Effect of Alternate Fuels on the Performance and Economics of Dispersed Fuel Cells

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Prepared by
Westinghouse Electric Corporation
Large, Pennsylvania

ELECTRIC POWER RESEARCH INSTITUTE

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Prepared by

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ABSTRACT

The purpose of this study was to assess the effects of alternative fuels on the performance and economics of conceptual Phosphoric Acid Fuel Cell (PAFC) power plants. The alternative fuels included natural gas, coal-derived medium Btu gas, methanol, ethanol, and naphtha. Properties of typical examples of these fuels were compiled along with the PAFC requirements to prepare preliminary specifications for fuel processing systems. Vendors were solicited to provide cost, performance, and operational data on fuel processing systems. These data were used to define ten different PAFC power plant configurations of nominally 7.5 MWe utilizing the five alternative fuels. Heat rates for the power plant configurations ranged from 7400 to 10,750 Btu/kWh. Direct capital costs were estimated in terms of prototype unit, 50th unit, and projected fully commercial units (which ranged from $420 to $593/kWe in 1980$) available in the 1990's. Parametric studies were conducted to determine the busbar cost of electricity as functions of: fuel costs, capacity factor, stage of commercial development, and production experience. Based on these studies, recommendations for primary and alternative fuel were developed: in the near term natural gas and methanol respectively were selected; in the longer term, the respective selections were methanol and coal gas.

EPRI PERSPECTIVE

PROJECT DESCRIPTION

RP1041 involves a group of contracts with the objective of expanding the range of fuels that can be used efficiently and economically in fuel cell systems. The project addresses issues including use of heavy liquid fuels (from coal and oil) in dispersed generators, use of coal in central station fuel cell systems, and integration of fuel processors with fuel cells. This final report describes the tradeoffs involved in using several light fuels (naphtha, natural gas, medium-Btu coal gas, or alcohols) in conceptual gas-cooled phosphoric acid power plants. The study assesses the effects that selected fuels, and the level to which the fuel processor is integrated into the system, have on power plant performance and cost.

PROJECT OBJECTIVE

The objective of this six-month study for RP1041-7 was to determine the preferred fuels and the preferred conceptual power plant configuration for dispersed generator applications. This study provides input to an effort sponsored by Southern California Edison Company to develop a detailed dispersed fuel cell generator specification. This specification will focus on the performance features and operational characteristics desired from the utility point of view for the conceptual power plant configurations. A third effort, as yet unfunded, would use these results to develop a detailed power plant design.

PROJECT RESULTS

Ten power plant configurations using the five alternative fuels were defined in this study. The cost and performance assessments indicate that the more closely integrated and therefore more fuel-efficient configurations were cost effective at current (and projected) fuel prices, even though their capital costs were higher. Methanol was determined to be the preferred fuel for the Westinghouse air-cooled phosphoric acid fuel cell from the standpoint of both capital cost and heat rate.

E. A. Gillis, Project Manager
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1.0 EXECUTIVE SUMMARY

The purpose of this study was to assess the effects of alternative fuels on the performance and economics of candidate Phosphoric Acid Fuel Cell (PAFC) power plants. The tasks performed during this study can be categorized as follows:

- The physical characteristics, impurities, and costs of alternative fuels were determined. Five fuels were considered which include natural gas, coal-derived gas, methanol, naphtha and ethanol.
- The technology status, cost and performance of fuel processing systems capable of handling these alternative fuels was determined. This effort included the preparation and issuance of a preliminary specification to potential fuel processor suppliers and evaluations of their responses.
- Fuel cell, fuel processor, and balance of plant performance and cost data were developed and used to estimate the overall performance and direct capital cost of PAFC power plants utilizing the alternative fuels.
- The economics of ten candidate PAFC power plant designs utilizing the alternative fuels were examined. These economic analyses included estimating the cost of electricity generated, estimates of break-even fuel costs, and an economic sensitivity study.

1.1 SUMMARY OF STUDY PARAMETERS

The study parameters include the assumptions, ground rules, and design conditions that guided or constrained the analyses. For the overall PAFC power plant the parameters were:

Fuel Cell Design Power Level = 7.5 MWe (DC)
Power Conditioner Efficiency = 96%
Ambient Air Temperature = 80°F
Heat Rejection Method = Dry Cooling Tower
Make up Water = None Required

The PAFC fuel cell design used herein features the DIGAS (Distributed Gas) cooling concept for power generation. Figure 1-1 illustrates the DIGAS stack which uses the same air for cooling and for feeding oxygen to the cathode side of the cell.

The fuel cell reject heat from the DIGAS stacks can be used to drive auxiliary rotating equipment within the fuel cell power plant. For a cogeneration application, the reject heat can also be used in an industrial facility.

The bipolar plate serves four primary functions: it separates the chemical reactants in adjacent cells, it conducts electrical current, it directs fuel to one cell and oxidant to the next, and it removes waste heat generated within the cell.

Other components of the fuel cell stack include the cooling plates which allow for additional heat removal from the cell. The fuel cell which produces dc power generates an appreciable amount of heat. The same air stream which supplies oxygen to the cathode can be used for the cooling channels. This DIGAS cooling concept represents a major step in simplified fuel cell design. The DIGAS approach significantly reduces demands for water treatment and manifolding connections, and permits coolant pressure to be independent of temperature.

Four stacks are combined to form a complete assembly (see Figure 1-2). The DIGAS concept permits the cathode gas to bathe the entire stack and be exhausted downward in the center section of the assembly. In this configuration, no separate cooling manifolds are necessary. Depending on the actual design, a fully assembled module would produce approximately 350-400 kW_e (dc). This approach to module design makes fuel cell systems extremely flexible. Adding capacity is accomplished simply by adding more assemblies to the total array. Similarly, individual assemblies can be replaced easily without disturbing the entire system.

Other features of the PAFC DIGAS fuel cell concept are:

- Low heat rate at rated conditions - potential for base load plant.
- Low heat rate at part power - potential for intermediate and peaking operation.
- No significant environmental impact.
  - No vapor plume
  - No significant contaminants, such as NO_x, SO_2 and CO
  - No outside water source required.
  - Adaptable to co-generation applications with minimal design changes.
- Fuel flexibility

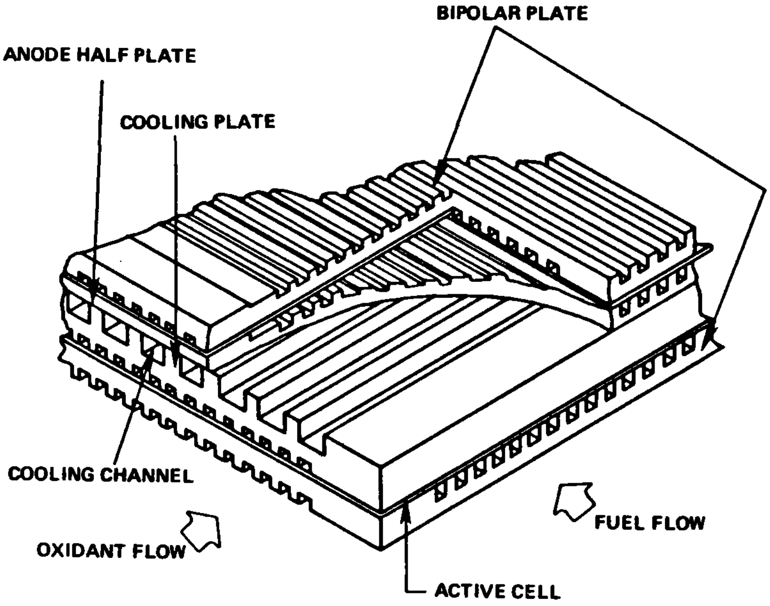


Figure 1-1. PAFC DIGAS Fuel Cell Stack

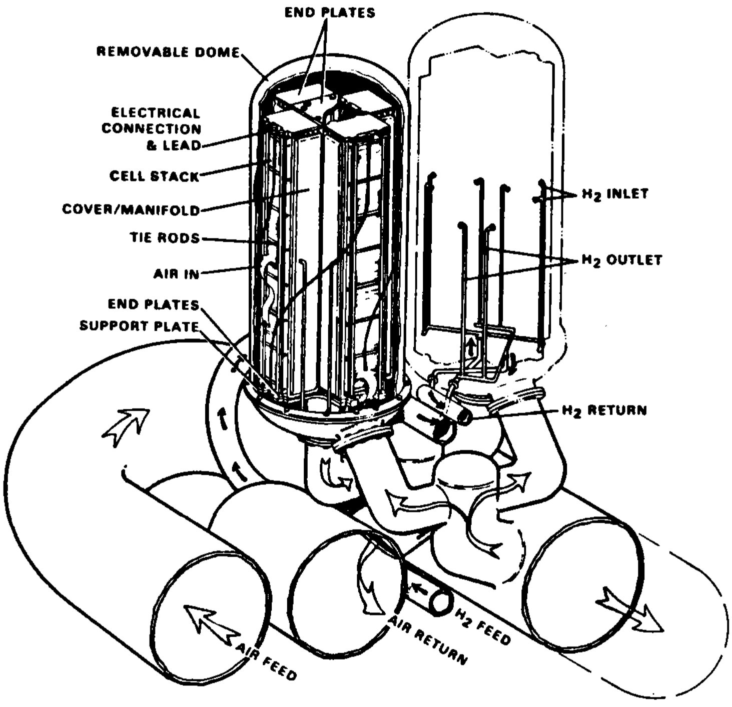


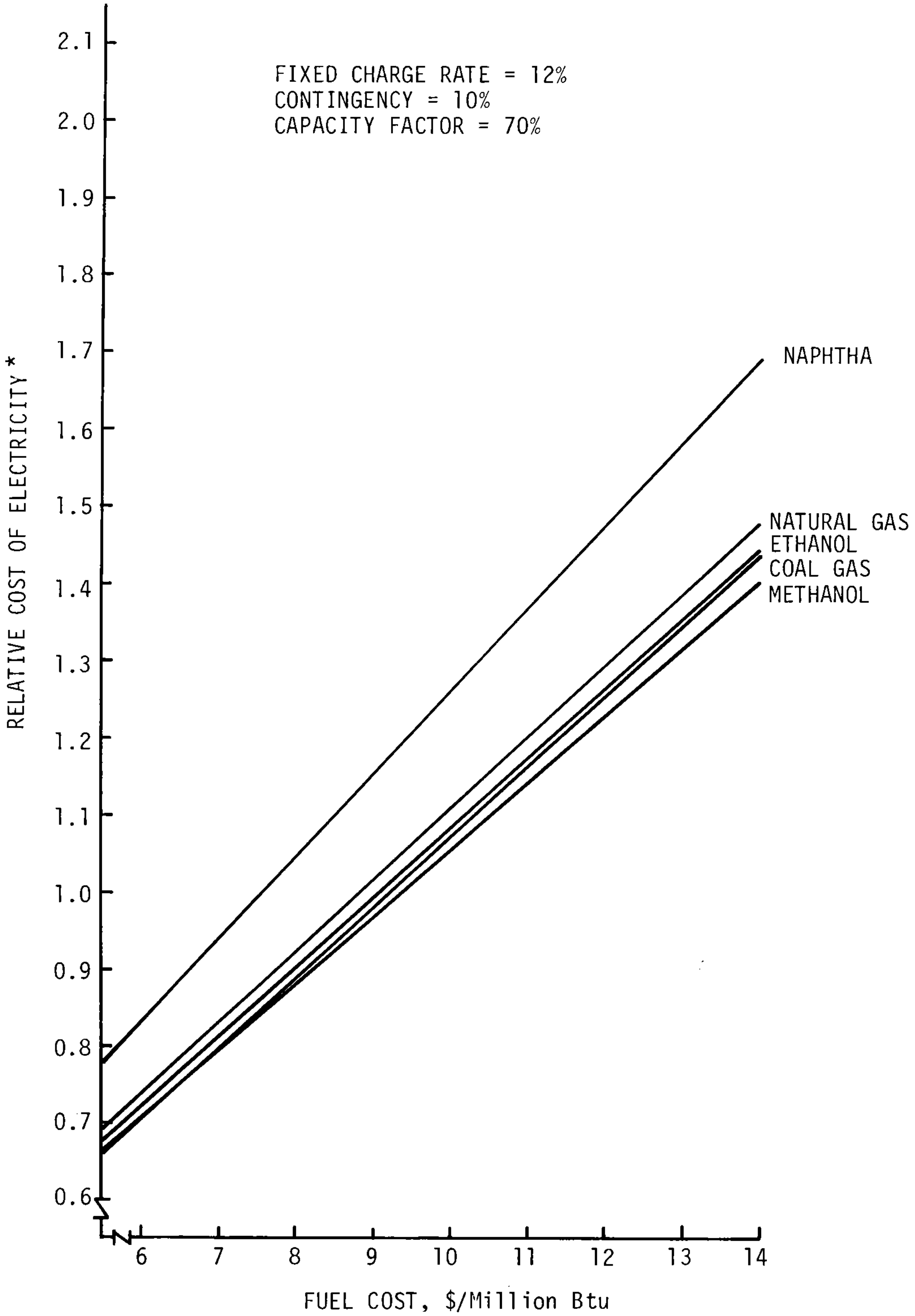
Figure 1-2. PAFC Fuel Cell Module With DIGAS Stacks

- Common manifolding for both coolant and reactant air.
- Coolant pressure is independent of temperature
- Reduced water treatment requirements
- Operation, maintenance, reliability, and sizing benefits due to modularity

Using the PAFC DIGAS concept, at least two types of overall plant designs were considered for each alternative fuel except ethanol where only one basic design was considered. These included: designs where the fuel processor was not integrated with the fuel cell; and integrated designs where the fuel cell cathode side reject heat was utilized to the extent practical to supply the thermal input requirements of the fuel processor. The nonintegrated designs have the advantage that the fuel processor can be replaced or modified as required for different fuels without major modifications to the rest of the PAFC power plant. In general, the non-integrated designs have higher heat rates and lower direct capital costs than the integrated designs.

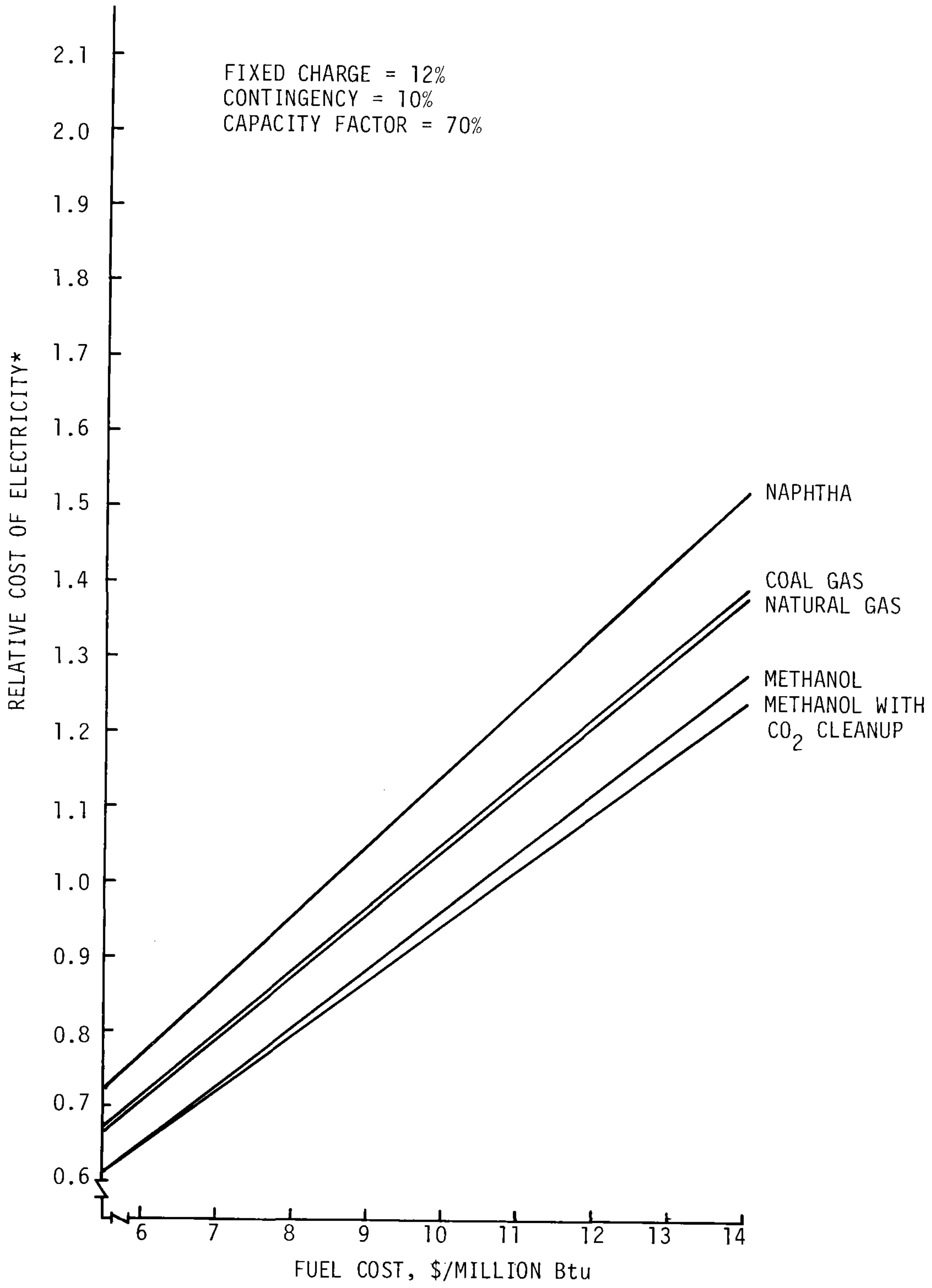
Typical properties for the five fuels were compiled from several existing sources. For the coal-derived gas properties typical of an O_2 blown medium BTU gasifer were used to conduct the performance and economic comparisons. The effects of the alternate types of fuels on the Fuel Processor System (FPS) were evaluated based on the solicitation of state-of-art FPS characteristics from credible suppliers, parameters for the PAFC power plant, and the types of alternative fuels. A specification was prepared and issued to potential suppliers which contained the requirements encompassing functional and process requirements and restrictions on diluents and impurities typical for fuel cell systems. This specification is included herein as Appendix A.

The cost of electricity (COE) was calculated for the PAFC power plant designs based on the assumption that the alternate fuels would be competitively priced over the 1990-2000 period. The effect of future fuel prices on the COE is illustrated graphically in Figures 1-3 and 1-4. The COE's for the ten configurations are also tabulated for estimated fuel price levels that are anticipated by the Fuel Cell Users Group to bracket competitively priced fuels in the 1990 to 2000 year time frame. These are:



* A Relative COE of 1.0 is Equal to 100 Mills/kWh

Figure 1-3. Relative Cost of Electricity as a Function of Fuel Cost for Nonintegrated Commercial Units



* A Relative COE of 1.0 is Equal to 100 Mills/kWh

Figure 1-4. Relative Cost of Electricity as a Function of Fuel Cost for Integrated Commercial Units

FUEL PRICE PROJECTIONS

	Constant 1980 $/ Million Btu	
	1990	2000
High Price Projection	8.16	10.97
Low Price Projection	7.04	7.04

It should be noted that questions as to the actual future availability of the various alternative fuels were beyond the scope of the present study.

Direct capital costs and total plant investment costs were estimated for prototype plants, 50th units of a production run, and projected commercial units. The 50th units are regarded as units that will be available for utility installations by 1990. Projected commercial units are regarded as a learned-out mature design that will be available in the late 1990's.

The sensitivity of COE estimates was also calculated for the following ranges:

Parameters	Study Range
Fixed Charge Rate	12%
Capacity Factor	30% and 70%
Fuel Cost Range	2-14 $/Million Btu
Experience Curve (50th Units)	90 ± 5%
" " (Projected Commercial 135th Unit)	85%

Experience curve effects have been observed in many types of production processes and they are a measure of the cost reductions that occur with increases in the cumulative size of a production run. These cost reductions are due to the combined effects of: improved labor efficiencies (i.e., learning curve effects); technical and manufacturing improvements; economies of scale; and volume purchasing of components and materials. A 90 percent experience curve means the cost per unit will decrease by 10 percent when the cumulative number of units produced is doubled. Similarly, an 85 percent experience curve will result in a 15 percent cost reduction when the cumulative number of units is doubled.

The O & M cost factors used in the COE estimates are based on the values given in 1979 EPRI Technical Assessment Guide but updated as follows to end-of-year 1980 dollars:

	1979 TAG (EOY 1978 $'s)	PRESENT STUDY (EOY 1980 $'s)
Fixed O&M Cost, $/kW-YR	3.3	3.9
Variable O&M Cost, Mills/kWH	3.2	3.8

1.2 SUMMARY OF STUDY RESULTS

Ten different PAFC plant configurations for five alternative fuels were defined during the study. The direct capital cost of the ten plant configurations, their heat rates, and their net electrical output are shown in Table 1-1. A detailed breakdown of the direct capital cost of each configuration can be found in Section 5.2.

Parametric Cost of Electricity calculations were also made for each of the ten plant configurations for the 50th units and the projected commercial units. The results were stated in terms of relative COE in order to facilitate comparisons between the candidate configurations and alternate fuels. Relative COE data for the 50th units, and for different capacity factors and fixed charge rates is included in Section 5.3 of the text.

The sensitivity of the relative COE to other parameters was also examined for one case. Given variation of the parameters individually by a fixed plus/minus percentage, the parameters in order of their decreasing impact on COE are:

Experience Curve
Capacity Factor
Heat Rate or Fuel Cost
Fixed Charge Rate or Direct Capital Cost
Interest During Construction
Contingency

Only three of the above parameters are controllable to a certain extent by the PAFC designer. They are the heat rate, direct capital cost, and the experience curve. But the designer is constrained by the trade-off between heat rate and direct capital cost for a given plant configuration. Since the fuel cost component of COE is much greater than the capital cost contribution, an incentive for improving the heat rate even at the expense of additional capital cost exists particularly at higher fuel prices. However, when other parameters including fuel price are held constant, COE is most sensitive to the experience curve factor for

Table 1-1

SUMMARY OF PAFC PLANT DIRECT CAPITAL COSTS WITH EFFECT OF EXPERIENCE CURVE, 1980 DOLLARS

FUEL	PLANT TYPE	HEAT RATE (Btu/Kwh)	NET OUTPUT MWe (AC)	PROTOTYPE DIRECT CAPITAL COST, $/KWe	50TH UNIT DIRECT CAPITAL COST, $/KWe			PROJECTED COMMERCIAL CAPITAL COST, $/KWe
					@ 95% EXPERIENCE	@ 90% EXPERIENCE	@85% EXPERIENCE	
Methanol	Non Integrated	8790	6.95	1504	1126	830	601	476
	Integrated	7820	6.92	1669	1250	921	667	528
	Highly Integrated With CO_2 Clean Up	7390	6.71	1940	1452	1070	775	614
Natural Gas	Non Integrated	9270	6.95	1604	1203	887	642	508
	Integrated	8440	6.95	1815	1361	1003	726	575
Coal Gas	Non Integrated	9110	8.10	1326	988	728	527	420
	Integrated	8520	6.95	1838	1370	1010	731	582
Naphtha	Non Integrated	10748	6.95	1666	1247	919	666	527
	Integrated	9448	6.95	1872	1400	1033	748	593
Ethanol	Non Integrated	9058	6.95	1519	1136	838	607	481

production units. The sensitivity of the COE to the experience curve reflects the fact that consistent achievement of even small experience curve benefits over a production run is more important that a one-time reduction in the capital cost of a specific unit.

Since the integrated methanol units have the lowest COE's, the incremental difference in mills per kwhr between the integrated commercial methanol unit and the other plant configurations was determined. The effect of the different plant types, capacity factor, and fuel cost on COE compared to the baseline methanol case is shown in Table 1-2. The COE increments, particularly with the commercial units at 70 percent capacity factor, illustrate the COE advantage that is due to the lower heat rates that can be achieved with methanol units. This advantage is maintained at the lower 30 percent capacity factor and the high fuel cost of 8.16 $/MBtu. At equal but higher fuel cost levels, the lower heat rate methanol units will show a greater incremental COE advantage. It should be noted that heat rate for each fuel and plant configuration was assumed constant and did not vary with capacity factor. A detailed yearly load profile as a function of capacity factor would be necessary to more precisely determine relative COE under part load conditions. Since, in general, PAFC plant efficiency tends to increase as loading is decreased, this COE data can only be used to compare PAFC plant configurations at different capacity factors. They cannot be used to compare PAFC plants to alternate generation units that decrease in efficiency when loading (or capacity factor) is reduced.

