment calculation.

Likewise, for calculating operating and maintenance costs, only the reforming and Rankine cycle sections were assumed to be affected by the change in gas quality. An Oil and Gas Journal Article [116] was used to estimate the effect on the change in natural gas requirement. The basis used was that one third of the natural gas was used for fuel and the rest for feedstock. The fuel rate but not the feedstock rate was assumed to be affected by a change in feedstock quality.

The economic table for ammonia production when charging low quality natural gas is as follows. The feedstock price is adjusted to give the same ammonia product price as for the economics using high quality natural gas. The calculated value of the LQNG is \$2.24/MMBtu as shown below.

Plant Capacity, 10 ³ Short tons/year of Ammonia	420
Total Capital, 10 ⁶ US\$ (USGC, December 1993)	154.9
	<u>\$/Short Ton - Ammonia Product</u>
Natural Gas Feed, \$2.24/10 ⁶ Btu	45.69
Operating and Maintenance Costs	82.73
Capital Charges, 14.2% of Total Capital	59.06
Product Price, \$/Short Ton	240.33

Variations of Ammonia Production Routes

Improvements have been made in the ammonia process mainly to decrease the energy consumption requirement. The M.W. Kellogg Company [117] reports that older ammonia plants had a cost of energy which was 80-90% of the selling price of ammonia. In the future, they expect that energy costs will still account for 50% of the product ammonia selling price. This represents a decrease of energy usage from 40 to less than 25 MMBtu/ST. Their strategies in new plant design included:

- Reduce power consumption
- Improve process heat recovery
- Reduce CO₂ removal energy consumption
- Improve power cycle efficiency
- Reduce stack gas losses

These improvements can be utilized not only in new plant designs but also in retrofitting existing plants.

KTI Corp. [116] has developed a process for producing ammonia in smaller plants in the 200 to 600 ton per day range in a more efficient manner. It is called the PARC process and includes a number of innovative features in its design. They claim a payout time of 1 to 2 years for natural gas valued at 5.54 \$/MMBTU. This is considerably higher than current natural gas price. Therefore the implementation of these energy saving design changes does not appear to be attractive at this time.

Pullman Kellogg [114] has compared the costs of producing ammonia from natural gas, naphtha, fuel oil and coal at 1979 prices and determined that natural gas was the most economical feedstock. Their economic methods could be applied to any sets of feedstock prices to determine comparisons at present and future times.

Synthesis production is a large part of the total cost of ammonia production, and so new improved methods of synthesis gas production should be examined. These methods were discussed in Section 8.1.

Hydrocarbon Processing [118] shows flow diagrams for licensors who have processes for producing ammonia from hydrocarbon feedstocks. The companies involved are:

- Brown & Root Braun with twenty-five single-train plants.
- Two ammonia plants have been built by the LCA process which has advantages at small scale such as 500 tpd. ICI licenses this technology to three contractors: Chiyoda Corporation of Japan, Davy McKee of England and Mannesmann/KTI of Germany and the Netherlands.
- Two plants have been built using the AMV process licensed by ICI Katalco.
- large-scale, single-train ammonia plants are on stream or have been contracted for throughout the world with M. W. Kellogg Company as the licensor.

Where both methanol and ammonia products are needed, it would be cost effective to produce synthesis gas from a single process to supply feed for both processes.

Summary

The process description, major reactions and economics have been shown for the ammonia process using M. W. Kellogg methodology. The ammonia process technology is mature and it is not likely that many new process units will be built in the U.S. in the near future, because of stagnant market demand. There are new process improvements which could be applied to new and existing units if energy prices were high enough to justify them.

Economics have been shown for producing ammonia when charging both high and low quality natural gas. The use of low quality gas would cause an increase in equipment size and the use of more energy than when charging high quality natural gas. The ammonia can be made economically if the price of low quality feedstock is greater than 24 cents per million Btu less than the price of high quality natural gas. In this case the low quality gas contains 13% nitrogen. The problem with use of this low quality gas for ammonia production is that it is in the parts of the western U.S. where the demand for ammonia is small.

9.3. Natural Gas for Power Production Using a Combined Cycle

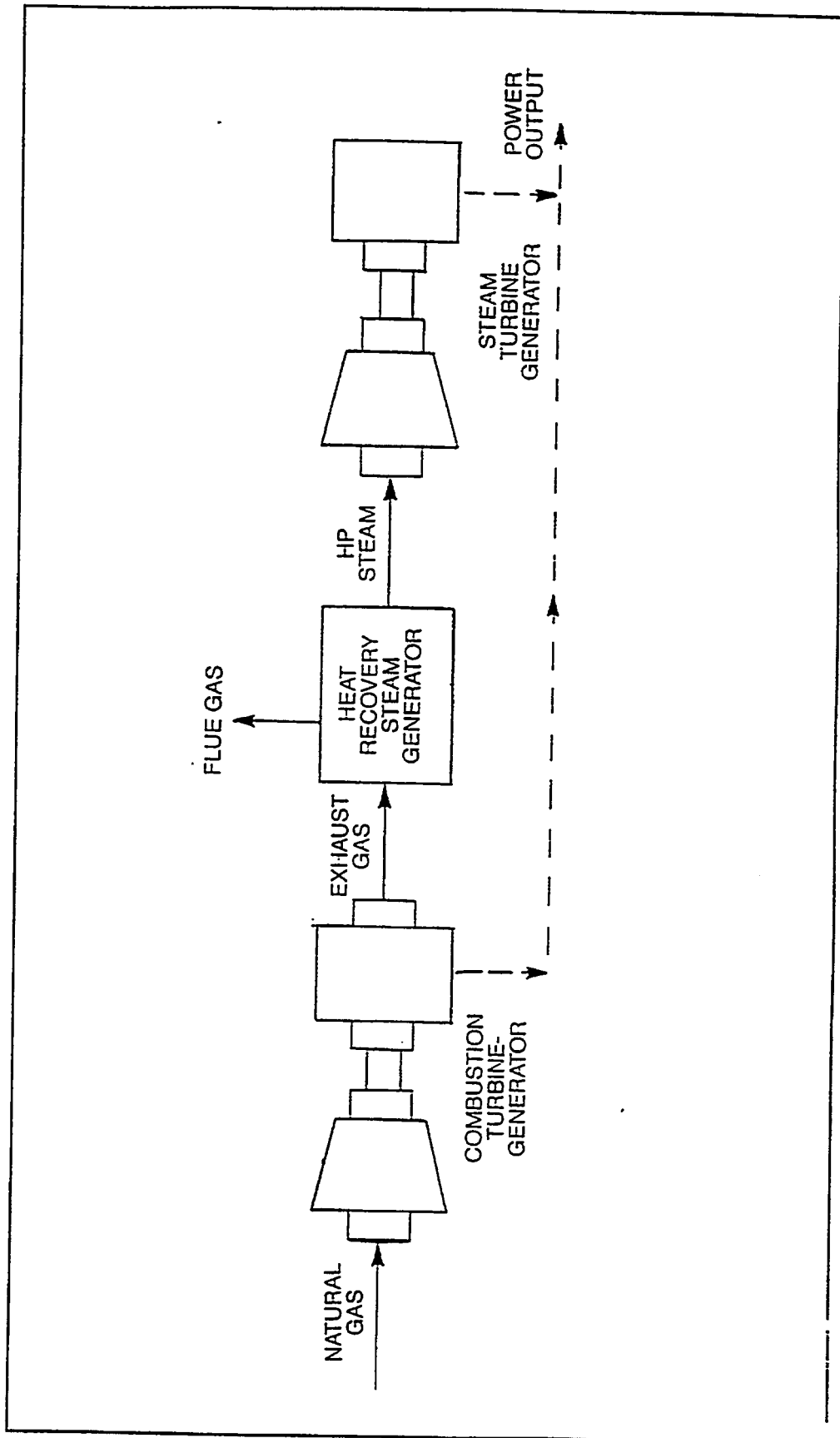
The term "combined cycle" refers to power cycles with a gas turbine and steam turbine combined in a series arrangement in which the exhaust from the gas turbine flows directly through a heat recovery steam generator (HRSG). Steam produced in the HRSG is used to drive the steam turbine. Electric generators, driven by both the gas and steam turbines, produce electric power for the grid. Combined cycle power plants, fueled by both natural gas and oil, have been in common use for power generation by electric utilities since the 1960s.

The power plant described in this section is rather large (426 MW) and is therefore quite efficient both because of its size and its use of modern technology. The following section will give details of this process. Information on alternate process routes and cost factors follows.

Process Description

Natural gas is fed to two General Electric PG7001F combustion turbines to generate 304 MW of power. The exhaust gas goes to a heat recovery steam generator where 1485 psig steam is produced. The steam is fed to a steam turbine generator to generate 129 MW of additional power. The total output of power is 426 MW. The net heat rate is 7700 Btu/kWh. A simplified flow diagram is shown in Exhibit 9-4.

**EXHIBIT 9-4
COMBINED-CYCLE POWER PLANT**



Economics

The following economics table is for a combined cycle power production facility charging high quality natural gas. It is based on the 426 MW case for Gilbert Station Site Alternate Case 2 from an EPRI study prepared by Jersey Central Power and Light Company and Sargent and Lundy [119].

Plant Capacity, MW of Electricity	426
Total Capital, 10 ⁶ US\$ (USGC, December 1993)	231.8
	<u>\$/kWh Electricity Product</u>
Natural Gas Feed, \$2.43/10 ⁶ Btu	0.019
Operating and Maintenance Costs	0.005
Capital Charges, 14.2% of Total Capital	0.011
Electricity Product Price, \$/kWh	0.037

Economics were also obtained for power production when charging low quality natural gas. This gas contained 13 volume percent nitrogen. The previous set of economics obtained when charging high quality natural gas was used as a basis for the price of electricity product. Based on a conversation with Peter Roberts of Solar [120] the mechanical part of the investment and maintenance costs were assumed to be 10% higher when charging low quality gas than when charging high quality gas. The increase in natural gas and chemical costs for a case charging low quality gas compared to the case charging high quality gas is based on an increase in gas volume. The increased gas volume is estimated from material balance data from a Virginia Power EPRI study [121]. The feedstock price is adjusted to give the same electricity product price as for the economics using high quality natural gas.

Plant Capacity, MW of Electricity	426
Total Capital, 10 ⁶ US\$ (USGC, December 1993)	246.4
	<u>\$/kWh Electricity Product</u>
Natural Gas Feed, \$2.32/10 ⁶ Btu	0.018
Operating and Maintenance Costs	0.006
Capital Charges, 14.2% of Total Capital	0.012
Electricity Product Price, \$/kWh	0.037

Alternate Process Routes and Cost Factors

Jersey Central Power & Light and Sargent and Lundy [119] have compared two methods of converting natural gas to power. These are steam - injected gas turbine units and combined cycle plants. Evaluations were made for plants located on a fresh water site and on a sea water site. For larger plants, 150 to 400 MW, the combined cycle units showed a distinct economic advantage and would definitely be preferred over the steam - injected gas turbines. Economics for smaller units were

much closer. The report also shows that there is a considerable advantage in terms of cost of electricity for plants at the 350 to 400 MW level compared to smaller plants.