1.3 SELECTION OF REFERENCE FUELS

Table 1-3 summarizes the characteristics of PAFC power plant configuration for the five alternative fuels and for variations in the thermal integration of the FPS. Included in this table are operational factors, performance factors, capital costs and incremental COE. This data is presented based on the following assumptions:

- 7.5 MWe (DC) fuel cell power
- Projected Commercial Unit
- 1990 start date with low fuel price of 7.04 $/Million Btu
- 70% capacity factor

Of the three liquid fuels, commercial grade methanol is preferred over commercial grade ethanol and naphtha. Comparison of methanol versus ethanol indicates that

Table 1-2

INCREMENTAL EFFECT ON COE* OF UNIT TYPE, CAPACITY FACTOR, AND FUEL COSTS IN MILS/kWh

FUEL	PLANT TYPE	50TH UNIT @ 70% CAPACITY FACTOR	COMMERCIAL UNIT @ 70% CAPACITY FACTOR	COMMERCIAL UNIT @ 30% CAPACITY FACTOR	COMMERCIAL UNIT @ 70% CAPACITY FACTOR & HIGH FUEL COST
METHANOL	Non Integrated	+15	+ 6	+23	+16
	Integrated	+10	0	+20	+ 9
	Highly Integrated With CO_2 Clean Up	+11	- 1	+22	+ 7
NATURAL GAS	Non Integrated	+20	+10	+29	+21
	Integrated	+19	+ 6	+27	+15
COAL GAS	Non Integrated	+15	+ 7	+22	+17
	Integrated With CO_2 Clean Up	+18	+ 7	+28	+17
NAPHTHA	Non Integrated	+32	+21	+40	+33
	Integrated	+25	+14	+35	+24
ETHANOL	Non Integrated	+17	+ 8	+26	+18
Fuel Cost, $/MBtu		7.04	7.04	7.04	8.16

Fixed Charge Rate = 12%, Contingency = 10%, Experience Curve = 90% for 50th Units, 85% for Commercial Units

*COE Increment Above (+) or Below (-) the Integrated Methanol Unit at 70 Percent Capacity Factor and a Fuel Cost of 7.04 $/MBtu

Table 1-3

COMPARISON OF FUEL ALTERNATIVES FOR THE PAFC POWER PLANT
(At a 70% capacity factor and a fuel cost of 7.04 $/million Btu)

		FUEL PROCESSING SYSTEM								
		REFORMER TEMPERATURE (°F)	TECHNOLOGY STATUS	OPERATIONAL FACTORS	PROTOTYPE SYSTEM DIRECT CAPITAL COST $ MILLIONS	HEAT RATE (BTU/KWHR)	NET PLANT POWER OUTPUT MW_e (AC)	PROTOTYPE DIRECT CAPITAL COST ($/KW)	PROJECTED COMMERCIAL DIRECT CAPITAL COST $/KW	INCREMENTAL COST OF ELECTRICITY Mills/KWH
METHANOL	**Nonintegrated**	500-1000	Near mid-term/ developmental.	More rapid than natural gas.	2.2	8790	6.95	1500	480	+6
	Integrated	500-1000	Less complex than natural gas.	High availability and reliability.	3.3	7820	6.92	1670	530	0
	Integrated with CO_2 removal	500-1000	Same		3.3	7390	6.71	1940	610	-1
NATURAL GAS	**Nonintegrated**	1600-1800	Near term/ commercial.	Very slow response. Massive complex system.	2.9	9270	6.95	1600	510	+10
	Integrated	1600-1800	Very complex system.	Lowest availability.	4.4	8440	6.95	1820	580	+6
COAL GAS	**Nonintegrated**	None	Near term.	Fastest response. Least problems visualized.	2.2	9110	8.1	1330	420	+7
	Integrated with CO_2 removal	None	Least complex.	Highest availability.	2.2	8520	6.95	1840	580	+7
NAPTHA	**Nonintegrated**	1600-1800	Near term/ commercial.	Slow response.	3.2	10750	6.41	1670	530	+21
	Integrated	1000-1800	Most complex.		4.8	9450	6.94	1870	590	+14
ETHANOL	**Nonintegrated**	500-1000	Developmental	Similar to methanol.	2.2	9060	6.95	1520	480	+9

it is a toss up in terms of technology and operational factors; however methanol should have a lower heat rate and COE. Naphtha rates over methanol in technology status however its projected heat rate and COE are considerably higher.

Of the two gaseous fuels, natural gas and coal gas are a toss up in terms of performance and COE. The natural gas fuel process technology is more mature than coal gas. However coal gas should be superior operationally. Ultimately coal gas will be preferred over natural gas when large coal gasification plants reach commercial maturity. Therefore, for near term application, natural gas is preferred over coal gas and vice versa for far term applications.

Methanol is preferred over natural gas and coal gas due to slight advantages in performance and COE. Its technology status is less certain than natural gas and the production of methanol in large guantities is also tied to producing methanol from coal which is not a mature technology. Therefore in near term applications natural gas is preferred over methanol. Although operationally coal gas may be better than methanol, the performance and cost advantages particularly in an integrated plant, favor methanol. Based on the above considerations, the fuel recommendations are as follows:

	Near Term Applications	Future Commercial Applications
Primary Fuel	Natural Gas	Methanol
Alternate Fuel	Methanol	Coal Gas

1.4 CONCLUSIONS AND RECOMMENDATIONS

The major conclusions from this study are as follows:

- Based on considerations of heat rate, COE, and operational factors, methanol is selected as the preferred fuel for future commercial PAFC plant applications with coal gas as the second choice.
- For near term applications where current technology status is an important consideration, natural gas is the preferred fuel with methanol as the second choice.
- While there is extensive commercial experience with fuel processors which convert naphtha and natural gas into hydrogen, the complexity of an FPS for either naphtha or natural gas will make it relatively more difficult to design a commercial utility unit with the desired operation and maintenance characteristics.

- Experience curve benefits in a production run of commercial units will be important in minimizing capital costs and COE. Production and manufacturing planning must be an inherent and early part of the fuel processing system and complete PAFC power plant design process in order to achieve such benefits in actual practice.

- The high cost of the CO_2 clean up system tends to offset the COE and heat rate advantages of the extensively integrated methanol system when fuel costs are low. At the highest fuel costs the integrated methanol system with CO_2 clean up will yield the lowest COE.

- The fuel cost alone is by far the largest component in the COE of all the units considered. For this reason, it is significant that methanol units have the potential for the lowest heat rates. A heat rate in the range of 7000 Btu/kWh is a credible objective with methanol.

The major recommendations resulting from this study are as follows:

- The effects of PAFC plant operating pressures and temperatures on plant heat rate at full and part-load conditions and costs should be evaluated. The degree of technical risk associated with projected improvements in performance and costs should also be evaluated.

- Operability and performance features such as varying degrees of part-load, transient, and startup/shutdown capabilities should be examined in detail.

- Less expensive methods of CO_2 removal and resulting effects on heat rate and complexity, should be examined.

- A conceptual design and cost estimate of a fuel processing system that is integrated thermally with a PAFC plant should be developed with close attention paid to negative affects on availability due to more difficult maintenance.

- The cost, performance, operational, and COE impacts of adding multifuel capabilities to the fuel processing system needs to be considered in a further study.

- A study of multifuel capability should include the cost, performance, operational and COE aspects for each potential fuel that are described herein. The future availability of the potential fuels and utility-owner fuel preference should be inputs to such a future study. An evaluation of current projections for the future availability of methanol and coal gas should also be considered.

- The developmental costs and technical risks associated with bringing methanol and coal gas fuel processing systems from their present status to that required for commercial PAFC plants should be evaluated through appropriate vendors.

The importance of achieving experience curve benefits and successful commercialization of the PAFC plant also lead to the following recommendations:

- Standard sizes for the major systems of the PAFC power plant should be defined early to permit manufacturing and production planning for volume production to become an early and inherent part of the design process.
- Overall power plant specifications should be developed early in the design process to: facilitate production and manufacturing planning; to allow the selection of standard sizes; to permit the maximum use of common components; and to allow the definition of modular, factory assembled systems.
- The PAFC power plant specifications should be complete and comparable to the specifications used by utilities for the purchase of alternate types of generating units that will be competing with PAFC power plants. Such specifications should be carefully prepared in order to avoid imposing performance or other requirements on fuel cell power plants that either exceed those imposed on competing generating units or unnecessarily increase PAFC power plant cost or complexity.
- Conversely, the PAFC specifications should emphasize the unique benefits of fuel cells such as low part-load heat rates, low emissions, modularity, and siting flexibility.

2.0 INTRODUCTION

The development of a Phosphoric Acid Fuel Cell (PAFC) Power Plant can be significantly affected by the selection of a fuel or fuels for plant operation. For example, fuel selection will impact the design of the Fuel Processor System (FPS), the thermodynamic cycle arrangement to achieve an optimal cycle energy balance and performance, plant operating characteristics, plant costs and the Cost of Electricity (COE). Since a fuel cell operates with a hydrogen rich gas, the potential selection of a fuel(s) is varied. Consideration of projected fuel costs and availability can, however, restrict the options from a practical standpoint.

To facilitate the development of a commercially viable PAFC powerplant, Westinghouse has undertaken a study to assess the effects of alternate fuels on the performance and economics of candidate powerplants. Under EPRI Contract RP 1041-7 Westinghouse completed the first phase of this study which was to assess the performance and economics of PAFC powerplants designed to operate with each of the following five fuels:

methanol
natural gas
coal derived gas
ethanol
naphtha

An objective of this study is to select a candidate fuel and a preferred alternate fuel from the above candidates. These recommended fuels will be used to guide subsequent conceptual system design and trade off studies that will lead ultimately to a prototype power plant. Furthermore, it is anticipated that the results and conclusions of this study will be used to support the definition of system requirements being developed by the Fuel Cell Users Group. It is also anticipated that other fuels and multifuel capabilities will be evaluated in subsequent studies.

The tasks performed during this study can be categorized as follows:

- The physical characteristics, impurities, and costs of five alternative fuels were determined.

- The technology status, cost and performance of fuel processing systems capable of handling the alternative fuels was determined. This effort included the preparation and issuance of a preliminary specification to potential fuel processor suppliers and evaluations of their responses.

- Fuel cell, fuel processor, and balance of plant performance and cost data were developed and used to estimate the overall performance and direct capital cost of PAFC power plants utilizing the alternative fuels.

- The economics of ten conceptual PAFC power plant systems utilizing the alternative fuels were examined. These economic analyses included estimating the cost of electricity generated, estimates of break-even fuel costs, and an economic sensitivity study.

From this information preferred fuel candidates were selected based on overall plant performance, operating considerations and economics.

Section 3.0 describes the approach used in this study and includes the major assumptions, cases considered, costing bases and assumptions, and sensitivity parameters. Section 4.0 discusses the results of assessment of fuels and fuel processors based on existing data. Section 5.0 discusses the results of the system performance analyses and economic analyses. Conclusions from this study are highlighted in Section 6.0.

3.0 STUDY APPROACH

3.1 OVERALL OBJECTIVES

To accomplish the objectives of this study, five subtasks were defined. The first subtask consisted of reviewing previous studies, reports, and general literature to assimilate and define representative characteristics of five fuels. References 1-15 were data sources for this effort.

The second subtask consisted of defining representative design, performance, cost and technology characteristics for Fuel Processing Systems designed for each of the these fuels. A vendor survey was conducted to develop a data base. A specification (Appendix A) was issued to several vendors and responses were received from four of the vendors. Typical characteristics were defined from these responses.

The third subtask consisted of assessing the impact of the five fuels on system performance of at least two types of designs (integrated and non integrated) for each fuel. A nonintegrated design is defined as a plant where the FPS is a complete module that receives fuel and delivers hydrogen to the fuel cell and which provides all the energy to vaporize and preheat the fuel gas. The integrated designs utilize the cathode exhaust gas to supply thermal requirements for the fuel processor. For each design a cycle arrangement was defined and state points were calculated to determine the plant heat rate.

The fourth task consisted of estimating capital costs for each of the designs and calculating the cost of electricity (COE) for each design. The sensitivity of the COE as a function of several design, operating and costing variables were estimated.

The fifth task consisted of ranking the five candidate fuels. Key design performance and cost parameters were tabulated and compared.

3.2 STUDY ASSUMPTIONS

Major assumptions used in this study include:

- Assume that all candidate fuels will be available in sufficient quantities to meet commercial demand
- Assume that deregulated controls and free competition have resulted in the same prices for the five fuels
- Base the plant design on the 7.5 MW_e(DC) prototype plant concept using the PAFC DIGAS fuel cells proposed by Westinghouse
- Base economics on the 50th follow on unit with a 1990 start date and the projected commercial plant
- Base plant concept to minimize or eliminate need for a water supply.
- Base COE on constant 1980 $'s and real growth projections using economic and cost assumptions consistent with the 1979 EPRI Technical Assessment Guide
- Consider fuel prices as a variable over a range consistent with the Fuel Cell Users Group projections for natural gas and #2 fuel oil.

3.3 ANALYSIS VARIABLES

The following variables were considered in the plant performance and economic analyses

- Fuels
    - ethanol
    - coal derived gas
    - methanol
    - naphtha
    - natural gas
- Plant configurations
    - Nonintegrated fuel processing system
    - Integrated fuel processing system
- Fuel prices (for 1990 in 1980 $'s)
    - high price 8.20 $/M Btu
    - low price 7.00 $/M Btu
- Plant Capacity Factor
    - 30%
    - 70%
- Experience Curve for 50th unit
    - 90 % ± 5%
- Experience Curve for Projected Commercial Units
    - 85% and 135 th Unit
- Fixed Charge Rate
    - 12%

4.0 TECHNOLOGY ASSESSMENT OF FUELS AND FUEL PROCESSING

Objective assessments of the impacts of the primary source fuels and the fuel processor subsystems were made for the PAFC power plant. In support of this effort, assessment of the source fuels required characterization of the fuel composition which included identifying the various constituents and their concentration levels and the type and amount of fuel contaminants. In order to perform the assessment of the Fuel Processing System (FPS), potential suppliers were identified and contacted. For each source fuel the following information was requested:

- Price of the FPS
- O & M cost estimates
- FPS utility requirements
- Multiple fuel capability
- Technology status
- Estimated emissions
- Overall envelope size
- Lead time

4.1 FUEL CHARACTERISTIC ASSESSMENT

Assessments were made of the primary candidate fuels; methanol, natural gas, ethanol, naphtha, and coal gas. Characterization and identification of these fuels requires a qualitative understanding of the fuel composition which includes the various constituents and their concentrations and the type and amounts of fuel contaminants. A review of published industry data, existing studies, and reports and discussions with major manufacturers, consultants and vendors was made. Coal gasification systems potentially compatible with a PAFC power plant were also studied.

4.1.1 METHANOL

Reference 1 presented the results of a study on methanol supply and demand. Synthetic methanol is produced in 75 plants throughout the world. The total capacity is estimated at greater than 14 million metric tons per year. In the United States present methanol capacity is 3.6 million metric tons. This is approximately 85% of the total North America production. Table 4-1 gives a breakdown of the present capacity (1979) and location of methanol plants along with future predictions in the United States, Canada and Mexico. Table 4-2 shows the breakdown and applications for present and future methanol demand in the United States (Note 6.2 weight percent is for power generation in the year 1990). Table 4-3 gives a global summary of the world methanol balance in 1990 in thousands of metric tons.

Traditional methanol feedstocks include naphtha, natural gas and liquified petroleum gas (LPG). As the world methanol demand increases, the price of these feedstocks will increase. Countries that have low-cost natural-gas reserves will probably build large methanol projects. Large consuming regions (see Table 4-3) such as Japan and the U.S. will increasingly be net methanol importers if methanol is produced from these traditional feedstocks. It is expected that the methanol producing technology for other potential sources such as coal (and biomass) will be developed as the price for standard feedstocks increases and as the far more efficient low pressure production is developed.

A study has been performed (Reference 2) which assesses fuels for power generation, in which coal-derived methanol was considered as a fuel for fuel cells. For this study methanol was to be synthesized via the ICI process from synthesis gas produced from gasified coal. The gasification system considered was demonstrated commercially with a variety of coal feedstock including caking type coals. The gasification operation was also assumed to be flexible enough that it could be done at high pressures eliminating the compression of the synthetic gases prior to the ICI methanol synthesis (highly developed commercial proven process). Typical representative compositions by volume percent reported for this gasification process were: (based on Illinois #6 and New Mexico coal)

CH_3 OH	97.7
Higher Alcohols	1.9
Water	0.4

Similar compositions were reported by Energy Research Corporation (ERC) and the Fuel Cell User's Group (FCUG) Fuel Subcommittee (Reference 3). The methanol can be generated from either natural gas, naphtha or coal. For fuel grade methanol

TABLE 4-1
NORTH AMERICAN METHANOL CAPACITY*
(THOUSAND METRIC TONS PER YEAR, 1979)

U.S.	Plant Location	Feedstock	Present Capacity	Future	Startup
Air products	Pensacola, FL	Natural gas	150	150	
Allemania	Plaquemine, LA	Natural gas	300	300	
Borden	Geismar, LA	Natural gas/offgas	480	570	1980
Celanese	Bishop, TX	Natural gas	435	435	
	Clear Lake, TX	Natural gas	690	690	
Du Pont	Beaumont, TX	Natural gas	660	825	1982
	Dee Park, TX	Fuel oil	-	600	1980
Georgia Pacific	Plaquemine, LA	Natural gas	360	360	
Getty	Delaware City, DE	Raffinate	-	300	
Monsanto	Texas City, TX	Natural gas	300	300	
Tenneco	Houston, TX	Natural gas/offgas	240	390	
Arco	Gulf Coast	NG/LPG	-	600	1983
			3,615	5,520	
Canada					
Alberta Gas Chemicals	-	Natural gas	360	720	1982
Celanese Canada	-	Naptha	51	51	
		Natural gas	-	705	1982
Ocelot	-	Natural gas	-	360	1982
Westcoast Transmission	-	Natural gas	-	360	1983
			411	2,196	
Mexico					
Pemex	-	Natural gas	171	825	1985
Total North America	-		4,197	8,541	

*See Reference 1
50:50 joint venture of Ashland and IMC.
Source: Chem Systems Inc. and CE estimates

TABLE 4-2

U.S. METHANOL DEMAND*
(WEIGHT PERCENT)

Application	Year 1979	Year 1990
Formaldehyde	42.5%	28.0%
DMT	4.4	2.0
Methyl Malides	8.3	5.0
Methylamines	5.0	2.8
Methyl Methacrylate	4.5	3.8
Solvents	9.4	7.1
Acetic Acid	6.7	10.8
Gasoline Blending	8.0	14.1
Miscellaneous	11.2	10.8
MTBE**	-	9.4
Power Generation	-	6.2

*See Reference 1

**MTBE - Methyl tert - butyl ether

TABLE 4-3

WORLD METHANOL BALANCE IN 1990*
(THOUSAND OF METRIC TONS)

Country	Methanol Demand	Methanol Potential Supply
Japan	2,120	950
Other Asian	1,087	1,130
Australia/New Zealand	125	360
Canada	312	1,150
U.S.	8,070	5,340
Latin America	578	1,641
Western Europe	5,960	4,025
Eastern Europe	5,000	5,620
Africa-Middle East	100	2,575
Total	23,352	23,141

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*See Reference 1

any excess water and higher grade alcohols can be removed by distillation. Table 4-4 gives the methanol purity that the FCUG presented (Reference 3) for raw grade and fuel grade product.

Physical properties of methanol are reported in Tables 4-5 and 4-6. These constants and composition relationships, i.e., mixtures with water, are given in Reference 4. The physical properties of methanol - water mixtures are of importance because methanol is so frequently used as an aqueous solution. The methanol and water to the FPS may be premixed prior to reforming.

Based on these data sources the following composition was assumed to be representative for methanol (produced from standard feedstocks) and was used in the present study.

CH_3 OH	96% (by volume)
Higher alcohol	1.5%
Water	2.5%

4.1.2 NATURAL GAS

Typical natural gas compositions for various U.S. cities, major transmission lines and various gas fields were collected in a survey by the American Gas Association and are given here in Tables 4-7 through 4-10. Note that the natural gas compositions generally contain a high percentage of methane (CH_4) with varying amounts of other hydrocarbons such as ethane (C_2H_6) and inerts (carbon dioxide, nitrogen, and helium). Most natural gas delivered by utilities will have under 10 percent of inerts. The heating value will generally range between 1000 and 1100 Btu/SCF (standard conditions of 62°F and 30 in Hg).

Table 4-8 classifies the various natural gases into groups, namely:

• High Inert Type	6.3 - 16.2% N_2; 71.9 - 83.2% CH_4
• High Methane Type	0.1 - 2.4% N_2; 87.6 - 95.7% CH_4
• High Btu Type	1.2 - 7.5% N_2; 85.0 - 90.1% CH_4

Average values for these compositions (average for 48 cities in the U.S.) were reported by Energy Research Corporation (ERC) to the FCUG Fuel's Subcommittee

TABLE 4-4

METHANOL PURITY

Process	Compounds	Mole %	
		Raw Product	Fuel Grade Product
*ICI - Low Pressure	CH_3OH	65	97
ICI - Low Pressure	H_2O	30	2
ICI - Low Pressure	Other	5	1
ICI - Low Pressure	Other - Primarily $C_4H_{10}OH$ and (CH_3)	2	0
High Pressure	CH_3OH	same	95
Vulcan-Cincinnati	H_2O		3
Wentworth	Other		2

*ICI - Imperial Chemical Industries

TABLE 4-5
PHYSICAL PROPERTIES OF METHANOL*

Constant	Value
F.p., °C.	-97.0 to -97.8
B.p., °C.	64.5-64.7
Density, gram/ml., at:	
15°C	0.79609
30°C	0.78208
D at 16°C.	1.3306
Viscosity, centipoise, at:	
	0.6405
15°C	0.623
20°C	0.5945
	0.5525
25°C	0.544
	0.5142
30°C	0.510
Surface tension, dynes/ sq.cm., at:	
15°C	22.99
20°C	22.55
30°C	21.69
Crit. temp., °C	240.0
Crit. pres., atm.	78.5
Crit. soln. temp., °C., in:	
CS_2	35.0 (app.)
C_6H_{14}	34.50
C_6H_{12}	46.05

Constant	Value
Sp. heat of:	
Liquid, cal./gram, at:	
0-5°C.	0.570-0.580
20-25°C	0.595-0.605
40-45°C	0.610-0.620
Ideal vapor, cal./gram-mole, at:	
0°C.	10.27
25°C.	10.76
100°C.	12.20
200°C.	14.40
Heat of vaporization at 64.7°C., cal./ mole	8430
Heat of combustion of liquid at 25°C., cal./mole.	-173,650
Heat of formation at 25°C., cal./mole, of:	
Liquid	-57,036
Vapor.	-48,100
Flash point (open cup)., °C	15.6
Flammability limits in air, vol.%	6-36
Ignition temp., °C. in:	
Air.	473
Oxygen	461
Solubility in:	
Water.	
Alcohol.	
Ether.	

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*Reference 4

TABLE 4-6

DENSITY, AND COMPOSITION RELATIONSHIP OF MIXTURES OF METHANOL AND WATER AT 25°C*

METHANOL, WT. %	METHANOL, VOL. %	Density g/cc	METHANOL, WT. %	METHANOL, VOL. %	Density g/cc
0.	0.00.	0.99708	60.	67.96.	0.8910
10.	12.46.	0.9804	70.	77.19.	0.8675
20.	24.53.	0.9649	80.	85.66.	0.8424
30.	36.20.	0.9492	90.	93.33.	0.8158
40.	47.37.	0.9316	100.	100.00.	0.7867
50.	57.98.	0.9122			

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*Reference 4

Table 4-7

NATURAL GAS DISTRIBUTED IN VARIOUS CITIES IN THE UNITED STATES*
(Surveyed by A.G.A.* in the Fall of 1962)

		COMPONENTS OF GAS, PERCENT BY VOLUME										
NO.	CITY	METHANE	ETHANE	PROPANE	BUTANES	PENTANES	HEXANES PLUS	CO_2	N_2	MISCEL.	HEAT VALUE† Btu/cu ft	Sp gr
1	Abilene, Tex.	73.52	13.23	4.35	0.56	0.06	0.11	0.16	8.01	...	1121	0.710
2	Akron, Ohio	93.30	3.49	0.69	0.18	0.04	0.00	0.50	1.80	...	1037	0.600
3	Albuquerque, N.M.	86.10	9.49	2.34	0.44	0.08	0.03	1.02	0.50	...	1120	0.645
4	Atlanta, Ga.	93.42	2.80	0.65	0.33	0.12	0.10	1.38	1.20	...	1051	0.604
5	Baltimore, Md.	94.40	3.40	0.60	0.60	0.00	0.00	0.60	0.50	...	1051	0.590
6	Birmingham, Ala.	93.14	2.50	0.67	0.32	0.12	0.05	1.06	2.14	...	1074	0.599
7	Boston, Mass.	93.51	3.82	0.93	0.28	0.07	0.06	0.94	0.39	...	1057	0.604
8	Brooklyn, N.Y.	94.52	3.29	0.73	0.26	0.10	0.09	0.70	0.31	...	1049	0.595
9	Butte, Mont.	87.38	3.02	1.09	0.11	0.06	0.00	1.93	6.35	...	1000	0.610
10	Canton, Ohio	93.30	3.49	0.69	0.18	0.04	0.00	0.50	1.80	...	1037	0.600
11	Cheyenne, Wyo.	91.00	4.73	1.20	0.30	0.06	0.04	1.85	0.81	...	1060	0.610
12	Cincinnati, Ohio	94.25	3.98	0.57	0.16	0.03	0.03	0.68	0.30	...	1031	0.591
13	Cleveland, Ohio	93.30	3.49	0.69	0.18	0.04	0.00	0.50	1.80	...	1037	0.600
14	Columbus, Ohio	93.54	3.58	0.66	0.22	0.06	0.03	0.85	1.11	...	1028	0.597
15	Dallas, Tex.	86.30	7.25	2.78	0.48	0.07	0.02	0.63	2.47	...	1098	0.641
16	Denver, Colo.	81.11	6.01	2.10	0.57	0.17	0.08	0.42	9.19	...	1011	0.659
17	Des Moines, Iowa	80.38	6.39	2.46	0.61	0.08	0.03	0.20	8.53	0.32 He	1012	0.669
18	Detroit, Mich.	89.92	4.21	1.34	0.34	0.09	0.01	0.59	3.30	0.20 He	1015	0.616
19	El Paso, Tex.	86.92	7.95	2.16	0.16	0.00	0.00	0.04	2.72	0.05 He	1082	0.630
20	Ft. Worth, Tex.	85.27	8.43	2.98	0.62	0.09	0.04	0.27	2.30	...	1115	0.649
21	Houston, Tex.	92.50	4.80	2.00	0.30	....	....	0.27	0.18	...	1081	0.623
22	Kansas City, Mo.	72.79	6.42	2.91	0.50	0.06	Trace	0.22	17.10	...	1945	0.695
23	Little Rock, Ark.	94.00	3.00	0.50	0.20	0.20	....	1.00	1.10	...	1085	0.590
24	Los Angeles, Calif.	86.50	8.00	1.90	0.30	0.10	0.10	0.50	2.60	...	1084	0.633
△25	Louisville, Ky.	94.05	3.41	0.40	0.13	0.05	0.09	1.20	0.67	...	1054	0.595
26	Memphis, Tenn.	92.50	4.37	0.62	0.18	0.07	0.10	1.60	0.56	...	1044	0.603
27	Milwaukee, Wis.	89.01	5.19	1.89	0.66	0.44	0.02	0.00	2.73	0.06 He	1061	0.627
28	New Orleans, La.	93.75	3.16	1.36	0.65	0.66	0.00	0.42	0.00	...	1072	0.612
29	New York City	94.52	3.29	0.73	0.26	0.10	0.09	0.70	0.31	...	1049	0.595
30	Oklahoma City, Okla.	89.57	6.31	1.36	0.36	0.00	0.00	13	2.06	0.21 O_2	1080	0.615
31	Omaha, Neb.	80.46	6.30	2.59	0.68	0.09	0.05	0.17	9.32	0.34 He	1070	0.669
32	Parkersburg, W. Va.	94.50	3.39	0.68	0.12	0.07	0.03	0.67	0.41	0.01 O_2	1049	0.592
33	Phoenix, Ariz.	87.37	8.11	2.26	0.13	0.00	0.00	0.61	1.37	...	1071	0.633
34	Pittsburgh, Pa.	94.03	3.58	0.79	0.28	0.07	0.04	0.80	0.40	0.01 O_2	1051	0.595
35	Providence, R.I.	93.05	4.01	1.02	0.34	0.08	0.09	1.00	0.42	...	1057	0.601
36	Provo, Utah	91.40	3.95	0.84	0.39	0.03	0.01	0.52	2.85	...	1032	0.605
37	Pueblo, Colo.	73.85	5.71	3.20	1.34	0.14	0.06	0.13	15.26	...	980	0.706
38	Rapid City, S.D.	90.60	7.20	0.82	0.19	0.03	0.03	0.18	0.93	0.02 He	1077	0.607
39	St. Louis, Mo.	93.32	4.17	0.69	0.19	0.05	....	0.98	0.61	...	....	.....
40	Salt Lake City, Utah	91.17	5.29	1.69	0.55	0.16	0.03	0.29	0.82	...	1082	0.614
41	San Diego, Calif.	86.85	8.37	1.86	0.15	0.00	0.00	0.41	2.32	0.04 He	1079	0.643
42	San Francisco, Calif.	88.69	7.01	1.93	0.28	0.03	0.00	0.62	1.43	0.01 He	1086	0.624
43	Toledo, Ohio	93.54	3.58	0.66	0.22	0.06	0.03	0.85	1.11	...	1028	0.597
44	Tulsa, Okla.	86.29	8.36	1.45	0.18	0.14	0.01	0.23	2.95	0.39 O_2	1086	0.630
45	Waco, Tex.	93.48	2.57	0.89	0.43	0.17	0.11	1.69	0.66	...	1042	0.607
46	Washington, D.C.	95.15	2.84	0.63	0.24	0.05	0.05	0.62	0.42	...	1042	0.585
47	Wichita, Kan.	79.62	6.40	1.42	1.12	0.48	0.14	0.10	10.62	0.10 O_2	1051	0.660
48	Youngstown, Ohio	93.30	3.49	0.69	0.18	0.04	0.00	0.50	1.80	...	1037	0.600

• Average analyses obtained from the operating utility company(s) supplying the city; the gas supply may vary considerably from these data — especially where more than one pipeline supplies the city. Also, as new supplies may be received from other sources, the analyses may change. Peak shaving (if used) is not accounted for in these data.

† Gross or higher heating value at 30 in. Hg. 60 F, dry. To convert to a saturated basis deduct 1.73 percent; i.e., 17.3 from 1000, 19 from 1100.

△ 1954 data.

* AGA - American Gas Association

Table 4-8

GROUP CLASSIFICATION OF NATURAL GASES

	GROUP	NITROGEN, %	SPECIFIC GRAVITY	METHANE, %	Btu/cu ft, DRY
I	High Inert Type	6.3 - 16.20	0.0660 - 0.708	71.9 - 83.2	958 - 1051
II	High Methane Type	8.1 - 2.39	0.59 - 0.614	87.6 - 95.7	1008 - 1071
III	High Btu Type	1.2 - 7.5	0.62 - 0.719	85 - 90.1	1071 - 1124

Table 4-9

NATURAL GAS DELIVERED BY MAJOR TRANSMISSION LINES*

NO.	TRANSMISSION LINE AND MAJOR SOURCE OF SUPPLY	COMPONENTS OF GAS — PERCENT BY VOLUME									GROSS HEAT VALUE Btu per cu ft, DRY†	Sp gr	WATER VAPOR lb per MMCF
		METHANE	ETHANE	PROPANE	BUTANE	PENTANE	HEXANE PLUS	CO_1	O_1	N_1			
1	Cities Service Gas Co. from Texas Panhandle	73.48	6.86	4.26	2.13	0.63	0.40	0.10	0.32	11.90	1077	0.694	9.80
2	Cities Service Gas Co. from Oklahoma Hugoton	75.28	6.39	3.76	1.45	0.29	8.29	....	....	12.54	1043	0.706	3.50
3	Cities Service Gas Co. from Kansas Hugoton	77.02	3.69	2.58	2.04	0.49	0.13	0.10	0.10	13.65	1005	0.698	3.00
4	Colorado Interstate Gas Co. from Kansas Hugoton	72.40	6.12	3.21	1.20	0.15	....	....	....	16.92	903	0.703	5.80
5	Colorado Interstate Gas Co. from Texas Panhandle	78.76	5.67	2.88	1.06	0.10	....	....	....	11.53	1007	0.683	8.40
6	El Paso Natural Gas Co. from Permian Basin	81.21	9.42	3.45	0.63	0.04	....	....	....	5.25	1097	0.665	10.00
7	Kansas Nebraska Natural Gas Co. from Hugoton	71.25	5.69	3.37	0.98	0.14	0.20	0.10	....	18.27	955	0.7161	....
8	Lone Star Gas Co. from Texas and Oklahoma	85.00	7.10	2.40	0.50	0.40	....	0.60	....	4.00	1059	0.650	15.00
9	Michigan Wisconsin Pipeline Co. from Panhandle	73.10	6,20	4.00	0.90	....	....	....	....	15.80	973	0.704	3.50
10	Mississippi River Fuel Corp. from Monroe	93.17	4.23	0.75	0.22	0.07	0.11	1.22	0.01	0.32	1049	....	3.0
11	Montana Dakota Utilities Co. from Montana and Dakotas	93.00	6.00	0.20	....	....	....	0.30	....	0.50	1010	0.600	....
12	Naturas Gas Pipeline Co. of America from Panhandle	79.00	6.00	3.70	1.00	0.10	....	....	....	10.20	1039	0.685	8.00
13	Northern Natural Gas Co. from Panhandle	75.78	4.97	3.24	2.10	....	....	0.20	0.14	13.55	1011	0.685	11.08
14	Pacific Gas & Electric Co 34° from El Paso Nat. Gas Co. Sys.	81.90	9.30	3.30	0.50	....	....	....	....	5.00	1100	0.660	Win 11 Sum 15
15	Panhandle Eastern Trans. Co. from Panhandle	72.40	15.70	0.10	....	....	....	0.20	0.10	11.30	1020	0.680	....
16	Southern California Gas Co. from El Paso Nat. Gas Co. Sys.	81.40	8.70	3.60	0.60	0.10	....	....	....	5.60	1092	0.670	10.00
17	Southern Natural Gas Co. from Monroe and Others	94.95	1.30	0.33	0.11	0.09	0.14	0.70	....	2.39	1008	0.590	5.00
18	Tennessee Gas Transmission Co. from Louisiana & Texas	94.61	3.30	0.99	0.38	0.03	0.18	0.35	....	0.16	1065	0.5967	7.28
19	Texas Eastern Transmission Corp. from Louisiana & Texas	92.58	4.27	0.97	0.21	0.04	0.06	0.90	....	0.95	1051	0.606	3.80
20	Texas Gas Transmission Corp from Louisiana & Texas	92.80	4.20	0.90	0.20	....	0.10	1.00	....	0.80	1049	0.600	....
21	Transcontinental Gas Pipelin Corp. from Louisiana & Texas	93.45	3.59	1.27	0.61	0.26	0.22	0.60	....	....	1085	0.6102	3.50
22	United Gas Pipeline Co. from Refugio to Houston	92.61	3.87	1.15	0.39	0.12	0.07	0.66	....	1.13	1056	0.6049	....
23	United Gas Pipeline Co. from Carthage Longview	91.46	4.18	0.94	0.24	0.02	0.09	0.89	....	2.18	1037	0.6074	....
24	United Gas Pipeline Co. from Waskom to Dallas	89.55	4.97	0.93	0.33	0.07	0.24	0.91	....	2.95	1046	0.6219	....
25	United Gas Pipeline Co. from Slico to Shreveport	92.77	3.35	0.83	0.45	0.18	0.06	1.03	....	1.33	1044	0.6054	....
26	United Gas Pipeline Co. from Lirette to Mobile	93.32	3.00	1.06	0.57	0.20	0.20	0.49	....	1.16	1062	0.6063	....
27	United Gas Pipeline Co. from Carthage to Sterlington	92.82	3.55	0.69	0.09	0.03	0.10	1.01	....	1.71	1029	0.5995	....
28	United Gas Pipeline Co. from Agua Dulce to Austin	84.73	7.90	2.49	0.76	0.32	0.25	2.54	....	1.01	1112	0.6695	....

*These are average analyses obtained from the operating companies. As gas from new fields may be acquired or changes made in the processing plants, the analyses may change from time to time.

†30 in. Hg, 60 F. To convert to a saturated basis deduct 1.73 percent: i.e., 17.3 from 1000, or 10 from 1100.

Table 4-10

NATURAL GAS FROM VARIOUS GAS FIELDS
(As of November 1961)

		COMPONENTS OF GAS — PERCENT BY VOLUME										
NO.	LOCATION AND FIELD	METHANE	ETHANE	PROPANE	BUTANE	PENTANE	HEXANE PLUS	CO_1	O_1	N_1	GROSS HEAT VALUE Btu per cu ft, DRY	Sp gr
1	California, Kettleman North Dome (b)	87.20	5.20	3.50	2.00	1.70	....	0.40	....	....	1212	0.690
2	California, Kettleman North Dome (a)	93.00	4.60	1.50	0.20	....	....	0.40	....	....	1030	0.602
3	California, Rio Vista	94.20	2.95	0.80	0.15	0.10	....	0.30	....	1.50	1038	0.590
4	California, Ventura(b)	83.60	5.40	6.10	3.70	1.40	....	0.30	....	....	1260	0.706
5	California, Ventura(a)	92.70	4.70	2.20	0.10	....	....	0.30	....	....	1083	0.604
6	Minn. - Indiana	97.00	....	....	....	....	....	1.00	0.40	1.60	983	.....
7	Kansas, Cunningham	62.30	21.00	....	....	....	....	0.20	0.30	16.00	1011	.....
8	Kansas, Hugoton	77.00	3.90	2.60	2.00	0.50	8.10	0.10	0.10	13.60	1005	0.558
9	Kansas,	71.80	13.90	....	....	....	....	0.30	0.20	13.80	976	.....
10	Kentucky	81.40	13.30	....	....	....	....	0.80	0.40	2.10	1063	.....
11	Louisiana	91.28	1.52	8.70	8.41	0.19	0.15	0.30	....	5.45	997	0,6075

(Reference 3). Table 4-11 gives these averages along with the characteristic conditions of pipeline gas.

The FCUG Fuel's Subcommittee published descriptions of selected fuels for intermediate and peaking duty PAFC plants (Reference 5). Natural gas was identified in this report as typically being 90 - 95% methane with some heavier hydrocarbon gases, nitrogen and/or carbon dioxide. Light hydrocarbons and sulfur compounds present in the gas from the well are removed at the source before transmission through the pipeline. The heating value ranges from 950 - 1050 Btu/Scf. Small amounts of sulfur compounds such as Mercaptans (C_2H_5SH), Methyl Sulfides ($(CH_3)_2S$) and hydrothiophenes (CH-CH-CH-CH-S) are added to the natural gas as an odorant in the local distribution system.

Figure 4-1 shows the flow schematic for natural gas as a fuel for a PAFC power plant application. Listed are the current spec requirements for each major component within the FPS and the fuel cell subsystems. Note that for the most part the FPS is limiting on impurity concentration, i.e., S < 1 ppm in the reforming. These limits and specs apply to the component designs and catalysts that are available today. New catalysts may allow for higher impurity levels.

The average composition for all groups gives a natural gas fuel consisting of:

• CH_4	89.4% (by volume)
• Higher hydrocarbons	7.0%
• Inerts	3.6%

Natural gas composition can vary significantly as a function of site location. However, the average composition was used for the present study and the potential use of peak shaving gas was neglected.

4.1.3 COAL GAS

The information gathered on the various coal gasification processes was for both low and medium Btu gas and specifically includes the detailed surveys conducted by ORNL and ERDA (References 6 and 7). Of over a hundred gasification processes studied, only the most promising were selected taking into account such major categories as degree of development or commercialization, technical complexity, availability, and coal restrictions.

TABLE 4-11
NATURAL GAS FUEL DEFINITION*

- Constituents, Volume Percent; Average

CH_4,	72.4 to 95.15;	89.4
C_2H_6,	2.5 to 11.32;	5.1
C_3,	0.71 to 5.07;	1.9
N_2,	0.3 to 17.1;	2.9
CO_2,	0 to 1.98;	0.7
He,	0 to 0.34;	--

- DEW Point Below 40°F
- Olefins to 0.2 Volume %
- O_2 to 0.3 Volume %
- For Home Use:
  - Pressure; 2" to about 20" H_2O
  - Odorized by Law
    Mercaptans, Methyl Sulfides, and Hydrothiophenes

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*Reported by Energy Research Corporation (ERC) to FCUG Fuel's Subcommittee in Denver (Reference 3).

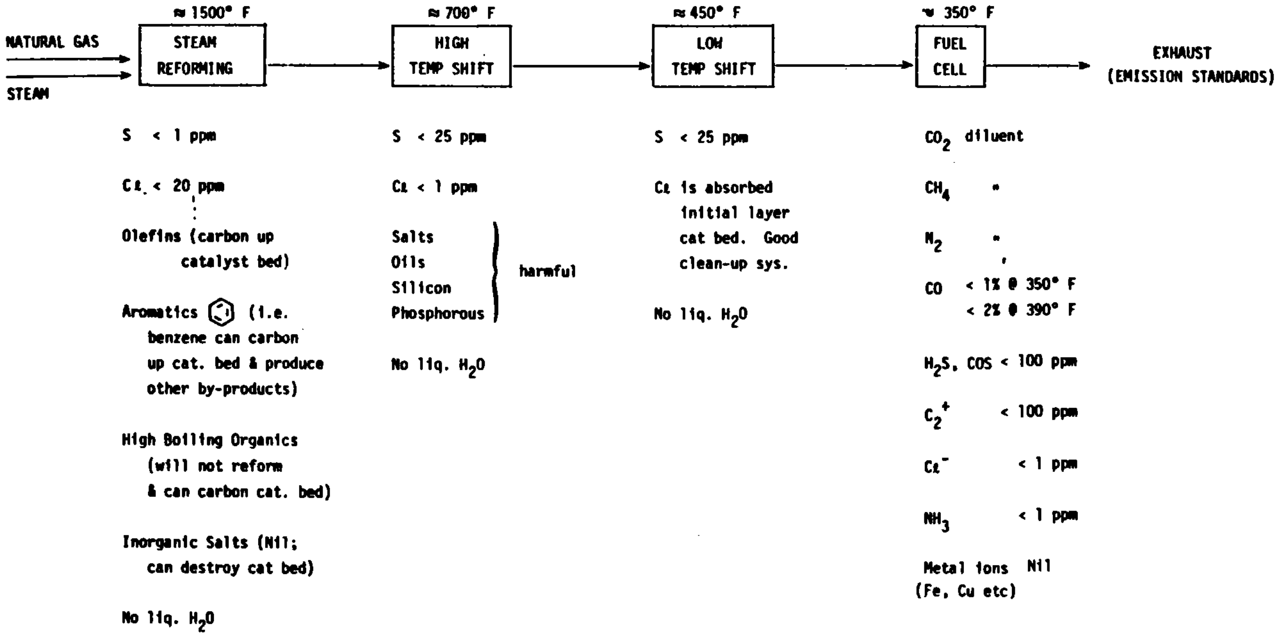


Figure 4-1. Natural Gas FPS/Fuel Cell Flow Schematic, Impurities and Specifications

Table 4-12 summarizes 21 of the more promising coal gasification processes resulting from the detailed survey. These processes have been grouped according to their generic gasifier type such as fixed, fluidized, entrained, and molten. The gas composition is shown in Table 4-12 for each system for oxygen and/or air blown systems. These compositions are average values to date representing the raw product gas out of the gasification plant. Additional clean-up is required prior to fuel processing in the fuel cell plant to remove sulfur and other impurities. Impurities in excess of several ppm are detrimental to the performance of the fuel processing and fuel cell systems.

The potential total hydrogen concentration out of a fuel processing system (FPS) was calculated for each of the gasification systems in Table 4-12. A 1.3 H_2O/CO mole ratio was assumed for the fuel processing shift reaction. For every mole of CO reacted, a mole of H_2 and CO_2, respectively, is generated. An additional 30 volume percent of H_2O is assumed to insure complete and total reaction of the CO and to minimize the potential for carbonization in the shift reactor catalyst bed.

For the oxygen blown gasifier cases the average hydrogen concentration after the shift reaction is approximately 50.1 mole %. The air blown cases were somewhat lower due to N_2 inerting but still acceptable for PAFC feed. The average hydrogen concentration for the air blown cases is 29.8 mole % ($\sigma = 3.0\%$). For both the oxygen and air blown cases, the hydrogen feed concentration to the fuel cell seems to be independent, on a first order basis, of the type of gasifier and the kind of coal used.

It is important to have a sufficient concentration of hydrogen in the fuel cell feed gas. Figure 4-2 shows the effect of reduced hydrogen on the overall fuel cell voltage. For the oxygen blown gasifiers the average hydrogen concentration after the FPS is 50 mole %. Depending on the fuel utilization factor (0.8 → 0.9) the voltage loss is 25 → 35 mv. For the air blown gasifiers with a 30 mole % hydrogen concentration, the voltage loss is 38 → 45 mv. The average differential voltage loss between the two cases is approximately 13 mv which amounts to a 2% reduction in overall fuel cell performance for air versus oxygen blown coal gas utilization assuming an operating voltage requirement of 0.66 volts per cell. For an oxygen blown coal gas the degradation in overall fuel cell performance compared to 100 mole of hydrogen feed is approximately 4%. The air blown case is approximately 6%. This means that the fuel cell system needs 4 → 6% more fuel feed gas

Table 4-12

COAL GASIFICATION SYSTEMS AND THEIR PRODUCT GAS (Representative Compositions)

REACTOR TYPE	PROCESS	REACTANTS	BTU/SCF HVV	GAS COMPOSITIONS (MOLE %)						COAL TYPE	H_2 % AFTER SHIFT 1.3 H_2O/CO	COMMENTS
				H_2	CO	CH_4	CO_2	H_2S(COS)	N_2(Ar)			
Fluidized Bed	Battelle Agglomerating Burner	O_2	300 → 350	NO DATA						-	-	Pilot plant O_2 runs; No data reported in literature,
Fluidized	Battelle Agglomerating Burner	Air	≈ 150	50 → 66	0 → 39	1 → 6	3 → 28	-	-	Bituminous Type Coal	-	Pilot unit shutdown 1978; approximately 65% N_2 removed from composition stream.
Fluidized Bed	BCR TRI-Gas	O_2	≈	NO DATA						-	-	PDU unit in U.S.A.; no O_2 information in the literature.
Fluidized Bed	BCR TRI-Gas	Air	≈ 150	15.8	31.2	-	0.5	0.2	52.3	Bituminous Coal	33.4	PDU unit in U.S.A.; no reformer required.
Fluidized Bed	CO_2 Acceptor	O_2	≈ 380	58.8	15.5	13.7	9.1	0	2.9	Lignite	61.8	Work completed in U.S.A.; not commercial; need reformer.
Fluidized Bed	CO_2 Acceptor	Air	-	NO DATA						-	-	Work completed in U.S.A.; not commercial; No information available for air blown case.
Fluidized Bed	CO Gas	O_2	≈ 335	57.9	31.2	4.0	6.6	-	0.3	Ill. #6	63.4	Pilot Plant in England; needs reformed.
Fluidized Bed	CO Gas	Air	-	NO DATA						-	-	Pilot Plant in England; needs reformed.
Fluidized Bed	HY Gas	O_2	370	30.2	23.8	18.6	24.5	1.2	0.1	Bituminous Subbituminous	41.2	Pilot Plant in U.S.A.; very high Methane needs reformer.
Fluidized Bed	HY Gas	Air	-	NO DATA						-	-	Pilot Plant in U.S.A.; very high methane

Table 4-12

COAL GASIFICATION SYSTEMS AND THEIR PRODUCT GAS (Representative Compositions) - (Continued)

REACTOR TYPE	PROCESS	REACTANTS	BTU/SCF HVV	GAS COMPOSITIONS (MOLE %)						COAL TYPE	H_2 % AFTER SHIFT 1.3 H_2O/CO	COMMENTS
				H_2	CO	CH_4	CO_2	$H_2S(COS)$	$N_2(Ar)$			
Fixed Bed	British Gas Lurgi Slag	O_2	370 → 380	27.8	60.6	7.6	2.6	-	1.0	Donisthorpe, Nearly all coals.	49.4	Semi-commercial; Outside U.S.A.; Molten Slag; Methane Conc. High; Will Need Reformer
Fixed Bed	British Gas Lurgi Slag	Air	≈ 150	-----	--------	-NO DATA	------	------------	---	-	-	No data reported
Fixed Bed	Lurgi, Dry Ash	O_2	285 → 300	39.4	16.9	9.0	31.5	0.8	1.6	Pittsburgh #8; Ill. #6; Bituminous; Nearly all coals.	46.2	Commercial; Outside U.S.A.
Fixed Bed	Lurgi, Dry Ash	Air	≈ 150	24.0	16.0	4.0	14.0	1.0	41.0	"	33.1	Commercial; Outside U.S.A.
Fixed Bed	Wellman-Galusha	O_2	270 → 305	-----	--------	-NO DATA	------	------------	---	Anthracite	-	O_2 tests performed - no data reported.
Fixed Bed	Wellman-Galusha	Air	120 → 168	15.0	28.6	2.7	3.4	-	50.3	Bituminous	31.8	Concentration of Methane is low enough that no reformer is needed; 12 in U.S.A.; Commercial units available.
Fixed Bed	Woodall-Duckham/ Gas Integrale	O_2	≈ 280	38.4	37.5	3.5	18.0	0.4	2.2	Bituminous	51.0	Commercial; Outside U.S.A.; Concentrations reported are after scrubbing; Methane concentration such may require a reformer.
Fixed Bed	Woodall-Duckham/ Gas Integrale	Air	175 → 205	17.0	28.3	2.7	4.5	0.3	47.2	Bituminous	33.1	As above; Methane concentration is low enough that a reformer may not be required.

Table 4-12

COAL GASIFICATION SYSTEMS AND THEIR PRODUCT GAS
(Representative Compositions) - (Continued)

REACTOR TYPE	PROCESS	REACTANTS	BTU/SCF HVV	GAS COMPOSITIONS (MOLE %)						COAL TYPE	H_2 % AFTER SHIFT 1.3 H_2O/CO	COMMENTS
				H_2	CO	CH_4	CO_2	H_2S(COS)	N_2(Ar)			
Fluidized Bed	Synthane	O_2	≈ 355	32.3	13.2	15.0	36.2	1.6	-	Ill. #6	38.8	Pilot Plant in U.S.A.; Needs reformer - methane concentration too high.
Fluidized Bed	Synthane	Air	≈ 150	21.5	10.1	5.6	17.9	0.7	43.5	Ill. #6	27.9	Pilot Plant in U.S.A., Needs reformer - methane concentration too high.
Fluidized Bed	U-Gas	O_2	≈ 320	41.4	16.1	35.8	5.9	0.2	0.6	Sub-bituminous	47.6	Pilot Plant in U.S.A.; very high methane; needs reformer.
Fluidized Bed	U-Gas	Air	≈ 154	17.5	19.6	3.4	9.9	0.7	43.5	Bituminous	29.6	Pilot Plant in U.S.A.; may need reformer.
Fluidized Bed	Westinghouse	O_2	≈ 285	29.8	49.1	3.2	17.2	0.5	0.3	Pittsburgh Seam; Eastern Coal.	48.2	PDU U.S.A.; may need reformer.
Fluidized Bed	Westinghouse	O_2	≈ 350	25.6	51.1	10.2	9.9	1.7	0.5	Pittsburgh Seam; Eastern Coal.	46.1	A proposed commercial composition; high methane - needs reformer. (See Ref. 8). Low methane can also be proposed for PAFC application.
Fluidized Bed	Westinghouse	Air	120→150	14.4	19.2	2.7	9.3	0.1	54.3	Eastern Coal	26.9	PDU U.S.A.; no reformer.
Fluidized Bed	Winkler	O_2	260→290	35.3	48.2	1.8	13.8	-	0.9	Lignite	51.3	Commercial units worldwide but outside U.S.A.; Methane low - no reformer needed.
Fluidized Bed	Winkler	Air	≈ 150	14.0	22.0	1.0	7.0	-	56.0	Sub-bituminous	28.0	Commerical units worldwide but outside U.S.A.; Methane low - no reformer needed.