Virginia Electric and Power Company [121] has shown a three-phase combined cycle power plant. Phases 1 and 2 consist of natural gas or distillate-fired combined cycle units. The third phase adds coal gasification facilities. Their conclusion was that (1) the price of natural gas and distillate would have to escalate significantly and (2) low price, high sulfur coal would have to be available to make the gasification plant installation economically attractive where coal is charged.

Summary

A process description and economics have been shown for power production using a combined cycle with natural gas as a feedstock. The technology is mature. Future improvements may be made in gas turbine technology or more efficient use of energy but whether the use of gas, fuel oil or coal as a feedstock is promoted in the future depends on the trend of relative price increases for these feedstocks.

Economics have been shown for power production when charging both high and low quality natural gas. The use of low quality gas would cause an increase in equipment size and energy usage and would require some equipment modifications to handle the low Btu gas but technologies exist to accomplish this. The power can be made economically if the price of low quality feedstock is greater than 11 cents per million Btu less than the price of high quality natural gas. In this case the low quality gas contains 13% nitrogen. The problem with use of this low quality gas for power production is that it is available in the parts of the western US where the demand for power is small.

9.4. Upgrading of Low Quality Natural Gas by Nitrogen Rejection

Since there are large volumes of the natural gas which is of low quality and contains excessive amounts of carbon dioxide, nitrogen and hydrogen sulfide, it must be upgraded by eliminating these contaminants in order to become pipeline quality gas. In this section, only the elimination of nitrogen will be considered to illustrate the contaminant rejection problem.

The following section will give details of the process for removal of the nitrogen from subquality gas by the cryogenic process. Later in the section, nitrogen rejection by other process routes will be discussed.

Process Description

Subquality natural gas is first fed to a compressor where the gas is raised from 315 to 610 psia at the front of the plant to furnish energy to produce a refrigeration effect by gas expansion downstream. Before the gas reaches the low temperature part of the process, the recompressed gas is scrubbed in a column with an aqueous solution of diethanolamine (DEA) to remove CO₂. Dehydration takes place

next in another column packed with molecular sieves. These two impurities are removed before hand to avoid solidification in downstream equipment.

Following dehydration, the gas stream is chilled by the returned N₂ and CH₄ stream to -143°F, 587 psia to condense 85% of the inlet stream. The prechilled stream is expanded through a Joule-Thomson valve where the pressure of the stream is reduced to 347 psia and the temperature to -168°F. The expanded stream enters near the bottom of the first column in which a preliminary separation produces a substantially pure N₂ stream overhead and a bottom stream containing the remaining N₂ and hydrocarbons. This preparation step rearranges the feed stream composition to match the expected composition profile in the second column, thereby improving separation efficiency and reducing energy requirements. The pre-separated top and bottom streams are separately flashed into the second column at 45 psia and 47 psia, respectively. The final separation in the second column produces a rejected N₂ stream overhead and a CH₄ - rich bottom product stream. After heat exchanges to recover refrigeration, the rejected N₂ stream is vented to the atmosphere. The sales gas stream is recompressed to 315 psia and cooled to 120°F before it leaves the battery limits. A flow diagram for the process is shown in Exhibit 9-5.