Table 4-12

COAL GASIFICATION SYSTEMS AND THEIR PRODUCT GAS
(Representative Compositions) - (Continued)

REACTOR TYPE	PROCESS	REACTANTS	BTU/SCF HVV	GAS COMPOSITIONS (MOLE %)						COAL TYPE	H_2 % AFTER SHIFT 1.3 H_2O/CO	COMMENTS
				H_2	CO	CH_4	CO_2	$H_2S(COS)$	$N_2(Ar)$			
Entrained Flow	Babcock & Wilcox	O_2	≈ 300	27.9	65.3	-	5.0	0.6	1.2	Pittsburgh #8	50.4	Semi-commercial; U.S.A.; no reformer - methane content low.
Entrained Flow	Babcock & Wilcox	Air	≈ 150	8.4	23.3	-	4.6	0.2	63.5	Pittsburgh #8	24.3	Semi-commerical; U.S.A.; no reformer - methand content low.
Entrained Flow	BI-Gas	O_2	≈ 356	32.0	29.3	15.7	21.5	0.8	0.7	Pittsburgh Seam	44.4	Pilot plant operation; U.S.A.; high methane concentration - needs reformed.
Entrained Flow	BI-Gas	Air	-	NO DATA						-	-	No data in the literature for air blown gasifier.
Entrained Flow	Combustion Engineering	O_2	-	NO DATA						-	-	No data in the literature for oxygen blown gasifier.
Entrained Flow	Combustion Engineering	Air	≈ 120	10.6	24.7	-	4.0	0.5	60.4	Pittsburgh Seam	26.7	PDU system; U.S.A.; no reformer required.
Entrained Flow	Foster-Wheeler	O_2	-	NO DATA						-	-	No data in the literature for oxygen blown gasifier.
Entrained Flow	Foster-Wheeler	Air	≈ 160	14.2	29.1	3.4	3.3	0.7	49.3	Ill. #6	31.4	Pilot facility in U.S.A.; may need reformer.
Entrained Flow	Koppers-Totzek	O_2	≈ 290	36.0	52.5	-	10.0	0.4	1.1	Eastern Coal	52.6	Commercial worldwide outside U.S.A.; no reformer required.
Entrained Flow	Koppers-Totzek	Air	-	NO DATA						-	-	No data found in literature search for air blown case.

Table 4-12

COAL GASIFICATION SYSTEMS AND THEIR PRODUCT GAS (Representative Compositions) - (Continued)

REACTOR TYPE	PROCESS	REACTANTS	BTU/SCF HVV	GAS COMPOSITIONS (MOLE %) H_2	CO	CH_4	CO_2	$H_2S(COS)$	$N_2(Ar)$	COAL TYPE	H_2 % AFTER SHIFT 1.3 H_2O/CO	COMMENTS
Entrained Flow	Texaco	O_2	≈ 250 (279)	39.0 (35.8)	37.6 (44.6)	0.5 (1.0)	20.8 (18.0)	1.5 (0.1)	0.6 -	Eastern; Western; Any Coal.	51.4 (51.0)	Semi-commercial; U.S.A.; no reforming (data provided by the Fuel Cell User's group).
Entrained Flow	Texaco	Air	-	------	--------	--------	NO DATA	--------	-------	-	-	No information in the literature for air blown case.
Molten Bath	Rockwell International	O_2	-	------	--------	--------	NO DATA	--------	-------	-	-	No information in the literature for O_2 blown case.
Molten Bath	Rockwell International	Air	≈ 150	13.2	29.7	1.5	3.5	< 5 ppm	49.4	Any Coal	31.0	Pilot plant; U.S.A.; no reforming.
Molten Bath	Saarberg/ Otto	O_2	≈ 250	30.7	53.6	0.5	14.0	0.5	0.7	Bituminous	49.7	Pilot plant; no reforming.
Molten Bath	Saarberg/ Otto	Air	-	------	--------	--------	NO DATA	--------	-------	-	-	No data in literature for air blown case
FCUG Moving Bed	All	O_2	315	40	20	12	28	-	-	-	47.6	Average Compositions reported for Fuel Cell User's Group Study; reformer needed (See Ref. 5).
FCUG Fluidized Bed	All	O_2	275	42	33	3	20	-	-	-	52.5	Average compositions reported for Fuel Cell User's Group Study; may need reformer.
FCUG Entrained Flow	All	O_2	290	36	53	0.1	9	-	0.8	-	52.7	Average compositions reported for Fuel Cell User's Group Study; no reformer required.
FCUG Entrained Flow	All	Air	100	12	20	0	7	-	61	-	25.4	Average compositions reported for Fuel Cell User's Group Study; no reformer required.

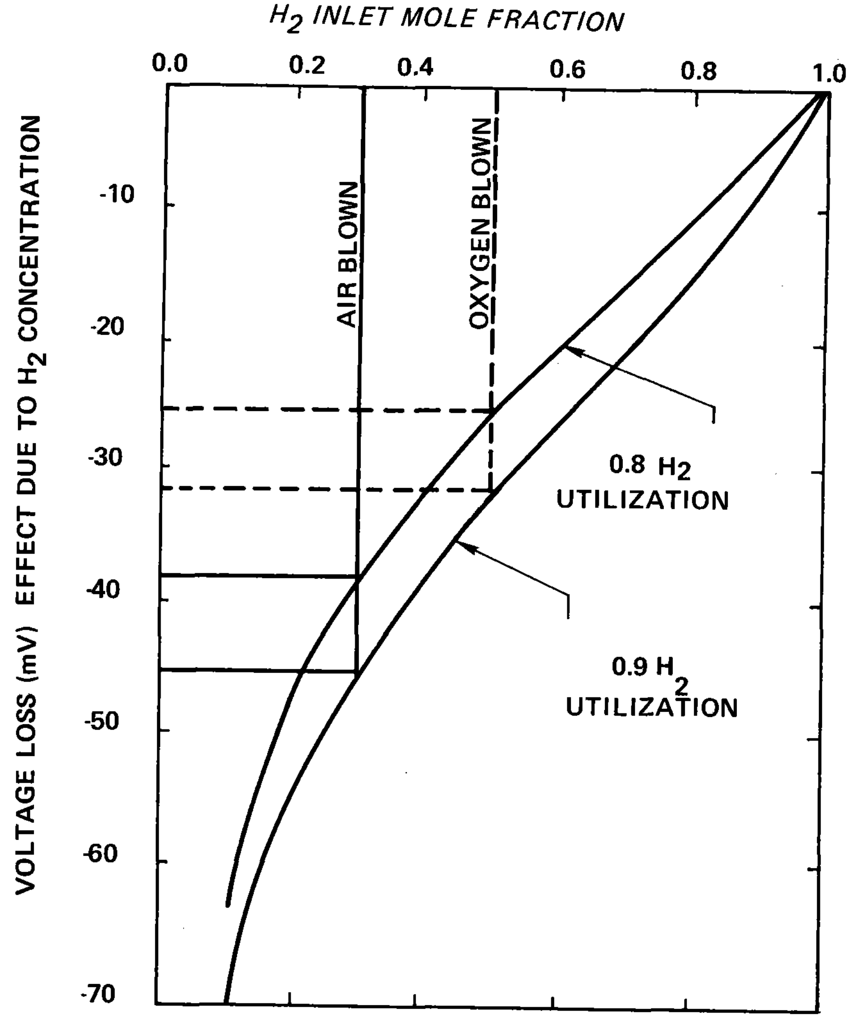


Figure 4-2. PAFC Voltage Loss Due to Decreased Hydrogen Concentration in The Anode Feed Gas

to obtain the same voltage output compared to pure hydrogen feed. Also, in order to utilize this increased fuel flow and maintain a fixed total voltage the number of fuel cells will have to be increased. This will result in approximately a 4 → 6% increase in direct capital cost of the fuel cell system over a pure H_2 feed.

Of the twenty-one coal gasification systems listed in Table 4-12, three representative processes producing medium or low Btu feed gas were chosen. Since the hydrogen concentration to the fuel cell is relatively constant for the low and medium Btu cases, other factors such as the methane content, systems availability, and gas impurities were considered.

One selection was based on using a oxygen blown system with low methane content (less than 3%) which would result in high hydrogen content without the need for high temperature reforming.

A second selection was based on air blown systems with low methane content. Although the low Btu feed to the PAFC plant results in a 2% reduction in fuel cell performance over medium Btu feed, the cost of air blown coal gas can be considerably less expensive. This gasifier produces a low methane fuel, thus no reforming is required. The gasifier is commercially available and twelve of these units are currently operating in the United States.

An oxygen blown system with a product gas high in methane (> 10 volume percent) content was selected as a third system. Since most industrial process applications strive for a high Btu gas, the gas composition may be representative of a large number of gasification systems. Because of the high methane gas concentration fuel reforming may be necessary in the PAFC FPS.

The typical characteristics for the three gasifer compositions are summarized in Table 4-13. For the inital fuel comparisons, properties representative of the medium Btu low methane gasifier (Reference 9) were used. The effects of the alternate compositions will be used in future comparisons.

4.1.4 NAPHTHA

The term "Naphtha" refers to any light oil product having properties intermediate between gasoline and kerosine. It is generally used as a catalytic reformer feed to produce high octane reformate for gasoline blending and aromatics. The range

Table 4-13

REPRESENTATIVE CHARACTERISTICS FOR COAL DERIVED GAS

	Medium Btu Low Methane	Low Btu Low Methane	Medium Btu High Methane
● Composition (Volume Percent)			
H_2	37.8	15.0	25.6
CO	50.4	28.6	51.1
CH_4	0.3	2.7	10.2
CO_2	10.1	3.4	9.9
H_2S	0.1	---	1.7
N_2	1.4	50.3	0.5
● Heating Value (Btu/SFC)	250*	150	350

* Supplied by TVA (Reference 9)

of true boiling point (TBP) for naphtha (also referred as Heavy-Straight-Run Gasoline) is 190°F to 380°F. Typical TBP cut points in the crude oil distillation process for naphtha are:

Initial Boiling Point (°F)	End Point (°F)	Process Use
180	380	Max. reforming cut
190	330	Max. jet fuel operation
220	330	Min. reforming cut

A typical analysis of crude oil from the North Slope of Alaska is shown in Table 4-14. The specific gravity and gravity is shown over the TBP cut range for naphtha. The molecular weight of naphtha varies from 88 lbs/lb mole at 180°F and 59° API to 154 lbs/lb mole at 380°F and 44° API.

The sulfur content of naphtha is dependent upon the percent of sulfur in the crude oil source. Figure 4-3 displays the sensitivity of sulfur content in straight-run product to the temperature level and crude oil sulfur content for typical United States low sulfur crude oil. The applicable levels in sulfur content for naphtha are shown in Table 4-15. Sulfur content in general will be less than 1%.

The limiting specifications based on Imperial Chemical Industries (ICI) recommendations for straight run naphtha are as follows:

EP	420 to 428°F
Aromatics	Below 35%
Olefins	3 Vol % max (Hydrogenation none to reformer)
Sulfur	.05 wt % max

Typical naphtha properties are as follows:

Components

Paraffins	65% Approx.
Olefins	0.1%
Naphthenes	20%
Aromatics	15%
Initial Boiling Point	100°F

Table 4-14

ALASKAN CRUDE PETROLEUM ANALYSIS

North Slope, Alaska, Crude Oil
Gravity, °API, 25.7
Sulfur, wt %, 1.12
Viscosity, SUS at 70°F, 182.5 sec; at 100°F, 94.1 sec
Conradson carbon: Crude, 5.99%: $1,000^{+}$, 19%

Gravity, specific, 0.900 (60/60°F)
Pour point, °F, 20

TBP Distillation

	TBP cut (°F)	Vol % on crude Frac	 Sum	Sp gr (60/60°F)	°API (60°F)
	178-214	2.1	6.2	0.740	59.7
	214-242	2.0	8.2	0.759	55.0
	242-270	2.0	10.2	0.764	53.8
	270-296	2.0	12.2	0.781	49.6
NAPHTHA	296-313	1.0	13.2	0.781	49.6
	313-342	2.0	15.2	0.791	47.3
	342-366	1.9	17.1	0.797	46.0
	366-395	2.0	19.1	0.806	44.0

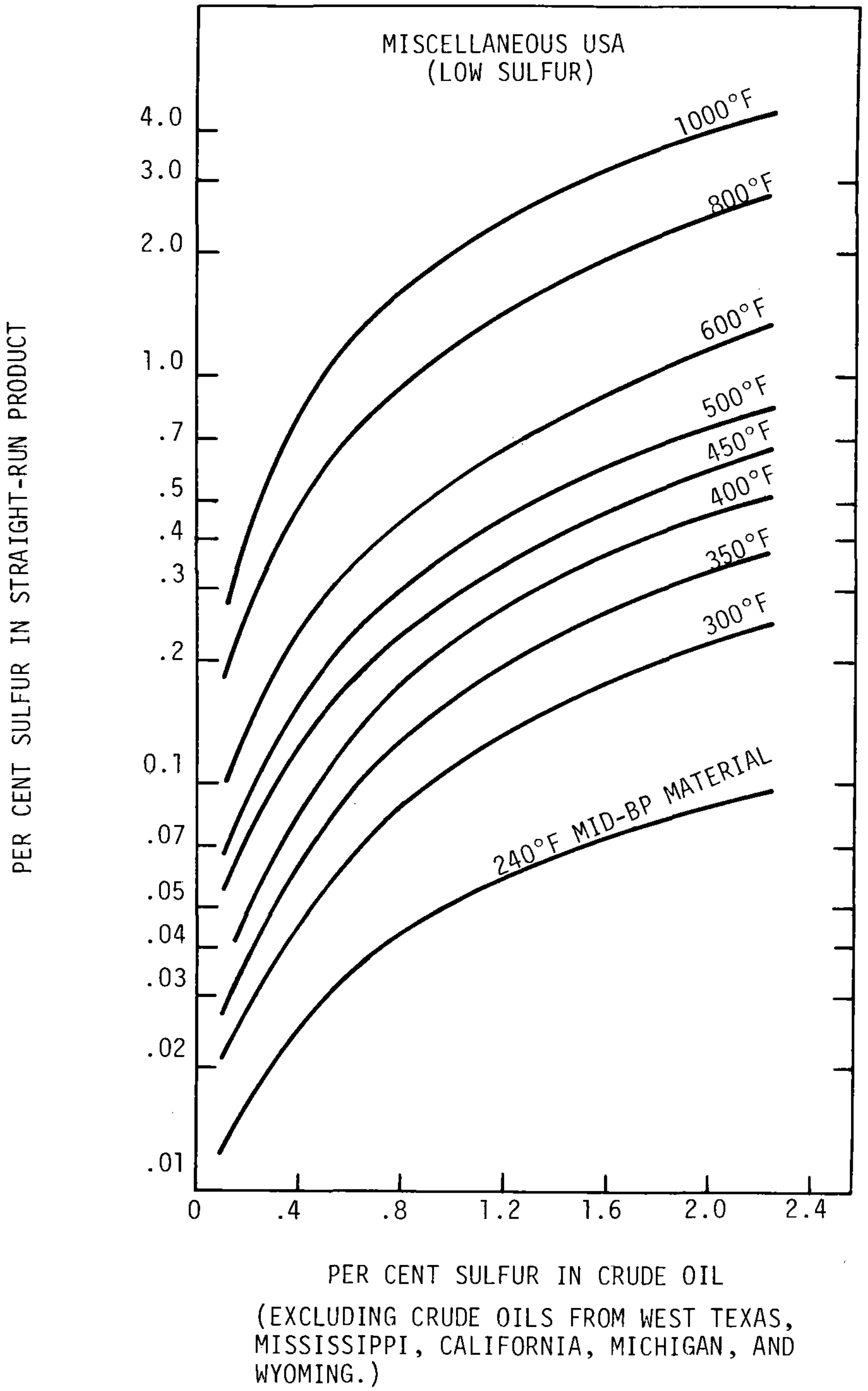


Figure 4-3. Sulfur Content of Products From Miscellaneous U.S. Crude Oils

Table 4-15

SULFUR CONTENT IN VARIOUS CRUDES AND NAPHTHAS

SOURCE OF CRUDE	% S IN CRUDE	% S IN NAPHTHA 289°F MID-BP CUT
Misc. U.S. Low Sulfur	0.1 - 2.2	0.02 - 0.22
West Texas	0.4 - 2.9	0.086 - 0.37
California	0.5 - 4.5	0.028 - 0.62
Middle East	0.3 - 2.6	0.02 - 0.085
Venezuelan	0.2 - 5.0	0.02 - 1.2

End Point	428°F
Sulfur	500 PPM, Max
Carbon to Hydrogen WT:WT	6.0
Molecular Weight	100

The system performance documented in Section 5.1 was calculated based on C_8H_{18} which is representative of this composition.

4.1.5 COAL DERIVED NAPHTHA

Table 4-16 presents the inspections of three boiling range fractions from H-Coal Syncrude from Illinois No. 6 Coal.

The concentration of ring structures in naphtha includes both cycloparaffins (naphthenes) and aromatics. Also of interest are the quantities of principally phenols and non-hydrocarbon constituents. In the case of the C_4 to 400 naphtha fraction, the high naphthenic content makes it potentially suitable for the production of gasoline-blending stock via catalytic reforming. However, hydrorefining of this fraction prior to catalytic reforming will be necessary to increase the saturates.

4.1.6 ETHANOL

Sources of Ethanol Include:

A.) Hydrocarbon Gas (mainly Ethylene) - by hydration.

B.) Saccharine Materials (Molasses, Sugar Cane, Sorgo, Sugar Beets).

C.) Cellulosic Materials (Wood, Agricultural Residue, and the waste sulfite liquor from Paper - Pulp Mill).

D.) Starchy Materials (Cereal Grains, Potatoes, etc.).

Ethanol from B), C) and D) is by fermentation.

Commercial production of ethanol by fermentation is based on the conversion of sugars with six carbon atoms, or C_6 sugars to ethanol by yeast (REFERENCE 15). Fermentation is the decomposition of organic compounds into simpler compounds through the use of enzymes.

Table 4-16

H-COAL PRODUCT INSPECTION FROM ILLINOIS NO. 6 COAL

Syncrude Mode

DISTILLATE FRACTION:	C_4-400°F (Naphtha)	400-650°F (Gas Oil)	650-920°F (Vacuum Gas Oil)
Constituents, wt. %			
Paraffins	11.99	6.5	1.4
Saturated Naphthenes	51.13	24.5	5.4
Unsaturated Naphthenes	11.20	4.3	1.6
Alkyl Benzenes	17.55	12.6	3.0
Other Aromatics*	7.03	47.0	73.3
Heterocyclics+	0.9	2.0	1.5
Other Non-Hydrocarbons	0.2	3.10	13.8
TOTALS	100.00	100.00	100.0

---

*Indans, tetralins, and polycyclic aromatics.

+Principally phenols.

Source: Battelle

Until the end of World War II, the fermentation of molasses, fruits, and grains was the main source of ethanol in the United States. Fermentation is still the only source for ethanol produced for human consumption. Today, however, approximately 80% of all industrial grade (nonbeverage) ethanol in the United States is produced synthetically by the direct hydration of ethylene. In 1977, approximately 202.9 million gallons of ethanol from ethylene, and 50 million gallons of ethanol from fermentation of sugars were produced for the industrial market. Production of beverage-grade ethanol that year was about 30 million gallons (REFERENCE 15). The November 1978 market price for ethanol was $1.18/gallon (delivered) for 190 proof ethanol, and $1.28/gallon (delivered) for 200 proof ethanol (REFERENCE 15). The December 1980 average market price was $1.75/gallon for 200 proof ethanol.

Since ethylene is produced from ethane and propane, components of natural gas, and from naphtha, industrial alcohol production in the United States is based primarily on domestic nonrenewable or imported resources.

More advanced technology for ethanol production is in a relatively early stage of research and development. This newer technology pertains not only to improvement in conventional processes, but particularly to utilization of agricultural and forest residues as raw materials for alcohol production. It is estimated that these technologies should develop to commercial feasibility levels within the next five to ten years. New motor fuel alcohol plants, based on current technology, should be designed and built to facilitate modification of auxiliary installations, to utilize the new technology as it becomes available.

In response to the gasohol demand, fermentation alcohol in the United States is growing rapidly. Over the near term, it is believed that fermentation alcohol producers will avoid the industrial ethanol market and concentrate on gasohol.

The following data is from Reference 15.

Maximum Projected Ethanol Production from Biomass
(Billion gallons per year)

	Year	
	1990	2000
Food processing wastes	0.7	0.9
Grains	2.8	2.3
Sugar	3.7	9.0
Wood	20.2	25.8
Agricultural residue	11.3	13.1
MSW (Municipal Solid Waste)	2.5	2.9
Total potential from all biomass	41.2	54.0

The following composition was assumed to be representative for ethanol produced from standard feedstocks, and was used in the present study.

Ethanol from Fermentation:

COMPOSITION	WEIGHT PERCENT
C_2H_5OH	95.6
H_2O	4.4
Higher Alcohols	Trace

Ethanol from Hydration of Ethylene:

COMPOSITION	WEIGHT PERCENT
C_2H_5OH	95.6
H_2O	4.4
Ether	Trace

The physical and chemical properties of ethanol are shown in Table 4-17.

4.2 ASSESSMENT OF FUEL PROCESSING SUBSYSTEMS

Westinghouse Advanced Energy Systems Division (AESD) directed a specification to potential vendors to obtain Fuel Processing Subsystem (FPS) cost and performance information for the five fuels described in Section 4.1. An objective assessment

Table 4-17

PHYSICAL AND CHEMICAL PROPERTIES OF ETHANOL*

Constant	Absolute alcohol	95% Alcohol
Melting Point,°C.	-112.3	
Boiling Point,°C.	78.4	
Specific Gravity	0.78510	
Viscosity at 20°C., poise	0.0122	0.0141
Surface tension, dynes/cm.	22.3	22.8
Specific Heat	0.581	0.618
Heat of fusion, cal./gram	24.9	
Heat of evaporation at b.p.,		
cal./gram	204	
B.t.u./lb.	368	
Heat of combustion, kg.-cal./gram-mole	328	
Flash point (A.S.T.M. open cup), °C	18.3	14
Explosive limits in air, per cent	3.7-13.7	
Apparent ignition temp. in air, °C	371-427	400-432
Electrical conductivity at 25°C., ohm $^{-1}$/cm.	1.35×10^{-9}	

* As Available From Volume 9 of the Encyclopedia of Chemical Technology

of the impact of alternative source fuels on the overall power plant requires that source fuels and their FPS be compared on an equal basis. The specification in Appendix A was prepared for this purpose.

Seven vendors were selected from an initial list of thirty three potential vendors who were contacted by Westinghouse during an earlier appraisal of their interest and capabilities. Responses were obtained from four of the vendors contacted. For each of the source fuels the potential FPS vendors were asked to provide the following type of information:

1) The price of the FPS
2) Operation and maintenance cost estimates for the FPS
3) Utility requirements to operate the FPS
4) Status of the FPS technology
5) Other performance and cost information as indicated in Tables 4-18 and 4-19

The requirements outlined for the FPS vendors considered both functional and process requirements. The top level functional requirements specified that the FPS shall convert the source fuel(s) into an acceptable hydrogen rich feedstream for the fuel cell subsystem. Also indicated was that a high efficiency FPS design is preferred in order to maximize the overall efficiency of the fuel cell power plant and that the FPS shall perform source fuel cleanup including desulfurization as required.

Cooling water, boiler feed water, electrical power, air, inerting gas, etc., were specified by the FPS vendor at rates and conditions as required.

The process requirements specified that the FPS produce the hydrogen rich feed gas at the following conditions:

H_2 production rate - 227,000 SCFH
Feed stream pressure - 200 psia*
Feed stream temperature - 300-400°F
H_2 gas content - 50 volume percent

---

*Higher feed stream pressure allows for better control of fuel cell pressure at 50 psia.

Table 4-18

SUMMARY OF VENDOR FPS DATA FOR NATURAL GAS AND METHANOL
(Hydrogen Production Rate 227,000 SCFH)

PARAMETERS	NATURAL GAS				METHANOL			
Vendor	A	B	C	D	A	B	C	D
Fuel	Natural Gas	Natural Gas	Natural Gas	Natural Gas	Methanol	Methanol	Methanol	Methanol
Process Fuel(3)	60 M SCFH	86.7×10^6 Btu/Hr	90.4×10^6 Btu/Hr	58 M SCFH (56.5×10^6 Btu/Hr)	6500 lb/hr	No Commercial Experience	No Commercial Experience	No Commercial Experience
Heat Fuel	45 M SCFH	(1)	(1)	41 M SCFH 40.2×10^6 Btu/Hr 10 M SCFH(4) 10×10^6 Btu/Hr	380 lb/hr			
Power, KWH	240	100	227	272	--			
Cooling Water, GPM (°F Rise)	500 (30°F)	500 (25°F)	-- (--)	300 (--)	-- (--)			
Boiler Feed Water	100 GPM	--	11,350 lb/hr	30 GPM	--			
Export Steam, PSIG/lbs	6000	--	--	--	0			
FPS Capital Cost(2)	$\$5.7 \times 10^6$	$\$5.4 \times 10^6$	$\$2.9 \times 10^6$	$\$3.1 \times 10^6$	$\$4.1 \times 10^6$	$\$4.1 \times 10^6$		$\$2.3 \times 10^6$
Initial Charge Catalyst & Absorbents	$240 K	$264 K	--	--	$250 K			
Total Annual O&M Cost	$585 K	--	$203 K	$193 K	--			
Annual Catalyst &(5) Absorbents Cost	$20 K	$53 K	$24 K	$31 K	--			

(1) Heat fuel included in process feed fuel
(2) Installed cost including initial charge of catalyst and absorbents. All cost numbers 1980 $.
(3) 64×10^6 Btu/Hr per 227,000 SCFH of H_2 (23,748 Btu/lb Natural Gas)
(4) With anode exhaust as a fuel
(5) Annual catalyst, labor, and major maintenance item costs are included in the total annual O&M cost.