Economics

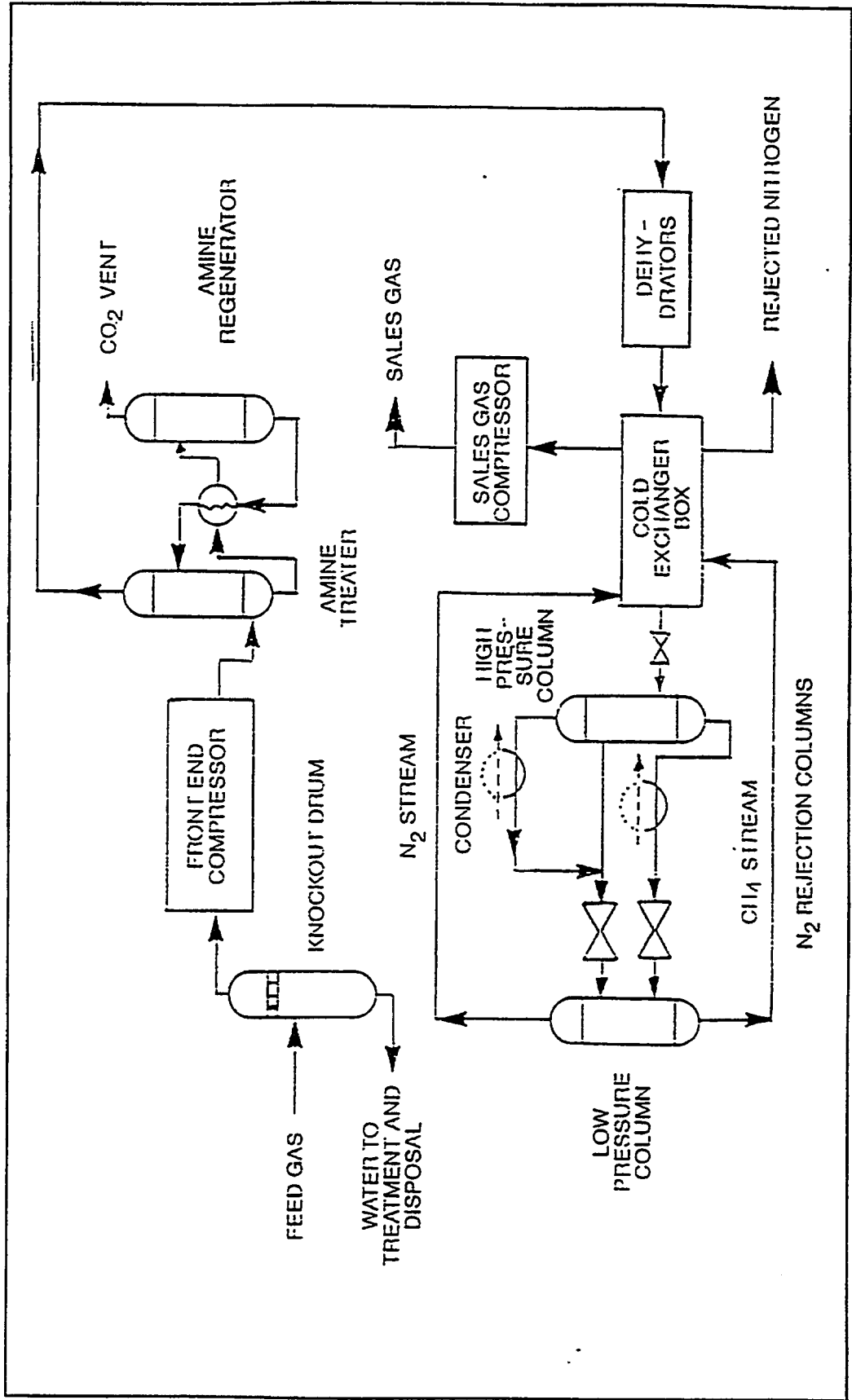
The following economics table is for nitrogen rejection from a low quality natural gas containing 13.7% nitrogen. The economics are based on the cryogenic portion of a report by James Ma of SRI International for GRI [122]. The low quality gas feedstock price is adjusted to give a high quality gas price after nitrogen rejection of \$2.43 per million Btu, the value used for high quality natural gas in this study. The resulting calculated value of the LQNG is \$1.71/MMBtu as shown below.

Plant Capacity, Billion Btu/Yr	5932
High Quality Natural Gas	
Total Capital, 10 ⁶ US\$	10.33
(USGC, December 1993)	
	<u>\$/10⁶ Btu Upgraded Natural Gas</u>
Natural Gas Feed, \$1.71/10 ⁶ Btu	1.579
Operating and Maintenance Costs	0.312
Capital Charges, 14.2% of Total Capital	0.242
Product Price of Upgraded Natural Gas, \$/10 ⁶ Btu	2.431

Other Nitrogen Rejection Process Routes

A study by Pullman Kellogg compares the economics of small scale nitrogen removal from natural gas [123]. The processes compared were cryogenic, Nitrotec's PSA and lean oil nitrogen removal processes. The cryogenic process had the lowest processing costs but the difference between PSA and cryogenic processes narrows as the nitrogen content approaches 15-25%. The lean oil method processing cost is high compared to both the cryogenic and PSA processes. The report verifies that

EXHIBIT 9-5
NITROGEN REJECTION FROM SUBQUALITY
NATURAL GAS BY A CRYOGENIC PROCESS



nitrogen removal from natural gas is expensive and several avenues for improving nitrogen removal technologies are identified. One important factor for improving economics is to develop the capability of the gas driven engines to accept lower heating value fuel gas.

The Mehra Process for nitrogen rejection has been compared to a cryogenic process for nitrogen rejection of a subquality natural gas [122]. The Mehra technology is a solvent absorption process. Based on results obtained for a 20 million Scfd capacity nitrogen rejection system, the Mehra technology was estimated to require 12% less total fixed capital investment than a cryogenic process. The cost savings by the Mehra technology were estimated at 3 cents per million Btu of upgraded natural gas. In addition, the Mehra process has advantageous qualitative features.

According to GRI [124] a PSA gas separation unit is ideal for small to mid-sized nitrogen rejection (i.e., up to 20 million scf/day of feed). The small amount of mechanical equipment and simple operation makes unattended operation possible. This unit uses carbon molecular sieves.

Summary

A process description and economics have been shown for a process which removes the 13% nitrogen content of a low quality gas using a cryogenic process. The cryogenic process is the recognized method for nitrogen rejection for large natural gas rates whereas the PSA method has advantages for low natural gas rates. It appears that the newer Mehra process has advantages over the cryogenic process but the Mehra process has not been experimentally tested and its capital cost advantage is partially offset by higher operating costs.

The nitrogen rejection process requires that low quality gas cost must be at least 72 cents per million Btu less than high quality gas. Use of this process makes it possible to convert the natural gas to pipeline quality. However, it is much more economical to use the low quality gas for chemical or power production, if that option is available.

9.5. Economics of Conversion of Natural Gas to Ethylene

The economics of converting natural gas to gasoline by direct conversion of methane was discussed in Section 8.3. In that analysis, it was shown that oxidative coupling, oxyhydrochlorination and partial oxidation were the most promising routes for direct conversion of methane but that none of them had sufficiently good selectivity and conversion to make commercial development economically feasible. Since conversion of methane to C_2 s and other hydrocarbons is a first step in the processes described in Section 8.3, much of what was said there would also apply to conversion of natural gas to ethylene. However, in this section some further discussion of economics will be given which concentrates on the conversion of methane by direct conversion to ethylene as the final product.

Parkyn et al [79] have reviewed the literature containing the economics of the direct conversion of methane to ethylene. Their general conclusion was that the technology is not ready to challenge syngas-based processes. In addition, the review explored novel approaches to methane conversion

which they believed might ultimately achieve the goal of commercialization of the direct conversion of methane. The economics reviewed included:

- Kuo and Ketkar did some studies for the U.S. Department of Energy [125] to screen direct methane conversion processes. They concluded that oxidative coupling was the most promising process. However, they said that ethylene production via oxidative coupling would only become competitive over traditional processes at conversion levels above 35% and selectivities above 88%.
- Union Carbide and Plastics Company Ltd. carried out a study on the effect that various operating parameters would have on the economic feasibility of a process based on oxidative coupling [126]. They recommended a process configuration using a demonstrated catalyst which would give low conversion (18%) but high selectivity (77% to C₂'s). This process is not yet commercially competitive. They believed that future catalyst improvements and high selectivity were the keys to success.
- Lee and Aitani [127] gave the economics for conversion of methane to ethylene at 65% selectivity using a LiCl/MnO catalyst operating at 2 bar and 750°C. They calculated that ethylene could be produced at 530 \$/tonne based on 1989 U.S. natural gas prices. These results are the most optimistic encountered even though this ethylene price is still higher than recent market prices.
- Another encouraging study was included in a report covering the work of Geerts et al [128] on Li/MgO coupling catalysts. They combine a coupling reactor followed by an ethane cracking unit. They showed that ethylene could be produced at 500 \$/tonne at something above 65% C₂ selectivity and 38% methane conversion.
- Menshchikov et al [129] found the optimum methane conversion level to be 40% and the C₂ selectivity to be 50% using coupling followed by pyrolysis. Even though the ethylene cost by this process was \$679/tonne, this USSR based process development has led to construction of a pilot plant.
- G. Renesme et al [130] in a review article concluded that the environment for oxidative coupling was unfavorable in North America but that it was more promising in Europe and Japan where ethane is already the favored feedstock for steam cracking. They favor the IFP process for direct conversion of methane to ethylene at oil prices above \$16/bbl.
- Parkyns et al [79] recommends that future work concentrate on catalyst development, mechanisms of surface reactions and gas separations.

A. D. Little has also reviewed the potential for developing a process for obtaining ethylene from methane by direct conversion [131]. Some of their conclusions are as follows:

- The C_1 process must have raw material advantages to be commercial since capital costs are high.
- Current economic analysis shows the potential for oxychlorination of methane and oligomerization of methanol (MTO) to become competitive against conventional ethylene production. U.S. Gulf Coast plants using these processes and producing 1 billion pounds of ethylene per year could realize a 15% ROI with an ethylene selling prices of 0.220 to 0.225 \$/lb. Of the two processes, oxychlorination has better process efficiencies because the oligomerization of methanol requires the production of methanol as an intermediate. However, the economics assume favorable yields which have not yet been demonstrated in a larger-scale process. Demonstration of process feasibility and development of these processes to commercial operation is necessary as a prelude to commercialization.
- The current limits of selectivity and conversion for oxidative coupling do not allow economic feasibility.
- A single world-scale ethylene plant based on oxychlorination would create a demand for 39.5 bcf of feedstock methane. Converting 25% of U.S. ethylene capacity to the oxychlorination process would create a yearly demand of 354 bcf of natural gas feedstock. Therefore, the potential for increasing the demand for natural gas is great.

Researchers at the Institute de Recherches sur la Catalyse in France have been studying the oxidative coupling of methane [132]. They say that while significant improvement in catalyst performance for the reaction of oxidative coupling of methane to C_2 hydrocarbons has been achieved in the last decade, C_2 yields still remain limited in view of a potential industrial application. They have done modeling and experimental work that demonstrates that membrane reactors can be used.

9.6. Economics of Conversion of Natural Gas to Methyl Chloride

In Section 8.3 the economics for converting natural gas to transportation fuels was shown when using the oxyhydrochlorination plus oligomerization process. If you stop the oxyhydrochlorination process before doing the oligomerization, methyl chloride is the resulting product. Following is a brief process description for this process.

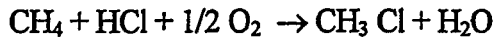
Process Description

Methane-rich gas is reacted with oxygen and hydrogen chloride in a fluid-bed reactor at 230 psia and 650°F over catalyst made of $CuCl$, KCl and $LaCl_3$. The reactants are cooled and condensed. Water containing unreacted hydrogen chloride and byproduct formic acid is removed and the HCl is recycled. The products of the oxyhydrochlorination step are primarily chloromethane and dichloromethane in a

4 to 1 ratio. The design is reported by Bechtel [42] based on research conducted at the Pittsburgh Energy Technology Center [133].

Major Reaction

The oxyhydrochlorination reaction proceeds according to the following reaction:



The net reaction is highly exothermic, with approximately 30% of the heating value of the methane feed being liberated as sensible heat.

Economics

The following table for the economics of converting natural gas to methyl chloride is based on a report by Bechtel [42]. Based on a bar chart on Page 2-4 of the Bechtel report, the operating and maintenance and capital costs for production of methyl chloride were taken as 77% of the corresponding costs of oxyhydrochlorination plus oligomerization. Feed costs were assumed to be the same as for oxyhydrochlorination plus oligomerization. The material balance is based on Table 5.1 on Page 5-3.