Table 4-18

SUMMARY OF VENDOR FPS DATA FOR NATURAL GAS AND METHANOL
(Hydrogen Production Rate 227,000 SCFH) - (Continued)

PARAMETERS	NATURAL GAS				METHANOL			
Vendor	A	B	C	D	A	B	C	D
Fuel	Natural Gas	Natural Gas	Natural Gas	Natural Gas	Methanol	Methanol	Methanol	Methanol
Annual O&M Labor Cost	$90 K	--	$121 K	$96 K	--	No Commercial Experience	No Commercial Experience	No Commercial Experience
Annual O&M Outage Hours	300	200	100	--	--			
Major Maintenance Items	--	Reformer Tubes, Catalyst	Reformer Tubes	Reformer Tubes, Pumps, Heaters	--			
FPS Operating Life, Years	3	15	20	20	2			
Lead Time to FPS Operation, Months	24	16	13	--	24			
FPS Envelope Size, Ft.	50' X 200'	50' X 80'	80' X 100'	--	50' X 50'			
Turn down Capability, %	30%	20%	33%	25%	50%			
Load Change, %/Hour	20%/Hr	40%/Hr	--	--	--			
Start up/Shut down, Hours	24	4→8	8/1	--	12			
Technical Status	Commercial	Commercial	Commercial	Commercial	Developmental	Near Term	Near Term	Near Term

Table 4-19

SUMMARY OF VENDOR FPS DATA FOR COAL GAS, NAPHTHA, AND ETHANOL
(Hydrogen Production Rate 227,000 SCFH)

PARAMETERS	COAL GAS				NAPHTHA				ETHANOL			
Vendor	A	B	C	D	A	B	C	D	A	B	C	D
Fuel	Coal Gas	Coal Gas	Coal Gas	Coal Gas	Naphtha	Naphtha	Naphtha	Naphtha	Ethanol	Ethanol	Ethanol	Ethanol
Process Fuel(3)	304 M SCFH	No Commercial Experience	No Commercial Experience	No Commercial Experience	3000 lb/hr	94.4 X 10^6 Btu/hr	90.4 X 10^6 Btu/Hr	2860 lb/hr	No Commercial Experience	No Commercial Experience	No Commercial Experience	No Commercial Experience
Heat Fuel	15 M SCFH				1820 lb/hr	(1)	(1)	39 X 10^6 Btu/Hr				
Power, KWH	200				280	120	227	8 X 10^6 Btu/Hr				
Cooling Water, GPM (°F Rise)	300 (30°F)				800 (--)	600 (25°F)	-- (--)	250 291				
Boiler Feed Water	650 GPM				25 GPM	--	11,350 lb/hr	34 GPM				
Export Steam, PSIG/lbs	--				1200	--	--	--				
FPS Capital Cost(2)	\$2.6 X 10^6	<\$4.1 X 10^6			\$7.3 X 10^6	\$6.0 X 10^6	\$3.0 X 10^6	\$3.1 X 10^6				\$2.3 X 10^6
Initial Charge Catalyst & Absorbents	\$240 K				\$280 K	\$360 K	--	--				
Total Annual O&M Cost	--				\$900 K	--	\$213 K	\$193 K				
Annual Catalyst &(5) Absorbents Cost	--				\$25 K	\$72 K	\$25 K	\$31 K				

Table 4-19

SUMMARY OF VENDOR FPS DATA FOR COAL GAS, NAPHTHA, AND ETHANOL
(Hydrogen Production Rate 227,000 SCFH) - (Continued)

PARAMETERS	COAL GAS				NAPHTHA				ETHANOL			
Vendor	A	B	C	D	A	B	C	D	A	B	C	D
Fuel	Coal Gas	Coal Gas	Coal Gas	Coal Gas	Naphtha	Naphtha	Naphtha	Naphtha	Ethanol	Ethanol	Ethanol	Ethanol
Annual O&M Labor Cost	--	No Commercial Experience	No Commercial Experience	No Commercial Experience	--	--	$126 K	$96 K	No Commercial Experience	No Commercial Experience	No Commercial Experience	No Commercial Experience
Annual O&M Outage Hours	--				300	200	100	--				
Major Maintenance Items	--				--	Reformer Tubes, Catalyst	Reformer Tubes,	Reformer Tubes, Pumps, Heaters				
FPS Operating Life, Years	2				3	15	20	20				
Lead Time to FPS Operation, Months	18				24	18	13	--				
FPS Envelope Size, Ft.	50' X 50'				50' X 200'	50' X 100'	80' X 100'	--				
Turn down Capability, %	0				50%	20%	33%	25%				
Load Change, %/Hour	--				10%/Hr	40%/Hr	--	--				
Start up/Shut down, Hrs	12				24	4→8	8/1	--				
Technical Status	Near Term	Near Term	Near Term	Near Term	Commercial	Commercial	Commercial	Commercial	Developmental	Near Term	Near Term	Near Term

The diluents in the feed gas can include CO_2, N_2, He, Ar, H_2, and CH_4 as long as the total diluent concentration does not exceed 50 volume percent.

Impurities which have an adverse effect on the FPS and fuel cell operation were limited to the following maximum values:

IMPURITY	MAX. ALLOWABLE
CO	1.0% by volume
H_2S	100 ppm by volume
COS	100 ppm by volume
C_2^+	100 ppm by volume
NH_3	1 ppm by volume
Cl	1 ppm by volume
Metal Ions	1 ppm by weight

Other FPS process requirements specified that 85% (by weight) of the hydrogen in the feed stream was utilized by the fuel cell subsystem. This allows for the fuel cell exhaust stream to be returned to the FPS for the unused hydrogen gas to be combusted or recovered. The representative conditions and compositions of the feed and exhaust streams given to the FPS vendors were as follows:

MOLE FRACTION	FEED STREAM	EXHAUST STREAM
H_2	0.726	0.346
H_2O	0.079	0.189
CO	0.010	0.023
CO_2	0.174	0.415
N_2	0.001	0.004
CH_4	0.010	0.023
H_2 SCFH	227,000	34,000
Pressure, psia	200	100
Temperature, °F	300-400	350-425

Methods for using the anode exhaust stream in the FPS were FPS designer options.

The source fuel compositions identified for the FPS vendors to consider are given in Section 4.1. Representative commercial fuel compositions for methanol, natural gas and coal gas are presented along with their appropriate heating values.

Table 4-18 and 4-19 summarize the data received from the four FPS vendors.

Table 4-20 is a summary of the four vendor's FPS installed capital cost data with additional vendors data for different system sizes included. The performance information and prices given were based on current designs and standard industry practices. All costs are stated in 1980 dollars and do not include escalation to reflect anticipated future delivery or installation dates. The FPS information was used to help establish system capital costs for the conceptual design layouts for each of the utility PAFC plants presented in Section 5.1. This vendor survey along with previous cost estimates, see Table 4-20, was used to establish FPS direct capital costs for the candidate fuels in either integrated or nonintegrated configurations.

Figure 4-4 shows the direct capital cost (based on the Table 4-20 data) as a function of hydrogen generating capacity. The data shown is for a nonintegrated natural gas FPS. The costs are in 1980 dollars and are for prototype designs. For the nonintegrated natural gas plant producing 6.95 MW_e AC, the FPS cost is 410 $/kW or $2,850,000. The methanol curve also given in this figure and is shown to be 20 to 30% less than the natural gas FPS. This curve was based on vendor predicted estimates of what a methanol and ethanol systems might cost since none has been built commercially. Again, based on the vendor survey and previous predicted estimates, the coal gas FPS will be slightly less than that for the methanol system. Four vendors have suggested that a typical naphtha FPS may be 10% higher in cost than a natural gas FPS.

The reasons for the lower methanol, ethanol, and coal gas FPS costs over natural gas are due to the reduced operating temperature and the simpler design for methanol reforming/shift converting and coal gas shift reaction. Reasons that a naphtha FPS is more expensive than for natural gas include the need for a more elaborate sulfur clean up system, fuel vaporizer, more expensive catalyst and that the system has to be sized to handle a higher water/carbon ratio due to the presence of more aromatics. The key operating characteristics, flow schematics, and specifications for the FPS for each of the fuels are discussed in Section 5.1.

For the methanol and natural gas integrated systems all costs were estimated to be 50% greater than the nonintegrated arrangement. This estimate was based upon the complexity in design with integrating the key FPS components with balance-of-plant (BOP) systems, i.e., cathode recirculation system. However, in order to arrive

TABLE 4-20

PROTOTYPE FUEL PROCESSING SYSTEM INSTALLED CAPITAL COST DATA AS SUPPLIED BY VARIOUS VENDORS

SYSTEM SIZE kW(e) d.c.	FUEL	COST(1) $$10^3$
7,500	Natural Gas	2,486
7,500	Methanol	1,243
40	Nautral Gas	140
40	Methanol	70
5,000	Natural Gas	1,700
20,000	Natural Gas	583
165	Natural Gas	317
500	Natural Gas	528
1,000	Natural Gas	730
3,315	Natural Gas	1,400
7,500	Natural Gas	5,393
7,500	Methanol	4,045
7,500	Coal Gas	4,045(2)
7,500	Natural Gas	5,740
7,500	Methanol	4,200
7,500	Coal Gas	2,590
165	Natural Gas	610(3)
826	Natural Gas	1,157(3)
8,260	Natural Gas	4,845(3)
7,500	Natural Gas	2,850
7,500	Natural Gas	3,058
7,500	Methanol	2,294
2	Natural Gas	41

---

1) Installed cost. Includes all fees, 1980$

2) Slightly less than methanol FPS cost.

3) CO_2 removal (99% H_2)

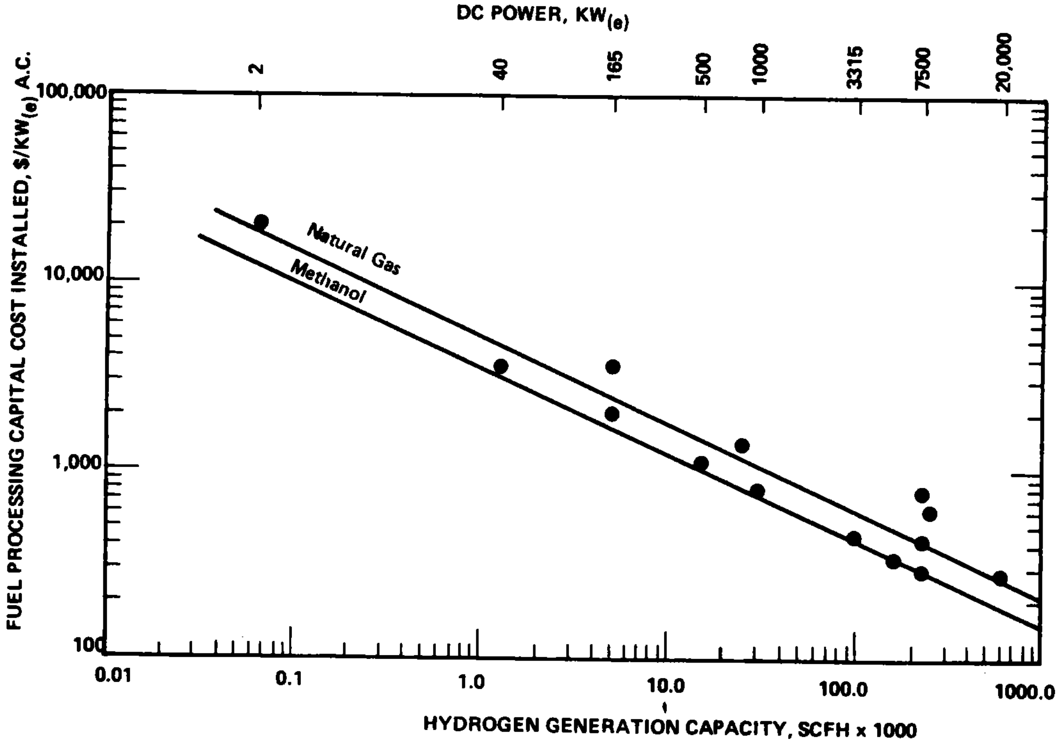


Figure 4-4. Prototype Fuel Processing System Direct Capital Cost (1980 $)

at a better cost for an integrated FPS, further vendor contact has to be made. Action needs to be taken to inform the various vendors of the individual integrated system arrangements. The vendors would then respond with manufacturing costs to fit the specific design.

For coal gas, the integrated FPS cost remained the same as the nonintegrated. This was due to the system arrangement whereby the FPS auxiliary components such as the gas turbine/generator system and CO_2 scrubber system were listed separately.

5.0 SYSTEM ANALYSIS AND ECONOMICS

5.1 SYSTEM ANALYSIS AND PERFORMANCE

The overall objective of the system analysis subtask was to assess the impact of the selected fuels on system performance. At least two types of designs (integrated and nonintegrated FPS) were considered for each fuel. A nonintegrated design is defined as a plant where the fuel processing subsystem is a complete module that receives fuel, liquid water, air and effluent from the anode of the fuel cell and provides all vaporization and major preheating requirements internal to the module. The nonintegrated design has the advantage that the FPS module can be replaced if a different fuel is selected, without major modifications to the rest of the fuel cell power plant. This generally results in lower plant capital costs. The integrated designs utilize reject heat (cathode side) to the extent practical to supply thermal requirements for the fuel processing subsystem. As a result the integrated designs are more efficient, although they generally have higher direct capital costs.

The assumptions listed in Table 5-1 are used as a base in this analysis. The forced draft dry cooling tower increases the heat rate for all the concepts by about 200 Btu/kWhr. A performance comparison for each nonintegrated system is listed in Table 5-12.

TABLE 5-1
ASSUMPTIONS USED IN SYSTEM ANALYSIS

Design Fuel Cell Power Level = 7.5 MW_e (DC)
Power Conditioner Efficiency = 0.96
Ambient Temperature = 80°F
Heat Rejection Method = Dry Cooling Tower
Make Up Water = None Required

5.1.1 NONINTEGRATED SYSTEMS

METHANOL FUELED SYSTEM

The nonintegrated methanol conceptual design flow schematic is shown in Figure 5-1, and the state points at the design rating are shown in Table 5-2. Hydrogen rich gas enters the fuel cell anode where 80% of the H_2 is consumed, and the H_2O formed passes out of the fuel cell with the cathode gas. The anode exhaust is burned with 50% excess air which produces a combustion temperature of about 2000°F, a moderate temperature that should produce very low levels of NO_x. The thermal energy in the combustion gas is utilized in the fuel processing subsystem for the endothermic reforming of the fuel, the generation of steam, and the vaporization of the methanol. The fuel processing subsystem is a complete module that receives fuel, hot water, air, and effluent from the anode of the fuel cell and produces a hydrogen rich gas for the fuel cell. The exhaust from this module has a high relative humidity and is directed to and mixed with air from the cooling tower so that a vapor plume will not be formed. Recirculation of exhaust gases in the fuel processing subsystem (flow stream (34)) may be necessary to control temperatures in the reformer.

The fuel cells are cooled by flow stream (12) to an operating average temperature of 350°F. The cathode effluent (13) is cooled in the low pressure steam generator. The steam generated is used to drive the steam turbine. At the design point, approximately six percent of the generated steam bypasses the turbine. The portion of the cathode effluent that is not recirculated is passed through an expander which provides some energy to the rotating group and also reduces the streams temperature so that the stream is supersaturated. The separator, with an efficiency of 75%, then recovers 5660 lb/hr of water (19), which is 840 lb/hr greater than the required input water. The exhaust stream (18) from the separator is supersaturated and is directed to the cooling tower inlet. The exhaust stream mixes with heated air thus preventing a vapor plume.

A direct contact condenser along with a dry cooling tower is used for removal of reject thermal energy. The exhaust steam from the turbine loses heat to the cooling water through condensation. A portion of the resulting condensate (22) is returned to the water storage and treatment, while the remaining condensate is pumped to the air cooled heat exchanger and returned to the condenser. To protect the condensate system from air-in leakage in the lines and heat exchanger, the condensate is pumped to above atmospheric pressure. In this dry cooling application, the turbine exhaust pressure is dependent on the ambient dry bulk

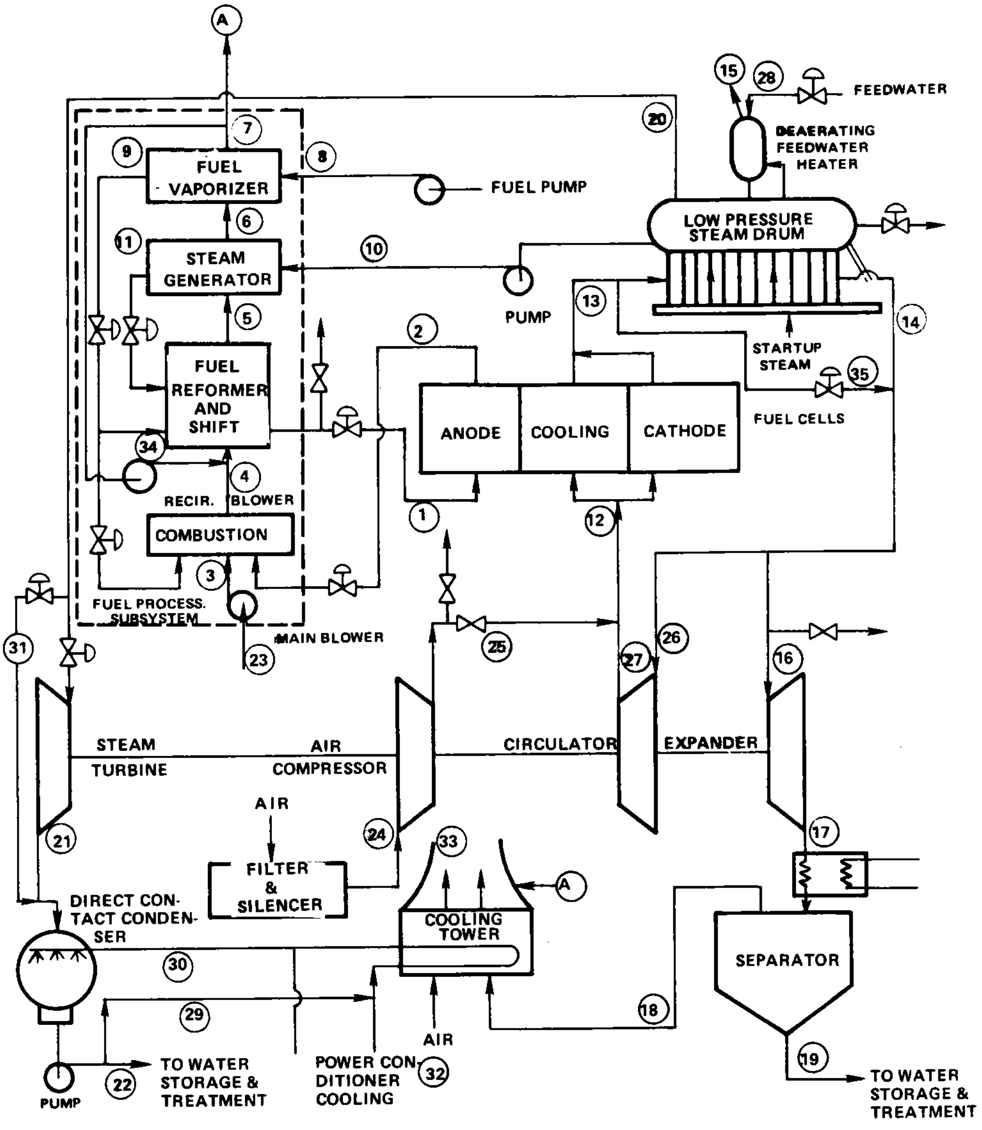


Figure 5-1. Methanol Fueled Nonintegrated Conceptual Fuel Cell Flow Schematic

Table 5-2

STATE POINTS METHANOL FUELED NONINTEGRATED SYSTEM

			Mole Fraction							
Station	Temp. °F	Pressure psia	H_2	H_2O	CO	CO_2	O_2	N_2	CH_3OH	Flowrate lb/hr
1	350	50	0.693	0.075	0.005	0.227	-	-	-	10840
2	350	50	0.311	0.168	0.011	0.510	-	-	-	9910
3	92	16	-	0.010	-	-	0.208	0.782	-	12720
4	2015	15	-	0.246	-	0.258	0.040	0.456	-	22630
5	1383	15	-	0.246	-	0.258	0.040	0.456	-	22630
6	736	15	-	0.246		0.258	0.040	0.456	-	22630
7	190	15	-	0.246	-	0.258	0.040	0.456	-	22630
8	80	80	-	-	-	-	-	-	1.000	6260
9	300	75	-	-	-	-	-	-	1.000	6260
10	240	105	-	1.00	-	-	-	-	-	4580
11	340	100	-	1.00	-	-	-	-	-	4580
12	277	50	-	0.187	-	-	0.100	0.713	-	1103350
13	352	50	-	0.197	-	-	0.094	0.709	-	1104190
14	269	49	-	0.197	-	-	0.094	0.709	-	1104190
15	249	30	-	1.000	-	-	-	-	-	240
16	269	49	-	0.197	-	-	0.094	0.709	-	65160
17	138	15	-	0.175g 0.022ℓ	-	-	0.094	0.709	-	65160
18	96	15	-	0.054g 0.042ℓ	-	-	0.106	0.798	-	60190
19	96	15	-	1.000	-	-	-	-	-	4970
20	249	30	-	1.000	-	-	-	-	-	24110
21	141	3	-	1.000	-	-	-	-	-	22610
22	135	3	-	1.000	-	-	-	-	-	24110
23	80	14.6	-	0.010	-	-	0.208	0.782	-	12720
24	80	14.6	-	0.010	-	-	0.208	0.782	-	64320
25	348	50	-	0.010	-		0.208	0.782	-	64320
26	269	49	-	0.197	-	-	0.094	0.709	-	1039030
27	273	50	-	0.197	-	-	0.094	0.709	-	1039030
28	135	30	-	1.000	-	-	-	-	-	28930
29	136	3	-	1.000	-	-	-	-	-	773930
30	106	15	-	1.000	-	-	-	-	-	773930
31	244	3	-	1.000	-	-	-	-	-	1500
32	80	15	-	0.010	-	-	0.208	0.782	-	4.8×10^6
33	100	15	-	0.010	-	-	0.208	0.782	-	4.8×10^6

g - vapor   ℓ - liquid

temperature (80°F), the approach temperature difference (25°F), a range of 30°F, and a terminal temperature difference of 6°F. These parameters result in an exhaust pressure of six inches of mercury absolute (141°F). The condenser uses a vacuum pump to remove noncondensibles.

The rotating group consists of a steam turbine, air compressor for cathode supply, recirculator, and expander. The steam turbine is small (approximately 930 hp) and is modeled with three stages and a 25" wheel (5000 rpm). The combustion air fan is driven by an electric motor and is included in the fuel processing subsystem. The compressor for the cathode air requires the most power (approximately 1600 hp), while the circulator requires only about 570 hp. The expander produces approximately 1350 hp. At rated conditions, the steam turbine and expander can produce more power than that required by the compressor and circulator; therefore, at rated conditions some steam (approximately six percent) is bypassed (13) around the steam turbine. At part power the compressor, circulator and expander powers are approximately proportional to flow rate cubed over pressure level, while the steam generation is approximately proportional to power level. Therefore, excess steam is available to accelerate the rotating group.

The conceptual design of the steam generator is that of a recirculating low quality steam boiler with both a steam separator and a deaerating feedwater heater. The design of this boiler can use highly finned tubes on the recirculating gas side (due to the clean characteristic of the gas) to produce an efficient compact heat exchanger. The pinch point temperature difference is 20°F and the log mean temperature difference is 72°F with a pressure loss ($\Delta P/P$) of two percent.

The conceptual design performance summary is given in Table 5-3. The power output at the dc bus is 7500 kW. The power conditioner efficiency is assumed to be 96 percent so that power available at the ac bus is 7200 kW. The parasitic loads are 252 kW so that the net output power is 6948 kW. The overall efficiency is 39 percent. With once-through cooling the efficiency would be 40 percent.

Startup of the power plant is accomplished by using 15 psig steam from an auxiliary boiler. This steam is directed to flow up through the tubes of the steam generator. This heats the water to saturation and pressurizes the steam lines. The condensing of this steam requires that some water be removed to maintain the proper water level in the steam drum. Once the steam generator is

TABLE 5-3

PERFORMANCE SUMMARY METHANOL FUELED NONINTEGRATED

Gross Electrical Output kW_e (DC)	7500
Gross Electrical Output kW_e (AC)	7200
Parasitic Losses	
Pumps	23
Fans (Cooling Tower)	210
Vacuum Pump	3
Air Compressor & Dryer	1
Controls	15
Net Electrical Output kW_e (AC)	6948
Input Energy (hhv = 9758 Btu/hr) Btu/hr	61.08×10^6
Overall Plant Efficiency	0.39
Heat Rate (Btu/kW-hr)	8791
Design Current Density (amp/ft^2)	300
Fuel Utilization Factor	0.8
Design Cell Voltage (volts/cell)	0.66

pressurized, the air compressor is valved off at (25) and vented to the atmosphere, and thevalve at (16) is opened to the atmosphere. The steam turbine can then be started, which drives the rotating group (circulator).