Plant Capacity, Tons/Day of Methyl Chloride	8112
Total Capital, 10 ⁶ U.S.\$ (USGC, December 1993)	740.8
	<u>\$/Ton Methyl Chloride</u>
Natural Gas Feed, \$2.43/10 ⁶ Btu	35.67
Operating and Maintenance Costs	34.28
Capital Charges	41.10
Product Price: \$/Ton	115.95

Process Status and Potential

A.D. Little has briefly reviewed the process for making methyl chloride from natural gas [131]. The chloromethanes such as methyl chloride are not in themselves valuable products. However, they can be converted to higher-valued olefins or hydrocarbon fuels. In Table 2-27, chlorination and oxychlorination research is summarized. Ethylene, vinyl chloride monomer and gasoline are the final products. The research has been conducted by KTI, IIT, PETC, USC and BP. Dow Corning has been doing research on a process for converting methane to methyl chloride via oxyhydrochlorination [134] and [135]. They have focused on developing a catalyst that will give stable, selective formation of methyl chloride. A Cu based catalyst system is most promising. They are also working on the design and construction of a process demonstration unit for this process.

The most serious reservations about the process are focused on costs associated with handling chlorides and disposal of chloride wastes. Concern was expressed that these problems and associated investments have not been adequately accounted for. The oxyhydrochlorination process is simple and occurs under moderate conditions. Methane conversion is high and methyl chloride selectivity is favored. This process has never been commercialized. Further process development and solutions for corrosivity problems are needed before this process can be commercialized.

10. Sensitivity Analysis of Economics in Converting Natural Gas to Fischer-Tropsch Liquids

The conversion of natural gas to Fischer-Tropsch liquids provides a method of obtaining desired transportation fuels but the cost can be relatively high. In this section, a number of parameters will be varied to show the effect on product cost. This will identify which element of cost most needs to be reduced to make Fischer-Tropsch gas-to-liquids a viable commercial process. Parameters to be examined include changing reactor type, gas price growth rate, plant size, location, use of barge mounted plants, oxygen production with ceramic membranes and natural gas price. All of the economics are based on costs before product upgrading.

10.1. Effect of Using Slurry Reactor

In Section 8.2, it was shown that investment costs and operating costs are lower for the slurry reactor than for the fixed-bed reactor for the Fischer-Tropsch process. Economics for the Fischer-Tropsch process using the slurry reactor have been obtained to better define this cost improvement. Economics obtained by Bechtel were used to determine differences in capital costs and operating and maintenance costs between Fischer-Tropsch plants with fixed-bed and slurry reactors [49]. Manpower was assumed to be the same for both reactor designs. The economics for obtaining Fischer-Tropsch liquids from natural gas using the slurry reactor are as follows:

Plant Capacity, BPSD		14,500
of Fischer-Tropsch Liquid		
Total Capital, 10 ⁶ U.S.\$		749.3
(USGC, December 1993)		
	<u>\$/B Fischer-Tropsch Product</u>	
Natural Gas Feed, \$2.43/10 ⁶ Btu		20.92
Operating and Maintenance Costs		11.88
Capital Charges,		22.56
14.2% of Total Capital		
Product Price:		
\$/Barrel		56.84
\$/Gallon		1.35

These economics were obtained using a 3.1% feed price growth rate. Economics obtained at the same feed price growth rate but using a fixed bed reactor gave a product price of 59.61 \$/Barrel. Thus the use of a slurry reactor instead of a fixed-bed reactor resulted in decreasing the product cost by about \$3/barrel, a substantial saving. This difference may understate the real advantage of using the slurry reactor since the slurry reactor produces a product with a much higher proportion of gasoline than the fixed-bed reactor.

10.2. Effect of Gas Price Growth Rate

In Section 7.1, the growth price rate for natural gas, in constant dollars, for the period 1992-2010 was predicted to be 3.5 - 4.0% annually. In Section 10.1, it was shown that for a Fischer-Tropsch plant with slurry reactor design the product cost is \$56.84/Barrel when the growth rate is 3.1%. Economics will now be given for the same plant except when the growth rates are 0.6, 1.6 and 3.9%. In all cases, plant capacity and total capital stay the same so only product costs will be shown. The economics for Fischer-Tropsch plants charging natural gas at various natural gas price growth rates all starting with first year cost of \$2.43/MMBtu are as follows:

Growth Rate for Natural Gas Price, % Annually	0.6%	1.6%	3.9%
Natural Gas Feed, \$2.43/10 ⁶ Btu	20.92	20.92	20.92
Operating and Maintenance Costs	11.88	11.88	11.88
Capital Charges, 14.2% of Total Capital	22.56	22.56	22.56
Product Price:			
\$/Barrel	52.76	54.04	58.87
\$/Gallon	1.26	1.29	1.40

These results show that if the feed cost growth rate is 3.9% instead of 3.1% the Fischer-Tropsch product price is \$2/B higher. Likewise if the feed cost growth rate is 0.6% instead of 3.1%, the Fischer-Tropsch product price is decreased by \$4/B.