Also, heat from the steam generator can be transferred to the fuel cell by the recirculating gas. The recirculating gas temperature can be controlled by utilizing the bypass (35) to ramp up the temperature in a controlled manner. After the fuel cells are heated to about 250°F, the fuel processing system can be started using methanol in the endothermic reformer to produce fuel gas. The fuel gas can then be introduced into the anode while the air compressor vent valve is closed and the valve at (25) is opened to introduce air into the recirculating stream for the cathode. The vent valve at (16) for the expander is closed, and the startup steam is shutoff. This puts the plant into the normal control mode.

ETHANOL FUELED SYSTEM

Figure 5-1 shows the methanol fuel cell conceptual flow schematic with the methanol fuel processing subsystem encircled with a dashed line. Figure 5-2 shows a conceptual ethanol fuel processing subsystem flow schematic to replace the methanol fuel processing subsystem. State points at the design rating for the ethanol fuel processing subsystem and those state points for the rest of the plant which differ from the methanol nonintegrated system are shown in Table 5-4.

Ethanol fuel is vaporized, reformed to supply a hydrogen rich gas to the fuel cell anode, and burned to supply process heat. Liquid ethanol (8) at 80°F is vaporized by heat exchange with a combustion gas (6) to a temperature of 300°F. The resulting vapor stream (9) is split into two streams, (37) and (38). 91 percent of the ethanol vapor (37) enters the fuel reformer after being mixed with 340°F steam in a ratio of 3.9 moles of steam per mole of ethanol. This ratio represents 30 percent excess steam in the mixture. A fuel gas at 350°F that contains 66.9 percent hydrogen (1) is produced in the reforming process and is directed to the fuel cell anode where 80 percent of the hydrogen is utilized. The other 9 percent of the ethanol vapor (38) is combined with the exhaust gas from the fuel cell anode (2) along with 50 percent excess air (3) and is burned to produce a 2155°F combustion gas.

The combustion gas provides heat for the endothermic fuel reforming steam generation, ethanol vaporization, and feedwater preheating. The combustion gas is

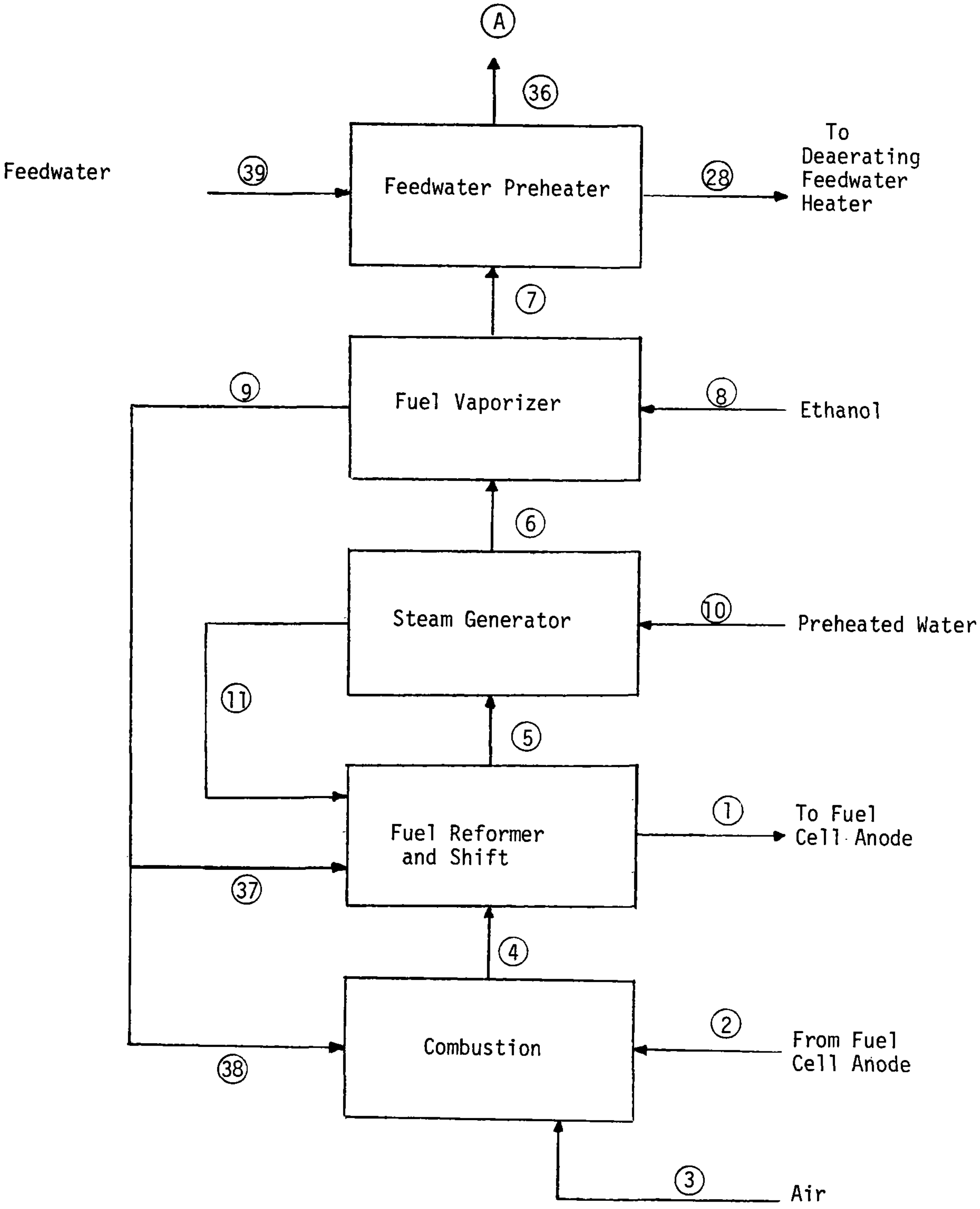


Figure 5-2. Ethanol Fuel Processing Subsystem

Table 5-4

STATE POINTS ETHANOL FUEL PROCESSING SUBSYSTEM

			Mole Fraction							
Station	Temp. °F	Pressure psia	H_2	H_2O	CO	CO_2	O_2	N_2	C_2H_5OH	Flowrate lb/hr
1*	350	50	0.669	0.106	0.005	0.220	-	-	-	11370
2*	350	50	0.288	0.228	0.011	0.473	-	-	-	10438
3	92	16	-	0.046	-	-	0.200	0.754	-	17842
4	2155	15	-	0.268	-	0.216	0.038	0.478	-	28705
5	1376	15	-	0.268	-	0.216	0.038	0.478	-	28705
6	589	15	-	0.268	-	0.216	0.038	0.478	-	28705
7	220	15	-	0.268	-	0.216	0.038	0.478	-	28705
8	80	80	-	-	-	-	-	-	1.000	4926
9	300	75	-	-	-	-	-	-	1.000	4926
10*	240	105	-	1.000	-	-	-	-	-	6869
11	340	100	-	1.000	-	-	-	-	-	6869
28	160	30	-	1.000	-	-	-	-	-	31220
36	158	15	-	0.268	-	0.216	0.038	0.478	-	28705
37	300	75	-	-	-	-	-	-	1.000	4501
38	300	75	-	-	-	-	-	-	1.000	425
39	135	30	-	1.000	-	-	-	-	-	31220

* Interface points

** See Figures 5-1 and 5-7

cooled to 1376°F by the reformer, 589°F by the steam generator, 220°F by the fuel vaporizer, and 158°F by the feedwater preheater. Although the combustion gascontains 27 percent water vapor, condensation does not occur at 158°F because the partial pressure of water is less than the vapor pressure. Thus, a pinch point in the feedwater preheater is avoided. Recirculation of the combustion gas may be necessary to control temperatures in the reformer.

The impact of an ethanol fuel processing subsystem versus a methanol fuel processing subsystem is small. The flow rate of water at interface point (10) for the ethanol subsystem is 50 percent greater than that required for the methanol subsystem. This requires more energy for preheating the water provided to the ethanol processing subsystem. However, by first directing the feedwater (39) through a heat exchanger warmed by 220°F combustion gas leaving the ethanol fuel vaporizer (7) (heats the water from 135°F to 160°F) before introducing the feedwater to the deaerating feedwater heater (28), the energy requirement is satisfied. Another impact of an ethanol fuel processing subsystem is that the flow rate of air (3) for combustion is 40 percent greater than for the methanol subsystem.

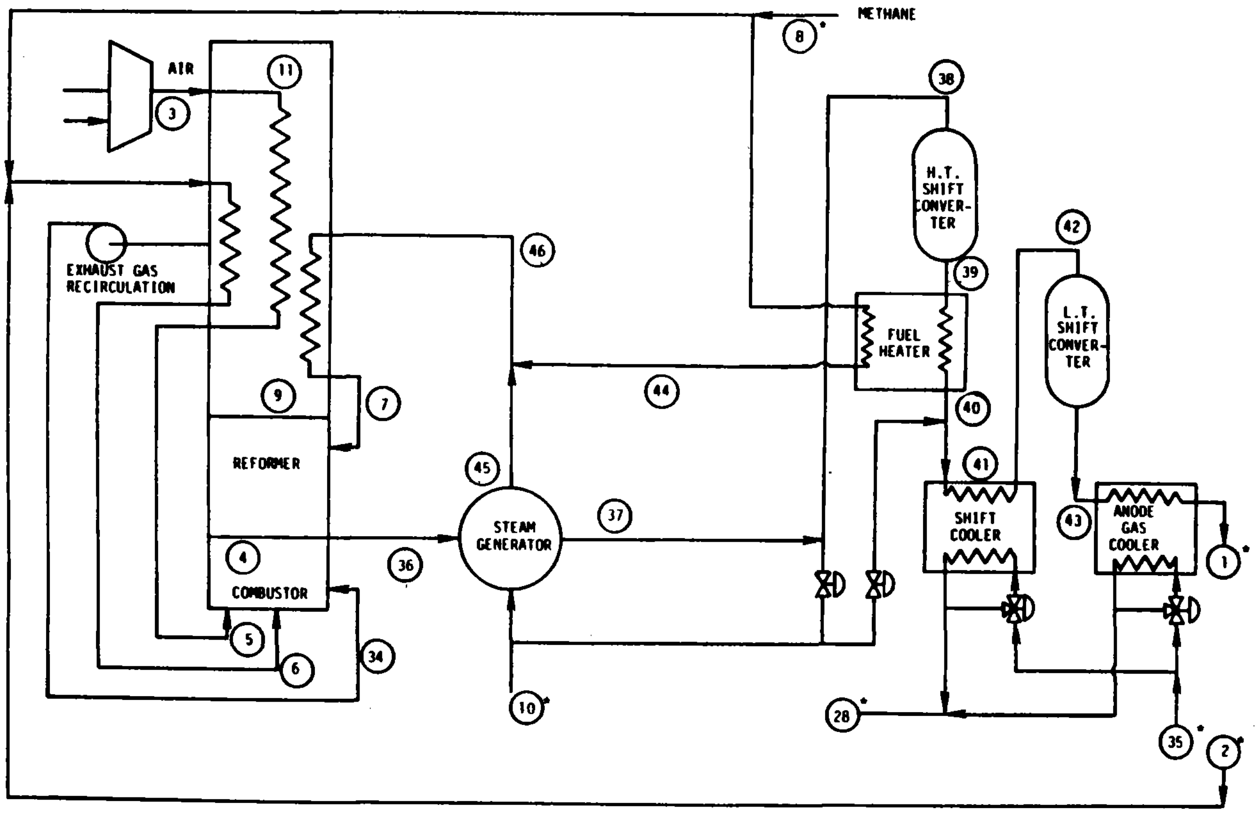
A performance summary for the ethanol fueled nonintegrated conceptual design is listed in Table 5-5.

NATURAL GAS FUELED SYSTEM (METHANE)

Figure 5-1 shows the methanol fuel cell conceptual flow schematic with the methanol fuel processing subsystem encircled with a dashed line. Figure 5-3 shows a conceptual natural gas fuel processing subsystem flow schematic to replace the methanol fuel processing subsystem, and the state points at the design rating which differ from the methanol non integrated system are shown in Table 5-6. About 93 percent of the methane (8) is directed to the fuel heater where the methane is heated to 611°F (4$\overset{8}{4}$) before mixing with steam in the ratio of 1.57 moles of steam per mole of methane. This mixture (46) is then preheated by the combustion gas to 1344°F (7). In the reformer about 95 percent of the methane is reformed with a small amount of shift taking place and exits the reformer at 1520°F (36). The fuel gas is then cooled in the steam generator to 714°F (37). Additional moisture is added via an attemperator before the fuel gas enters the H.T. shift converter (38). The shift reaction is exothermic so that the temperature of the fuel gas increases in the shift converter to 693°F (39). More moisture is added and the gas is cooled in the shift cooler to 330°F (42) before entering the second

TABLE 5-5
PERFORMANCE SUMMARY ETHANOL FUELED NONINTEGRATED

Gross Electrical DC Output (kW)	7500
Gross Electrical AC Output (kW)	7200
Parasitic Losses	
Pumps	23
Fans (Cooling Tower)	210
Vacuum Pump	3
Air Compressor & Dryer	1
Controls	15
Net Electrical AC Output (kW)	6948
Input Energy (hhv = 12776 Btu/lb) Btu/hr	62.93×10^6
Overall Plant Efficiency	0.38
Heat Rate (Btu/kW-hr)	9058
Design Current Density (amp/ft^2)	300
Fuel Utilization Factor	0.8
Design Cell Voltage (volts/cell)	0.66



*Interface Points

Figure 5-3. Natural Gas (Methane) Fueled Nonintegrated Fuel Processing Subsystem

Table 5-6

STATE POINTS NATURAL GAS (METHANE) FUEL PROCESSING SUBSYSTEM

		Mole Fraction							
Station**	Temp. °F	H_2	H_2O	CO	CO_2	O_2	N_2	CH_4	Flowrate lb/hr
1*	350	0.726	0.079	0.010	0.174	-	0.001	0.010	8828
2*	350	0.346	0.189	0.023	0.415	-	0.004	0.023	7897
3	80	-	0.016	-	-	0.204	0.780	-	14625
4	3590	-	0.0284	-	0.208	0.006	0.502	-	22701
5	1300	-	0.016	-	-	0.204	0.780	-	14625
6	1300	0.335	0.183	0.022	0.402	-	0.004	0.054	8076
7	1344	-	0.609	-	-	-	0.003	0.388	6901
8*	80	-	-	-	-	-	0.008	0.992	2695
9	1864	-	0.284	-	0.208	0.006	0.502	-	22701
10*	240	-	1.000	-	-	-	-	-	6311
11	230	-	0.284	-	0.208	0.006	0.502	-	22701
18	96	-	0.032g 0.030ℓ	-	-	0.110	0.828	-	58753
19	96	-	1.000	-	-	-	-	-	6409
28*	175	-	1.000	-	-	-	-	-	30421
35*	135	-	1.000	-	-	-	-	-	30421
36	1520	0.665	0.110	0.185	0.028	-	0.002	0.011	6901
37	714	0.665	0.110	0.185	0.028	-	0.002	0.011	6901
38	476	0.609	0.185	0.169	0.028	-	0.002	0.010	8057
39	693	0.704	0.090	0.074	0.120	-	0.002	0.010	8057
40	544	0.704	0.090	0.074	0.120	-	0.002	0.010	8057
41	407	0.665	0.139	0.070	0.114	-	0.002	0.010	8828
42	330	0.665	0.139	0.070	0.114	-	0.002	0.010	8828
43	462	0.726	0.079	0.010	0.174	-	0.001	0.010	8828
44	611	-	-	-	-	-	0.008	0.992	2517
45	328	-	1.000	-	-	-	-	-	4384
46	440	-	0.609	-	-	-	0.003	0.388	6901

** See Figures 5-1 and 5-3.

shift converter. The fuel gas is then cooled to 350°F (1) before entering the fuel cell stack anode where 80 percent of the hydrogen is utilized. The anodeexhaust gas has a small amount of methane added before being preheated to 1300°F (6). This gas is burned with ten percent excess preheated air (5) to produce a combustion temperature of approximately 3590°F (4). The products of combustion gas is cooled to about 1860°F (9) in the reformer and then to about 240°F (11) by giving up sensible heat for preheating the air, anode exhaust and the methane steam mixture. A recirculation path (34) is provided for control of temperatures in the reformer. The gas composition entering the fuel cell stack anode (1) was estimated based on maps in Reference 1.

The impact of a natural gas fuel processing subsystem versus a methanol fuel processing subsystem is small. The flow rate of water at interface point (10) for the natural gas system is 38 percent greater than that required by the methanol system. This could require more energy for preheating the water provided to the natural gas fuel processing subsystem, but by directing the feedwater first through the anode gas and shift coolers (35) (heats the water from 135°F to 175°F) before introducing the feedwater to the deaerating feedwater heater (28) (Figure 5-3), the energy requirement is reduced by two percent. Since methane reforming requires more water, the separator efficiency must also be increased from 75 percent to 85 percent so that slightly more water will be produced (19) versus that required (10).

The combustion temperature (4) is quite high and may introduce NOX problems. However, the log mean temperature differences and pinch point differences are quite high, so that either more excess air may be used or recirculation of combustion gases can be used to lower the combustion temperatures without significantly increasing the heat exchanger surface area requirements.

A performance summary for the natural gas nonintegrated conceptual design is given in Table 5-7.

COAL GAS FUELED SYSTEM

The product gas from coal gasification systems has a significant concentration of CO. The CO is a poison to the PAFC if the concentration is greater than about one volume percent. Therefore, the product coal gas CO is shifted by adding steam to produce CO_2 and H_2. An oxygen blown entrained flow gasifier product gas was selected as representative of a medium Btu, low methane gas and is used as a basis for this study.

TABLE 5-7
PERFORMANCE SUMMARY NATURAL GAS (METHANE) NONINTEGRATED

Gross Electrical dc Output (kW)	7500
Gross Electrical ac Output (kW)	7200
Parasitic Losses (kW)	
Pumps	26
Fans (Cooling Tower)	210
Vacuum Pump	3
Air Compressor & Dryer	1
Controls	15
Net Electrical ac Output (kW)	6945
Input Energy (10^6 Btu/hr)	64.38
Overall Plant Efficiency	0.368
Heat Rate (Btu/kW-hr)	9270

Figure 5-1 shows the methanol fuel cell conceptual flow schematic with the methanol fuel processing subsystem encircled with a dashed line. Figure 5-4 shows a conceptual nonintegrated coal gas fuel processing subsystem flow schematic to replace the methanol fuel processing subsystem, and the state points which differ from the methanol system are shown in Table 5-8. The methane concentration is low and methane acts as an inert in the fuel cell; reforming of the coal gas is not required. The fuel utilization in the fuel cell is 80 percent which produces an exhaust fuel mixture (2) that has 15.9 volume percent hydrogen. When this is mixed with compressed air (23) at a stoichiometric ratio of 1.1, the resulting mixture will burn without the use of a catalytic combustion chamber. The flame temperature with this mixture is about 1750°F (3). The hot gas is then expanded through a turbine, which drives an ac generator and the air compressor. The turbine exhaust (4) at 1200°F is cooled in the low pressure anode steam generator to 360°F (5). Steam is also generated in the shift cooler and anode gas cooler. These three heat exchangers produce more steam than the amount required for the shift reaction. The excess steam (36) is directed to the steam turbine.

The conceptual nonintegrated coal gas fueled PAFC performance is summarized in Table 5-9. The gross ac output comes from two sources, the fuel cell and the anode side turbocompressor. The net electrical output for this configuration is significantly greater than those PAFC plants that require reforming in the fuel processor. The heat rate is 9110 Btu/kWhr with a dry cooling tower.

NAPHTHA FUELED SYSTEM

Figure 5-1 shows the methanol fuel cell conceptual flow schematic with the methanol fuel processing subsystem encircled with a dashed line. Figure 5-5 shows a conceptual naphtha fuel processing subsystem flow schematic to replace the methanol fuel processing subsystem, and state points at the design rating which differ from the methanol plant are shown in Table 5-10. Octane was used to represent naphtha in this design. About 79 percent of the fuel (11) is directed to the fuel vaporizer and heater where the fuel is preheated to about 600°F before mixing with steam in the ratio of four moles of steam per mole of carbon in the fuel.

This ratio of steam to carbon is required to prevent coking in the reformer because of the high molecular weight and high carbon to hydrogen ratio of the fuel. Before entering the reformer, the fuel and steam mixture is preheated to about 1344°F (9). The operating pressure at the inlet to the reformer is 95 to 100 psia. The reforming reaction takes place at approximately 1500°F, and the

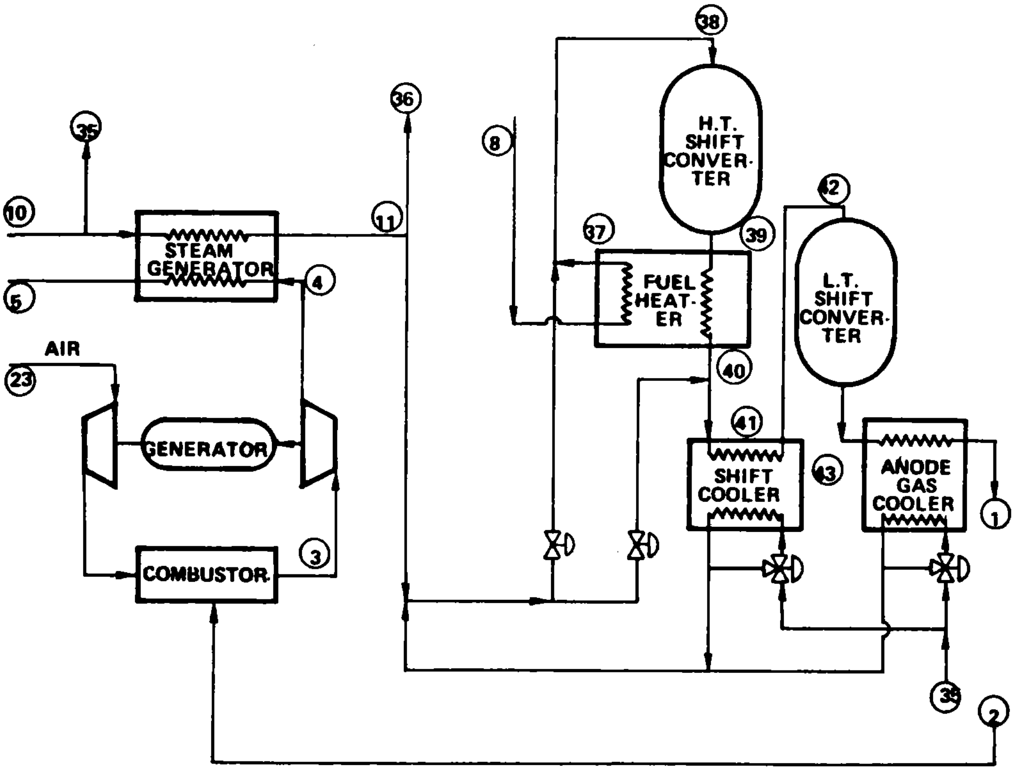


Figure 5-4. Coal Gas Nonintegrated Fuel Processing System

Table 5-8

STATE POINTS COAL GAS FUEL PROCESSING SUBSYSTEM

		Mole Fraction								
Station**	Temp. °F	H_2	H_2O	CO	CO_2	O_2	N_2	H_2S	CH_4	Flowrate lb/hr
1*	350	0.482	0.128	0.01	0.365	-	0.004	0.01	0.003	24320
2*	350	0.156	0.207	0.015	0.593	-	0.006	0.015	0.005	23395
3	1754	-	0.264	-	0.434	0.007	0.295	-	-	34685
4	1200	-	0.264	-	0.434	0.007	0.295	-	-	34685
5	360	-	0.264	-	0.434	0.007	0.295	-	-	34685
8	80	0.39	-	0.376	0.208	-	0.006	0.015	0.005	16487
10*	135	-	1.000	-	-	-	-	-	-	15718
11	280	-	1.000	-	-	-	-	-	-	9986
23	80	-	-	-	-	0.253	0.747	-	-	11290
35	135	-	1.000	-	-	-	-	-	-	5732
36	280	-	1.000	-	-	-	-	-	-	4198
37	600	0.39	-	0.376	0.208	-	0.006	0.015	0.005	16487
38	521	0.305	0.219	0.294	0.162	-	0.005	0.012	0.004	20357
39	820	0.44	0.078	0.15	0.303	-	0.005	0.012	0.004	20357
40	452	0.44	0.078	0.15	0.303	-	0.005	0.012	0.004	20357
41	436	0.365	0.244	0.125	0.249	-	0.004	0.010	0.003	24320
42	350	0.365	0.244	0.125	0.249	-	0.004	0.010	0.003	24320
43	587	0.482	0.128	0.010	0.365	-	0.004	0.010	0.003	24320
44	280	-	1.000	-	-	-	-	-	-	11520
45	280	-	1.000	-	-	-	-	-	-	5732

** See Figures 5-1 and 5-4

TABLE 5-9

PERFORMANCE SUMMARY COAL GAS FUELED NONINTEGRATED

Gross F.C. dc Electrical Output (kW)	7500
Gross F.C. ac Electrical Output (kW)	7200
Turbogenerator ac Power (kW)	1178
Gross ac Power (kW)	8378
Parasitic Losses (kW)	
Pumps	23
Fans (Cooling Tower)	210
Vacuum Pump	3
Air Compressor & Dryer	1
Controls	15
Net ac Output (kW)	8126
Input Energy (HHV) Btu/hr	$74.03\ 10^6$
Overall Plant Efficiency	0.375
Heat Rate (Btu/kW-hr)	9110

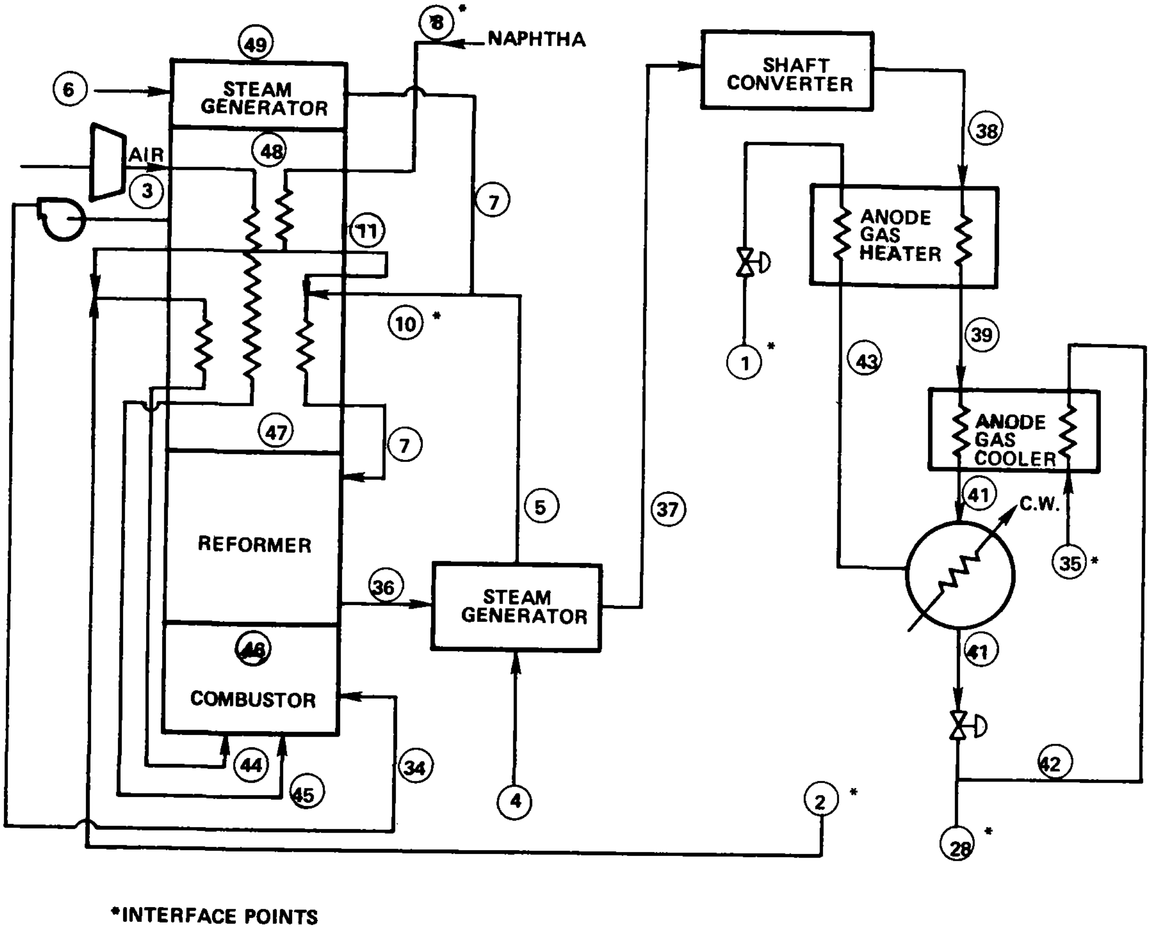


Figure 5-5. Naphtha Fueled Nonintegrated Fuel Processing Subsystem

Table 5-10

STATE POINTS NAPHTHA FUEL PROCESSING SUBSYSTEM

Station**	Temp. °F	Pressure psia	Mole Fraction								Flowrate lb/hr
			H_2	CH_4	CO	CO_2	H_2O	O_2	N_2	Fuel	
1*	350	50	0.708	0.010	0.007	0.231	0.044	-	-	-	10463
2*	350	50	0.326	0.024	0.017	0.531	0.102	-	-	-	9525
3	80	18	-	-	-			0.21	0.79	-	24658
4	240	105	-	-	-	-	1.000	-	-		8797
5	328	100	-	-	-	-	1.000	-	-	-	8797
6	240	105	-	-	-	-	1.000	-	-	-	5582
7	328	100	-	-	-	-	1.000	-	-	-	5582
8*	80	105	-	-	-	-	-	-	-	1.000	3622
9	1344	98	-	-	-	-	0.97	-	-	0.03	17231
10*	328	100	-	-	-	-	1.000	-	-	-	14382
11	400	100	-	-	-	-	-	-	-	1.000	2849
35	135	32	-	-	-	-	1.000	-	-	-	31695
36	1520	90	0.414	0.007	0.077	0.087	0.415	-	-	-	17227
37	350	85	0.414	0.007	0.077	0.087	0.415	-	-	-	17227
38	500	80	0.486	0.007	0.005	0.158	0.344	-	-	-	17227
39	310	78	0.486	0.007	0.005	0.158	0.344	-	-	-	17227
40	160	76	0.486	0.007	0.005	0.158	0.344	-	-	-	17227
41	141	70	-	-	-	-	1.000	-	-	-	6764
42	216	30	-	-	-	-	1.000	-	-	-	31725
43	141	74	0.708	0.010	0.007	0.231	0.044	-	-	-	10463
44	1300	48	0.320	0.024	0.016	0.522	0.100	-	-	0.018	10298
45	1300	16	-	-	-	-	-	0.21	0.79	-	24658
46	3845	15	-	-	-	0.219	0.195	0.014	0.572	-	34956
47	2768	15	-	-	-	0.219	0.195	0.014	0.572	-	34956
48	814	15	-	-	-	0.219	0.195	0.014	0.572	-	34956
49	267	15	-	-	-	0.219	0.195	0.014	0.572	-	34956

* Interface Points
** See Figures 5-1 and 5-5

reformed products (36) are cooled to 350°F (37), by generating part of the steam (5) required for reforming. The reformed gas then enters the shift converter where most of the CO content is reacted with H_2O to form more hydrogen. The CO-shift reaction operates at 80 to 85 psia and is moderately exothermic. The reaction heat is employed to preheat the makeup water to 216°F and to preheat the fuel gas to 350°F (1) before it enters the fuel cell stack anode where 80 percent of the hydrogen is utilized. A final processing of the exhaust gas from CO-shift converter removes the water from the fuel gas to a saturation temperature of 141°F in a recuperator/condenser unit. The anode exhaust gas is added to the vaporized naphtha fuel before being pre heated to 1300°F (44). The gas is burned with ten percent excess preheated air (45) to produce a combustion temperature of approximately 3845°F (46). The products of combustion gas is cooled to about 2768°F in the reformer and then to about 814°F (48) by giving up sensible heat for vaporizing the fuel, preheating the air, preheating the anode exhaust, and preheating the fuel and steam mixture. The products of combustion gas are finally cooled to 267°F by giving up sensible heat in a steam generator to produce part of the steam required in the reformer.

The flow rate of water at interface point (10) for the naphtha system is 214 percent greater than that required by the methanol system. The additional energy needed for preheating the water used in the naphtha fuel processing subsystem is provided by directing the feedwater first through the anode gas cooler (heats the water from 135°F to 216°F) before introducing the feedwater to the deaerating feedwater heater (28).

The combustion temperature (46) is quite high and may introduce an NO_x problem. However, the pinch point differences are quite high, so that recirculation of combustion gases (34) can be used to lower the combustion temperature without significantly increasing the heat exchanger surface area.

A performance summary for the naphtha nonintegrated conceptual design is shown in Table 5-11.

The performance of all the nonintegrated systems is summarized in Table 5-12.

5.1.2 INTEGRATED SYSTEMS

One of the design features of the nonintegrated designs was that the fuel processing system (FPS) effectively operated independent of the fuel cell system. In the

TABLE 5-11

PERFORMANCE SUMMARY NAPHTHA FUELED NONINTEGRATED

Gross Electrical DC Output (kW)	7500
Gross Electrical AC Output (kW)	7200
Parasitic Losses (kW)	
Pumps	27
Fans (Cooling Tower)	210
Vacuum Pump	3
Air Compressor and Dryer	1
Controls	15
Net Electrical AC Output (kW)	6944
Input Energy (hhv = 20593 Btu/lb) Btu/hr	74.57×10^6
Overall Plant Efficiency	0.318
Heat Rate (Btu/kW-hr)	10739

Table 5-12

PERFORMANCE COMPARISON NONINTEGRATED SYSTEMS

Fuel	Methanol	Ethanol	Natural Gas	Coal Gas	Naphtha
Gross electrical dc output (kW)	7500	7500	7500	7500	7500
Gross electrical ac output (kW)	7200	7200	7200	8378	7200
Parasitic Losses					
Pumps	23	23	26	23	27
Fans (cooling tower)	210	210	210	210	210
Vacuum pump	3	3	3	3	3
Air compressor	1	1	1	1	1
Controls	15	15	15	15	15
Rotating group motor	0	0	0	0	0
Net electrical ac output (kW)	6948	6948	6945	8126	6944
High heating value (Btu/lb)	9758	12776	23890	4490	20593
Input energy (10^6 Btu/hr)	61.08	62.93	64.38	74.03	74.57
Overall plant efficiency	0.39	0.377	0.368	0.375	0.318
Heat rate (Btu/kW-hr)	8791	9058	9270	9110	10739
Fuel utilization	0.8	0.8	0.8	0.8	0.8

nonintegrated concepts, the hydrogen utilization in the fuel cell is relatively low (0.8) in order to provide anode exhaust gas with sufficient hydrogen to fire the reformer, the steam generator for the reformer, and preheat and/or vaporize the fuel. This results in relatively high heat rates for these concepts.

Lower heat rates can be attained by utilizing more of the hydrogen in the fuel cell and using the reject heat from the fuel cell to raise steam for the reformer and to vaporize/preheat the fuel. However, high hydrogen utilization in the fuel cell results in low hydrogen concentrations in the anode exhaust which must be mixed with excess air to reduce the concentration of inert gases and allow efficient burning. The use of fuel cell reject heat to supply some of the thermal needs of the fuel processor results in a more complex and compact interface between the fuel processing and fuel cell power plant system.

A performance comparison for each integrated system is shown in Table 5-23.

METHANOL FUELED SYSTEM

A schematic of a methanol fueled integrated plant arrangement is shown in Figure 5-6, and the state points at design rating are shown in Table 5-13. In this configuration, the fuel conditioner combustor is pressurized to fuel cell pressure level, and the pressurized exhaust is combined with that from the cathode of the fuel cell and then passed through the expander. Since the steam available for the steam turbine is reduced in this configuration, as compared to the nonintegrated configuration, an electric motor must be added to the rotating group to balance the power requirements. The performance summary for this concept is given in Table 5-14.

The relatively low catalyst reforming temperature requirements for high grade methanol (Figure 5-7) suggest the possibility of supplying a significant portion of the reforming energy by means of reject heat from the fuel cell cathode recirculation stream. A detriment to this concept is the requirement of a carbon dioxide removal system to treat the exhaust fuel (anode exit) stream due to vastly reduced combustion thermal energy requirements. CO_2 removal systems are available, simple in concept, and reliable in operation for the percent CO_2 removal acceptable.

A concept schematic for a methanol fueled integrated power plant with a CO_2 removal system is shown in Figure 5-8, and the design rating state points are

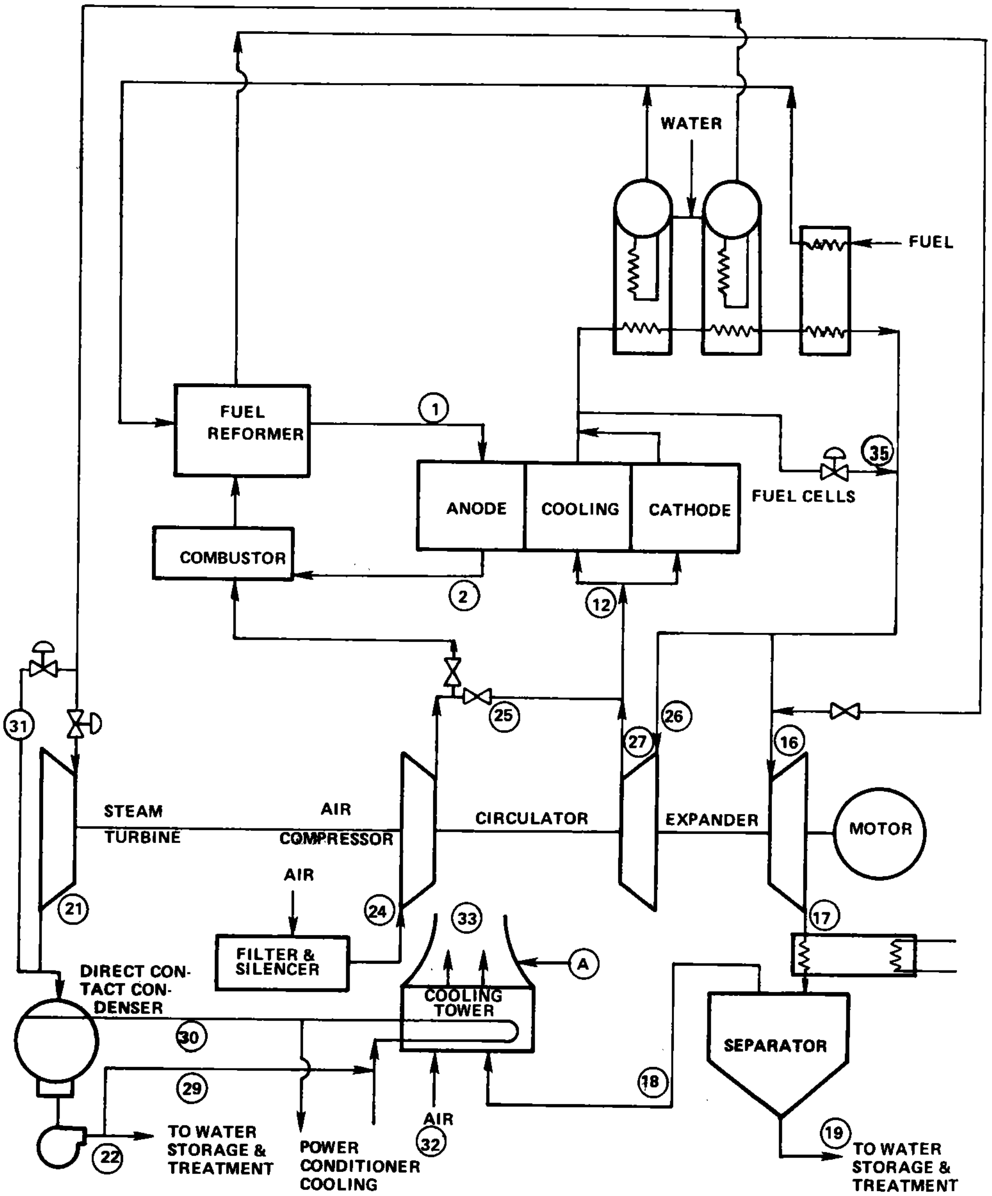


Figure 5-6. Methanol Fueled Integrated Conceptual Fuel Cell Flow Schematic

Table 5-13

STATE POINTS METHANOL FUELED INTEGRATED SYSTEM

STATION**	Temp. °F	Pressure psia	Mole Fraction							Flowrate lb/hr
			H_2	H_2O	CO	CO_2	O_2	N_2	CH_3OH	
1	350	50	0.693	0.075	0.005	0.227	-	-	-	9612
2	350	50	0.184	0.198	0.014	0.604	-	-	-	8682
3	348	50	-	0.010	-	-	0.208	0.782	-	4282
4	1790	50	-	0.273	-	0.431	0.007	0.289	-	12964
5	782	50	-	0.273	-	0.431	0.007	0.289	-	12964
8	80	80	-	-	-	-	-	-	1.000	5552
9	300	75	-	-	-	-	-	-	1.000	5552
10	340	100	-	1.000	-	-	-	-	-	4060
11	317	89	-	0.565	-	-	-	-	0.435	9612
12	277	50	-	0.187	-	-	0.100	0.713	-	1103350
13	352	50	-	0.197	-	-	0.094	0.709	-	1104190
14	269	49	-	0.197	-	-	0.094	0.709	-	1104190
16	355	49	-	0.208	-	0.060	0.082	0.650	-	78124
17	151	15	-	0.208	-	0.060	0.082	0.650	-	78124
18	96	15	-	0.056g 0.004	-	0.072	0.097	0.771	-	69987
19	96	15	-	1.000	-	-	-	-	-	8137
20	249	30	-	1.000	-	-	-	-	-	17632
21	141	3	-	1.000	-	-	-	-	-	17632
22	135	3	-	1.000	-	-	-	-	-	17632
24	80	14.6	-	0.010	-	-	0.208	0.782	-	68602
25	348	50	-	0.010	-	-	0.208	0.782	-	64320
26	269	49	-	0.197	-	-	0.094	0.709	-	1039030
27	273	50	-	0.197	-	-	0.094	0.709	-	1039030
28	135	30	-	1.000	-	-	-	-	-	22212
29	135	3	-	1.000	-	-	-	-	-	574215
30	105	15	-	1.000	-	-	-	-	-	574215
32	80	15	-	0.010	-	-	0.208	0.782	-	3.559×10^6
33	100	15	-	0.010	-	-	0.208	0.782	-	3.634×10^6

** See Figure 5-6

Table 5-14

PERFORMANCE SUMMARY METHANOL FUELED INTEGRATED

Gross Electrical dc Output (kW)	7500
Gross Electrical ac Output (kW)	7200
Parasitic Losses	
Pumps	22
Fans (Cooling Tower)	150
Vacuum Pump	3
Air Compressor and Dryer	1
Controls	15
Rotating Group Motor	93
Net Electrical ac Output (kW)	6916
Input Energy (hhv = 9758 Btu/lb) Btu/hr	54.1×10^6
Overall Plant Efficiency	0.44
Heat Rate (Btu/kW-hr)	7822
Fuel Utilization Factor	0.9

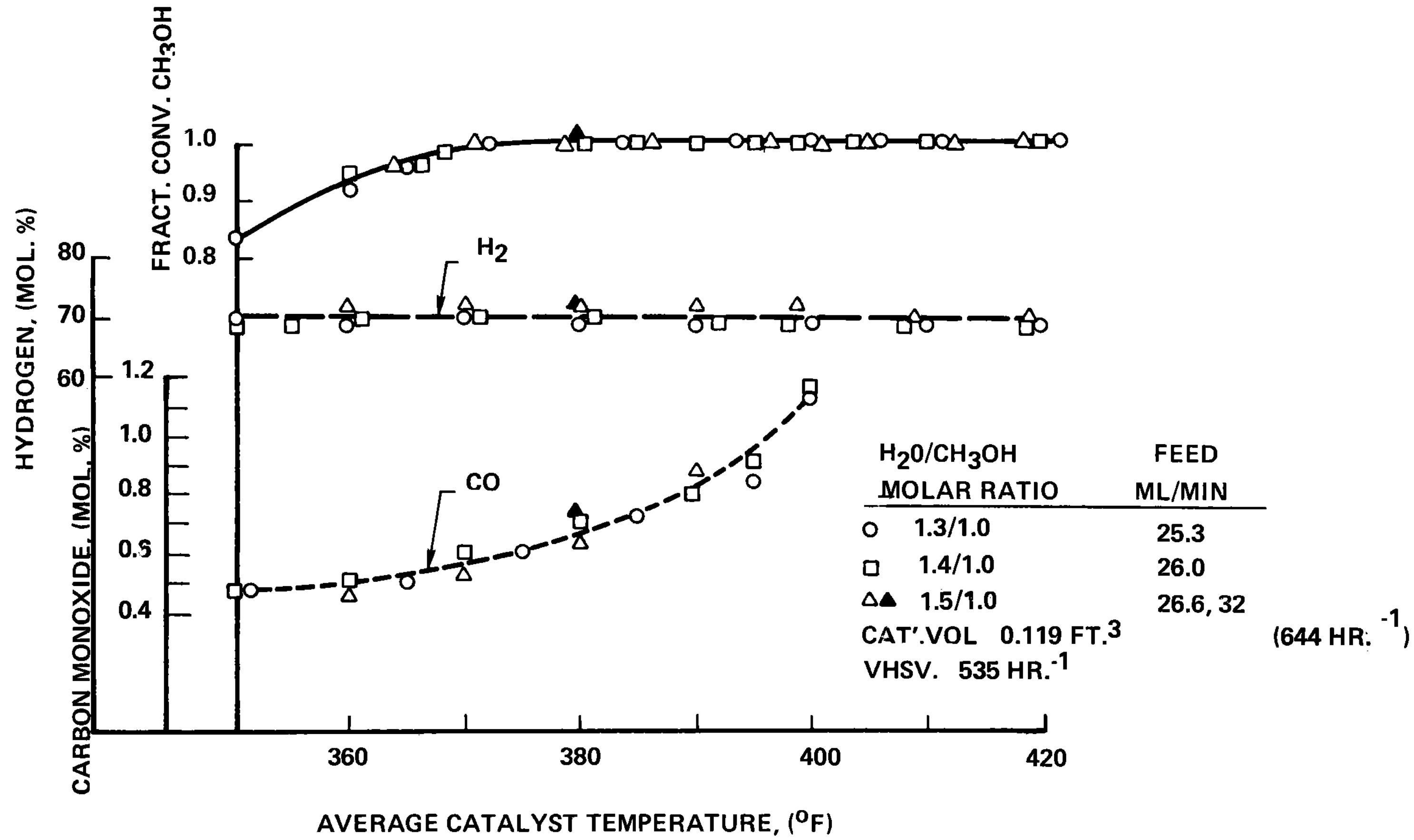


Figure 5-7. Methanol Fuel Conditioner Performance

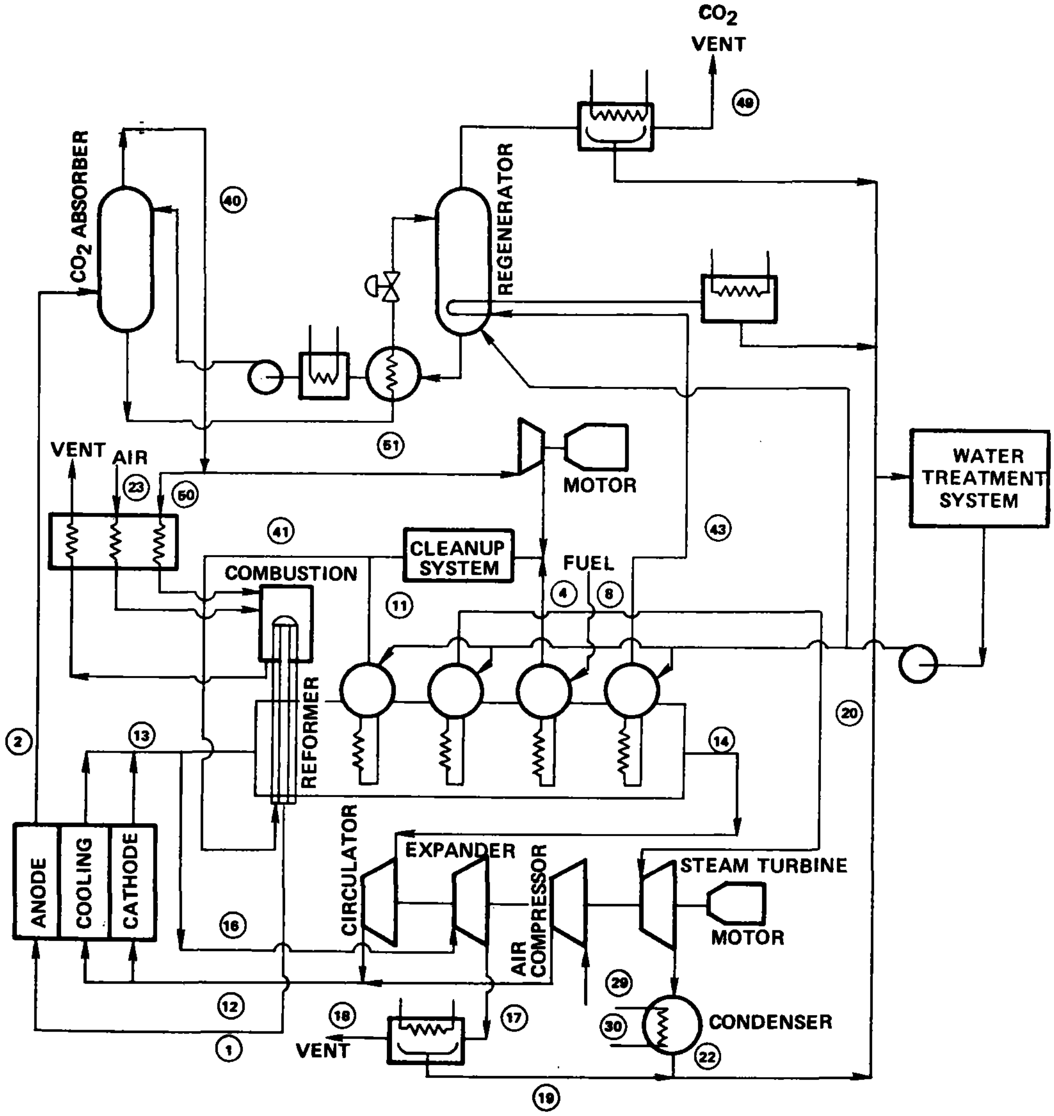


Figure 5-8. Methanol Fueled Integrated Conceptual Fuel Cell Flow Schematic With CO_2 Removal System

listed in Table 5-15. The methanol in liquid form is pumped through the fuel vaporizer, combined with hydrogen and water vapor returning from the CO_2 removal system through the cleanup system, and then mixed with steam. The source of heat for fuel and water vaporization is from the recirculation loop of the fuel cell. The introduction of the exhaust fuel stream (with CO_2 removed) back into the main fuel stream before admission to the reformer reduces the reformer steam generation requirement. The mixture created when these two streams are combined is then introduced into the low temperature reformer where it is heated to 350°F and 80 percent of the methanol is converted. The partially reformed gas is then recuperated and heated to 600°F in the high temperature reformer which is heated by combustion gases. The reformed fuel stream is then recuperatively cooled and introduced into the fuel cell anode.