10.3. Effect of Plant Size

The effect of plant scale was analyzed in a case in which Fischer-Tropsch plant size was made four times larger than the one used in Section 10.1. A report by Bechtel was used to determine the effects on capital and operating and maintenance costs [31]. The Bechtel report is for the methanol process, but it was assumed that the same cost/capacity relationships could be applied to Fischer-Tropsch. This report shows that capacity exponents for process field costs of 0.96 should be used, capacity exponents for offplots field costs of 0.35 are recommended and capacity exponents for home office and engineering should be 0.47. The number of operators per shift was nearly doubled. The economics for a plant four times as large as the one in Section 10.1 but similar in other respects are as follows:

Plant Capacity, BPSD of Fischer-Tropsch Liquid	58,000
Total Capital, 10 ⁶ U.S.\$ (USGC, December 1993)	2010.6
	<u>\$/B Fischer-Tropsch Product</u>
Natural Gas Feed, \$2.43/10 ⁶ Btu	20.92
Operating and Maintenance Costs	7.47
Capital Charges, 14.2% of Total Capital	15.05
Product Price:	
\$/Barrel	45.67
\$/Gallon	1.09

The economics showed that by quadrupling the size of the plant, the product price can be decreased by \$11/B. Thus it is advantageous to build as large a plant as possible as long as the product demand remains high.

10.4. Effect of Barge Mounted Plants and Alaska North Slope Location

The cost of mounting a Fischer-Tropsch plant on a barge or locating it on the North Slope of Alaska adds substantially to the cost of producing Fischer-Tropsch Liquids. These costs largely are balanced by assuming a very low cost for the natural gas.

Two sets of economics for plants similar to the one in Section 10.1 will be given to show the effect of making different assumptions about type and location of plant on the product price. The effects of remote location and/or barge mounting are based on estimates for methanol plants but it is assumed that the same relationships would apply to Fischer-Tropsch plants.

Chem Systems did a study for the U.S. Department of Energy in which they showed the differences in capital costs, maintenance costs and labor in terms of varying location [37]. There are four categories of location. The economics shown here are based on Category IV costs compared to Category I costs in Table I-5 and I-13. Category I costs are U.S. Gulf Coast and Category IV includes barge mounted plants and Alaska North Slope plants. In addition to increases in capital, operating and maintenance costs, costs are also included for adding infrastructure. Natural gas is priced at \$0.50/10⁶ Btu. The economics are as follows:

Plant Capacity, BPSD of Fischer-Tropsch Liquid	14,500
Total Capital, 10 ⁶ U.S.\$ (Alaska North Slope or Barge-Mounted, December 1993)	1,764.1 (Note: includes 636.3 for infrastructure)
	<u>\$/B Fischer-Tropsch Product</u>
Natural Gas Feed, \$0.50/10 ⁶ Btu	4.30
Operating and Maintenance Costs	18.13
Capital Charges, 14.2% of Total Capital	54.10
Product Price:	
\$/Barrel	71.88
\$/Gallon	1.71

In Table 1 in a paper prepared by Bechtel [31], capital and operating costs are compared for a plant on the U.S. Gulf Coast and a plant on the Alaska North Slope. These cost relationships were used to obtain the following economics for a Fischer-Tropsch plant on the Alaska North Slope. There was no separate accounting for infrastructure. Natural gas is again priced at \$0.50/10⁶ Btu.

Plant Capacity, BPSD of Fischer-Tropsch Liquid	14,500
Total Capital, 10 ⁶ U.S.\$ (Alaska North Slope, December 1993)	1223.7
	<u>\$/B Fischer-Tropsch Product</u>
Natural Gas Feed, \$0.50/10 ⁶ Btu	4.30
Operating and Maintenance Costs	19.60
Capital Charges, 14.2% of Total Capital	37.19
Product Price:	
\$/Barrel	57.44
\$/Gallon	1.37

The economics show that when comparing costs for Fischer-Tropsch liquids on the U.S. Gulf Coast with costs on the North Slope or a barge mounted plant that Fischer-Tropsch liquids cost from about the same up to \$15/Barrel more. This is assuming the natural gas price on the U.S. Gulf Coast is about 5 times as great as on the North Slope or at a barge-mounted remote location. Thus in most cases the manufacture of Fischer-Tropsch liquids cannot be economically justified in a remote location where facility and operating costs would be significantly greater than on the U.S. Gulf Coast.

10.5. Effect of Using Ceramic Membranes for Oxygen Production

Considerable work has been done to develop a ceramic membrane that could be used for oxygen production as an integral part of a process for obtaining synthesis gas from natural gas. Argonne National Laboratory has made long tubes of Sr-Fe-Co-O membrane [136, 137]. These tubes have given methane conversion efficiencies as high as 99% for up to 350 hours. Oxygen is transported across the ceramic material in the form of oxygen ions. Only oxygen is transported so that nearly pure oxygen is available for the synthesis gas reaction on the other side of the membrane and there is no longer a need for air separation by means of a cryogenic oxygen plant.

Worcester Polytechnic Institute is studying the oxidative coupling of methane in an inorganic catalytic membrane [138, 139]. Their target is to convert methane into C₂ hydrocarbons using a membrane which allows a controlled amount of oxygen to pass through the membrane to the reaction site. Their work has concentrated on using porous VYCOR membrane reactors in laboratory studies and synthesizing La-Sr-Fe-Co oxide perovskites as oxygen transfer materials for dense membranes. They are doing both experimental work and modeling studies.

The concept of using a ceramic membrane to separate oxygen from air and then deliver the oxygen to take part in a reaction with methane for the production of hydrocarbons is very attractive and presents possibilities for reducing the costs of synthesis gas production and thus also the costs of methanol, Fischer-Tropsch products and ammonia. The process not only delivers nearly pure oxygen but can do so at a wide range of temperatures and pressures. In the natural gas to gasoline process, the syngas production step represents 56% of total capital cost; thus a cost reduction in this step could have a significant impact in overall process economics.

Unfortunately, no cost estimates are yet available for the capital and operating costs for the membrane system in a synthesis gas process. The new perovskite type ceramic membranes are quite different from other membranes used in gas separations and so it is not advisable to use cost data for older membrane systems for the newly developed ceramic membranes. High priority should be given to estimating costs of the new membrane system, so that economics could be obtained. This would permit a comparison with more conventional synthesis gas generation using cryogenic oxygen production. This will help to justify the continued research in this type of ceramic membrane development.

10.6. Effect of Changing Natural Gas Price

In some parts of the world, particularly in remote areas, natural gas prices are much lower than the typical U.S. gas prices. In such locations, it may be attractive to convert the gas to a liquid fuel or chemical and transport it to the U.S. or major consumer markets around the world. In order to determine how much of a decrease in the cost of the production of Fischer-Tropsch liquids from natural gas can be obtained by using low cost gas, economics have been obtained when initial natural gas prices are \$0.50 and \$0.00/10⁶ Btu and the gas price growth rate is equal to inflation (0% constant \$'s). It is assumed that capital and operating and maintenance costs are the same as in the U.S. Gulf Coast. There may be places in Asia and other remote areas where this situation is applicable.

The economics for a Fischer-Tropsch plant with slurry reactor with low cost natural gas are as follows:

Plant Capacity, BPSD	14,500	14,500	58,000
of Fischer-Tropsch Liquid			
Total Capital, 10 ⁶ U.S.\$	749.3	749.3	2,010.6
(USGC, December 1993)			
Natural Gas Feed Cost, \$/10 ⁶ Btu	0.50	0.00	0.00
	<u>\$/B Fischer-Tropsch Product</u>		
Natural Gas Feed	4.30	0.00	0.00
Operating and Maintenance Costs	11.88	11.88	7.47
Capital Charges,	22.07	21.94	14.86
14.2% of Total Capital			
Product Price:			
\$/Barrel	35.95	31.81	25.03
\$/Gallon	0.86	0.76	0.60

When comparing these economics to the economics of Section 10.1, it can be seen that a decrease of 1.93 to 2.43 \$/10⁶ Btu in natural gas price results in a decrease in the Fischer-Tropsch liquid price of 21 and 25 \$/B. When natural gas can be obtained at no cost, Fischer-Tropsch liquids can be produced which are only slightly more expensive than gasoline. However, the cost of transportation to market and upgrading still needs to be added to these costs. According to UOP, the cost of upgrading is \$5.78/Bbl of liquid product [61].

10.7. Summary

The sensitivity cases for the FT gas-to-liquids process are summarized in Exhibit 10-1. In Case 1, the base case, the cost of product is unattractive both because of a high feed cost and a high capital cost. The Case 5 versus Case 1 indicates that the feed cost can be reduced by moving the facility to a remote gas production location where the wellhead gas has significantly lower market value. Comparing Case 4 and Case 1 shows that the capital cost component can be reduced by using the slurry process and increasing the plant scale to 58 MB/D, a moderate sized plant in comparison to most refineries producing gasoline or diesel product. Case 7 demonstrates that if the plant of Case 4 could be built at no greater cost in a low gas cost remote area the product could be produced at 25 \$/Bbl - which brings it into the competitive price range with production of products from crude oil. Further achievement in reduction of the capital cost component could potentially be achieved with improvement in the syngas production step which represents over half of the capital cost of the production facility. The membrane syngas process may possess the potential of providing further capital cost reduction and merits further exploration.

It must be emphasized that making FT look economically attractive requires reduction of both feed and capital cost components. This means that a process that would utilize lower 48 gas is unlikely to be a candidate for an FT process. It can be seen as primarily a potential future competitor to LNG production for utilization of remote gas.

Exhibit 10-1. Fischer-Tropsch Process Economics Sensitivity Analysis

CASE	SYNGAS PROCESS	TYPE FT REACTOR	PLANT LOCATION	CAPACITY B/D	GAS PRICE: \$/MMBTU		UNIT COST OF PRODUCT: \$/BARREL							
					INITIAL	GWTH RATE	FEED	O&M	CAP CHRG	PLT GAL	FEED	O&M	CAP CHRG	PLT GAL
1	PARTIAL OXIDATION	ARGE FIXED-BED	GULF COAST	14,500	2.43	3.1	21.68	11.91	24.53	59.61				
2	PARTIAL OXIDATION	FT SLURRY	GULF COAST	14,500	2.43	3.1	20.92	11.88	22.56	56.84				
3	PARTIAL OXIDATION	FL SLURRY	GULF COAST	14,500	2.43	1.6	20.92	11.88	22.56	54.04				
4	PARTIAL OXIDATION	FOUR RX SLURRY	GULF COAST	58,000	2.43	3.1	20.92	7.47	15.05	45.67				
5	PARTIAL OXIDATION	FT SLURRY	ASIA	14,500	0.50	0	4.30	11.88	22.07	35.95				
6	PARTIAL OXIDATION	FT SLURRY	ASIA	14,500	0.00	0	0.00	11.88	21.94	31.81				
7	PARTIAL OXIDATION	FT SLURRY	ASIA	58,000	0.50	0	4.30	7.47	14.86	25.03				