The anode exhaust is passed counterflow to a circulating stream of aqueous hot potassium carbonate in a packed column. This operation removes the CO_2 by absorption into the liquid solution. The solution from the bottom of the absorber is sent to the top of a packed regenerator column and passes counter current to a stripping stream at low pressure. The stripping process results in CO_2 and H_2O (vapor) leaving the top of the regenerator and entering the condenser where a significant portion of the H_2O is removed for reuse. The gas stream leaving the top absorber has a CO_2 content of about two volume percent. A small portion of this stream is utilized for providing reforming thermal energy. Most of the gas, however, is returned for recycling through the fuel cell.

This configuration has some desirable features. For example, control of the fuel utilization is not critical, because the hydrogen that is not used in the fuel cell is recycled. The temperature used in the high temperature reformer can be adjusted for the quality of the fuel. A performance summary for this system is given in Table 5-16.

NATURAL GAS FUELED SYSTEM

A schematic of a natural gas fueled integrated plant arrangement is shown in Figure 5-9, and the design rating state points are listed in Table 5-17. In this configuration, reformer steam and turbine steam are both produced by reject heat in the cathode recirculation loop. Since the steam available for the steam turbine is reduced in this configuration, as compared to the nonintegrated configuration, additional rotative power is obtained by expanding cathode

Table 5-15

STATE POINTS METHANOL FUELED INTEGRATED SYSTEM WITH CO_2 REMOVAL

			Mole Fraction							
Station**	Temp °F	Pressure psia	H_2	H_2O	CO	CO_2	O_2	N_2	CH_3OH	Flowrate lb/hr
1	350	50	0.710	0.066	0.005	0.219	-	-	-	9031
2	350	50	0.197	0.183	0.014	0.606	-	-	-	8099
8	80	80	-	-	-	-	-	-	1.000	5089
9	250	75	-	-	-	-	-	-	1.000	5089
11	312	81	-	1.000	-	-	-	-	-	3024
12	277	50	-	0.187	-	-	0.100	0.713	-	1103350
13	352	50	-	0.197	-	-	0.094	0.709	-	1104190
14	269	49	-	0.197	-	-	0.094	0.709	-	1039030
16	352	50	-	0.197	-	-	0.094	0.709	-	65160
17	148	15	-	0.197	-	-	0.094	0.709	-	65160
18	96	15	-	0.054g 0.042ℓ	-	-	0.106	0.798	-	60190
19	96	15	-	1.000	-	-	-	-	-	4970
20	307	75	-	1.000	-	-	-	-	-	9064
22	135	3	-	1.000	-	-	-	-	-	9064
23	80	15	-	0.010	-	-	0.208	0.782	-	919
29	135	5	-	1.000	-	-	-	-	-	301853
30	105	15	-	1.000	-	-	-	-	-	301853
40	250	45	0.491	0.459	0.035	0.015	-	-	-	1148
41	275	70	0.101	0.503	0.007	0.003	-	-	0.387	9031
43	249	30	-	1.000	-	-	-	-	-	7124
49	135	15	-	0.173	-	0.827	-	-	-	7547
50	250	45	0.491	0.459	0.035	0.015	-	-	-	230
51	250	45	0.491	0.459	0.035	0.015	-	-	-	918

** See Figure 5-8

Table 5-16

PEROFRMANCE SUMMARY METHANOL FUELED INTEGRATED SYSTEM WITH CO_2 REMOVAL

Gross Electrical dc Output (kW)	7500
Gross Electrical ac Output (kW)	7200
Parasitic Losses	
Pumps	23
Fans (Cooling Tower)	135
Vacuum Pump	3
Air Compressor and Dryer	1
Controls	15
Rotating Group Motor	318
Net Electrical ac Output (kW)	6705
Input Energy (hhv = 9758 Btu/lb) Btu/hr	49.51×10^6
Overall Plant Efficiency	0.46
Heat Rate (Btu/kW-hr)	7384

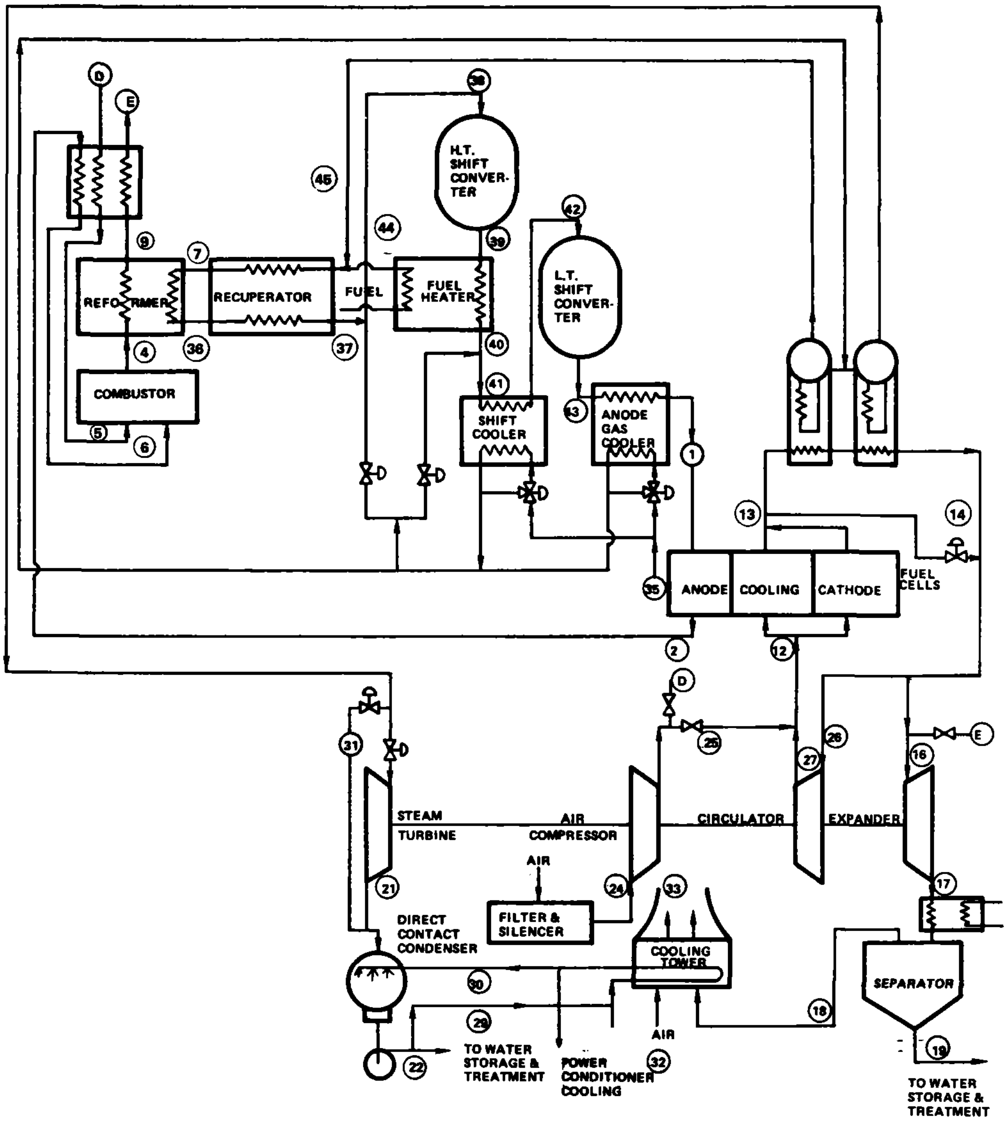


Figure 5-9. Natural Gas Fueled Integrated Conceptual Fuel Cell Flow Schematic

Table 5-17

STATE POINTS NATURAL GAS (METHANE) INTEGRATED SYSTEM

Station**	Temp. °F	Mole Fraction							Flowrate lb/hr
		H_2	H_2O	CO	CO_2	O_2	N_2	CH_4	
1	350	0.726	0.079	0.010	0.174	-	0.001	0.010	8828
2	350	0.346	0.189	0.023	0.415	-	0.004	0.023	7897
4	3500	-	0.292	-	0.227	0.011	0.470	-	19683
5	1300	-	0.010	-	-	0.208	0.782	-	11786
6	1300	0.346	0.189	0.023	0.415	-	0.004	0.023	7897
7	1344	-	0.609	-	-	-	0.003	0.388	6901
9	1392	-	0.292	-	0.227	0.011	0.470	-	19683
12	277	-	0.187	-	-	0.100	0.713	-	1103350
13	352	-	0.197	-	-	0.094	0.709	-	1104190
14	269	-	0.197	-	-	0.094	0.709	-	1104190
16	294	-	0.218	-	0.049	0.076	0.657	-	84842
17	140	-	0.187g 0.031ℓ	-	0.049	0.076	0.657	-	84842
18	96	-	0.056g 0.004ℓ	-	0.059	0.091	0.790	-	75329
19	96	-	1.000	-	-	-	-	-	9513
20	249	-	1.000	-	-	-	-	-	20013
21	141	-	1.000	-	-	-	-	-	20013
22	135	-	1.000	-	-	-	-	-	20013
24	80	-	0.010	-	-	0.208	0.782	-	73747
25	348	-	0.010	-	-	0.208	0.782	-	64320
26	269	-	0.197	-	-	0.094	0.709	-	1039030
27	273	-	0.197	-	-	0.094	0.709	-	1039030
29	135	-	1.000	-	-	-	-	-	651757
30	105	-	1.000	-	-	-	-	-	651757
32	80	-	0.010	-	-	0.208	0.782	-	4.04×10^6
33	100	-	0.010	-	-	0.208	0.782	-	4.04×10^6
35	135	-	1.000	-	-	-	-	-	25168
36	1520	0.665	0.110	0.185	0.028	-	0.002	0.011	6901
37	714	0.665	0.110	0.185	0.028	-	0.002	0.011	6901
38	476	0.609	0.185	0.169	0.028	-	0.002	0.010	8057
39	693	0.704	0.090	0.074	0.120	-	0.002	0.010	8057
40	544	0.704	0.090	0.074	0.120	-	0.002	0.010	8057
41	407	0.665	0.139	0.070	0.114	-	0.002	0.010	8828
42	330	0.665	0.139	0.070	0.114	-	0.002	0.010	8828
43	462	0.726	0.079	0.010	0.174	-	0.001	0.010	8828
44	611	-	-	-	-	-	0.008	0.992	2517
45	328	-	1.000	-	-	-	-	-	4384

** See Figure 5-10

gas that was previously combined with pressurized fuel conditioner combustion gases. The performance summary for this concept is given in Table 5-18.

COAL GAS FUELED SYSTEM

A conceptual schematic of a coal gas fueled integrated plant is shown in Figure 5-10, and the design rating state points are given in Table 5-19. Because coal gas has a low concentration of methane, the hydrogen containing anode exhaust must be used effectively to obtain high PAFC plant efficiencies. The most effective way to use the hydrogen is in the fuel cell. The anode exhaust gas can be passed through gas purification systems where impurities such as CO, CO_2, CH_4, and N_2 are absorbed and hydrogen with purity of over 99.9 percent can be obtained. However, such a high degree of hydrogen purity is not required for the fuel cell, so a simple Benfield process is used as representative.

The CO_2 removal system allows all of the hydrogen produced to be consumed in the fuel cell. This system also reduces the steam requirements for the shift reaction, since the stream returning from the CO_2 removal system has high concentrations of both hydrogen and water vapor.

A performance summary for the coal gas integrated case is given in Table 5-20.

NAPHTHA FUELED SYSTEM

A schematic of a Naphtha fueled integrated plant arrangement is shown in Figure 5-11, and the state points of the design rating are listed in Table 5-21. The hydrogen rich gas supplied to the fuel cell is produced by steam reforming Naphtha fuel. This fuel was represented in the fuel processing calculations by octane. The fuel passes through a vaporizer and desulfurizer with a small bleed stream of hydrogen added to it for desulfurizing. This bleed stream is now shown. The fuel stream is preheated to about 600°F and superheated steam is added to the fuel stream in a molar ratio of 4:1 steam-to-carbon of the fuel. The steam-fuel mixture is preheated to about 1100°F, after which it passes into the reformer. The operating pressure at the inlet to the reformer is 90 to 100 psia. The reforming reaction occurs at approximately 1500°F, and the reformer products are cooled to about 800°F, by preheating the reactant stream, and second, by vaporizing (and desulfurizing) the fuel. The reformed gas enters the shift converter where the CO content is reacted with H_2O to form more hydrogen.

Table 5-18

PERFORMANCE SUMMARY NATURAL GAS FUELED INTEGRATED

Gross Electrical dc Output (kW)	7500
Gross Electrical ac Output (kW)	7200
Parasitic Losses	
Pumps	26
Fans (Cooling Tower)	200
Vacuum Pump	3
Air Compressor and Dryer	1
Controls	15
Net Electrical ac Output (kW)	6955
Input Energy (hhv = 23890 Btu/lb) Btu/hr	58.56×10^6
Overall Plant Efficiency	0.40
Heat Rate (Btu/kW-hr)	8438

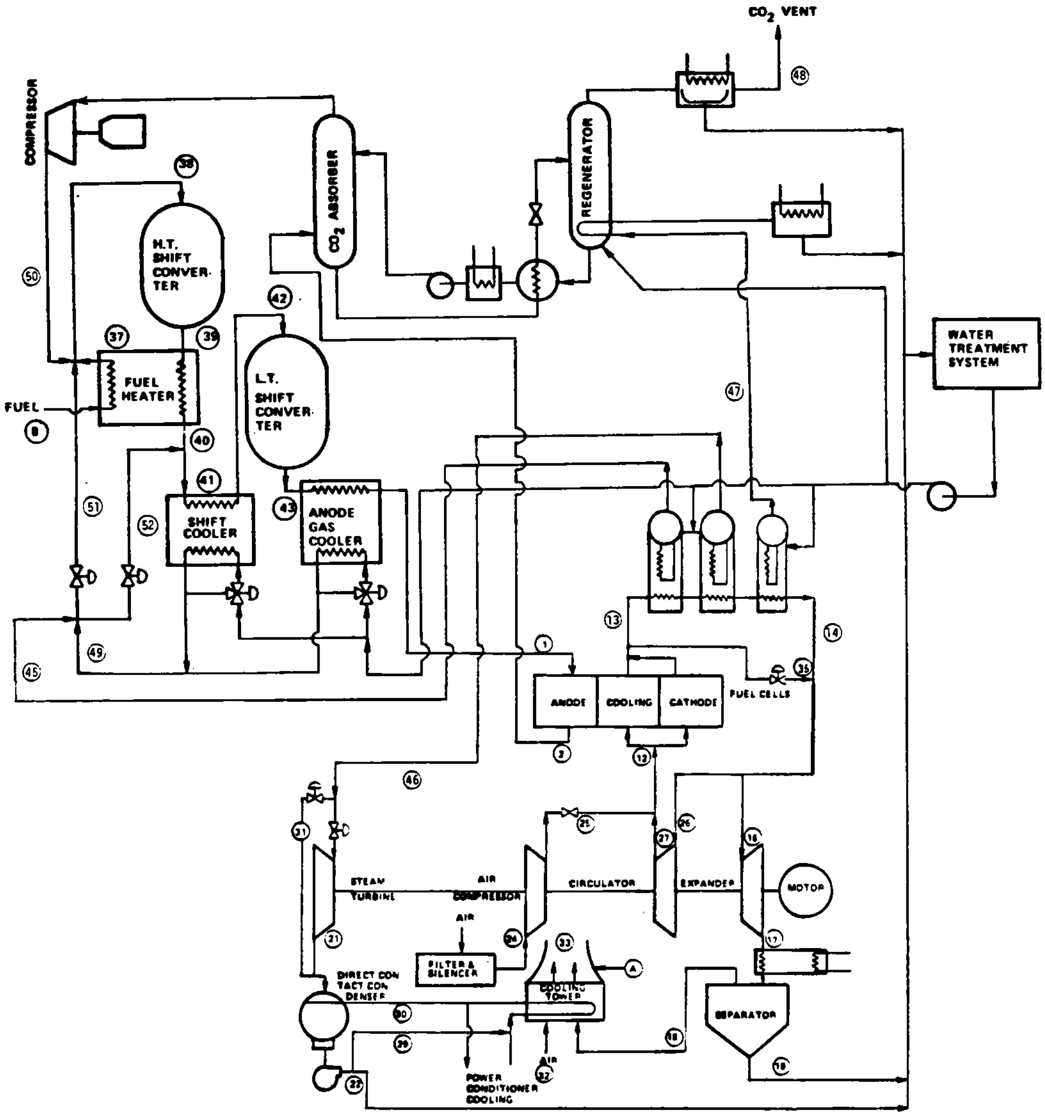


Figure 5-10. Coal Gas Integrated Conceptual Fuel Cell Schematic

Table 5-19

STATE POINTS COAL GAS INTEGRATED SYSTEM

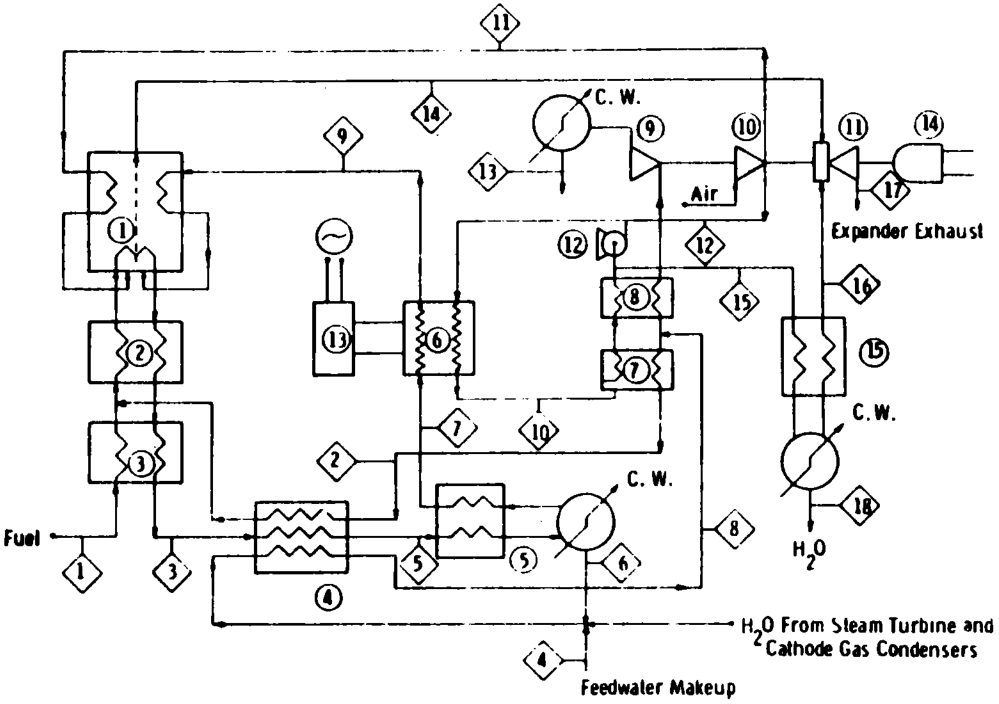
Station**	Temp. °F	Pressure psia	Mole Fraction							Flowrate lb/hr
			H_2	H_2O	CO	CO_2	O_2	N_2	CH_4	
1	350	50	0.540	0.114	0.010	0.330	-	0.003	0.003	19513
2	350	50	0.190	0.200	0.018	0.581	-	0.006	0.005	18581
8	80	80	0.390	-	0.376	0.208	-	0.006	0.005 CH_4 0.015 H_2S	13212
12	277	50	-	0.187	-	-	0.100	0.713	-	1,103,350
13	352	50	-	0.197	-	-	0.094	0.709	-	1,104,190
14	269	49	-	0.197	-	-	0.094	0.709	-	1,104,190
16	269	49	-	0.197	-	-	0.094	0.709	-	65160
17	138	15	-	0.175g 0.022ℓ	-	-	0.094	0.709	-	65160
18	96	15	-	0.054g 0.042ℓ	-	-	0.106	0.798	-	59500
19	96	15	-	1.000	-	-	-	-	-	5660
21	141	3	-	1.000	-	-	-	-	-	4934
22	135	3	-	1.000	-	-	-	-	-	4934
24	80	15	-	0.010	-	-	0.208	0.782	-	64320
25	348	50	-	0.010	-	-	0.208	0.782	-	64320
26	269	49	-	0.197	-	-	0.094	0.709	-	1,039,030
27	273	50	-	0.197	-	-	0.094	0.709	-	1,039,030
29	135	3	-	1.000	-	-	-	-	-	167,685
30	105	15	-	1.000	-	-	-	-	-	167,685
32	80	15	-	0.010	-	-	0.208	0.782	-	1.0394 x 10^6
33	100	15	-	0.010	-	-	0.208	0.782	-	1.0394 x 10^6
37	600	105	0.396	-	0.382	0.211	-	0.006	0.005	12894
38	483	85	0.371	0.242	0.242	0.138	-	0.004	0.003	17504
39	725	85	0.492	0.122	0.121	0.258	-	0.004	0.003	17504
40	419	84	0.492	0.122	0.121	0.258	-	0.004	0.003	17504
41	390	83	0.441	0.212	0.108	0.232	-	0.004	0.003	19513
42	350	82	0.441	0.212	0.108	0.232	-	0.004	0.003	19513
43	558	81	0.540	0.114	0.010	0.330	-	0.003	0.003	19513
45	280	49	-	1.000	-	-	-	-	-	1998
46	249	30	-	1.000	-	-	-	-	-	4934
47	249	30	-	1.000	-	-	-	-	-	17121
48	135	15	-	0.173	-	0.827	-	-	-	16849
49	280	49	-	1.000	-	-	-	-	-	2020
50	250	45	0.479	0.506	-	0.015	-	-	-	2603
51	280	49	-	1.000	-	-	-	-	-	2009
52	280	49	-	1.000	-	-	-	-	-	2009

** See Figure 5-10

Table 5-20

PERFORMANCE SUMMARY COAL GAS INTEGRATED SYSTEM

Gross Electrical dc Output (kW)	7500
Gross Electrical ac Output (kW)	7200
Parasitic Losses	
Pumps	23
Fans (Cooling Tower)	205
Vacuum Pump	3
Air Compressor and Dryer	1
Controls	15
Net Electrical ac Output (kW)	6953
Input Energy (hhv = 4490 Btu/lb) Btu/hr	59.22×10^6
Overall Plant Efficiency	0.40
Heat Rate (Btu/kW-hr)	8520



Legend ○ Equipment Item ◇ Stream Number

1. Reformer
    - i. Reform CH_{2+}
    - ii. Preheat Anode Gas
    - iii. Preheat Combustion Air
2. Reformer Reactant/ Product Heat Exchanger
3. Desulfurizer & Fuel Vaporizer
4. Co-Shift
    - i. Shift Reform Gas
    - ii. Superheat Reformer Steam
    - iii. Heat Feedwater to Steam Generators
5. Moisture Removal From Fuel Gas
6. Fuel Cell
    - i. Anode Gas - Fuel
    - ii. Process Air - Oxidant
    - iii. Recirculating Air - Coolant
7. Reformer Steam Generator
8. Auxiliary Steam - Turbine Steam Generator
9. Steam Turbine and Condenser
10. Compressor
    - I. Fuel-Cell Air
    - ii. Reformer Combustion Air
11. Expander
    - i. Fuel-Cell Exhaust Air
    - ii. Reformer Combustion Exhaust
12. Recirculator & Motor
13. DC/ AC Inverter
14. Motor
15. Moisture Removal from Cathode Gas

Figure 5-11. Naphtha Fueled Integrated System

Table 5-21

STATE POINTS NAPHTHA FUELED INTEGRATED SYSTEM

			Mole Fraction								
Station**	Temp. °F	Pressure psia	H_2	H_2O	CO	CO_2	O_2	N_2	CH_4	Fuel	Flowrate lb/hr
1	77	100	-	-	-	-	-	-	-	1.000	2936
2	328	100	-	1.000	-	-	-	-	-	-	14831
3	834	85	0.418	0.417	0.072	0.086	-	-	0.007	-	17767
4	77	105	-	1.000	-	-	-	-	-	-	921
5	500	80	0.488	0.347	0.001	0.157	-	-	0.007	-	17767
6	141	70	-	1.000	-	-	-	-	-	-	7136
7	375	70	0.716	0.042	0.002	0.229	-	-	0.011	-	10631
8	254	100	-	1.000	-	-	-	-	-	-	26791
9	375	50	0.341	0.098	0.005	0.531	-	-	0.025	-	9663
10	362	50	-	0.216	-	-	0.083	0.701	-	-	2.21×10^6
11	341	50	-	0.010	-	-	0.208	0.782	-	-	12316
12	341	50	-	0.010	-	-	0.208	0.782	-	-	59800
13	341	3	-	1.000	-	-	-	-	-	-	11960
14	434	42.5	-	0.250	-	0.281	0.011	0.458	-	-	21986
15	322	50	-	0.216	-	-	0.083	0.701	-	-	60770
16	280	42.5	-	0.064	-	-	0.099	0.837	-	-	53996
17	171	15	-	0.115	-	0.077	0.075	0.733	-	-	75983
18	141	46.25	-	1.000	-	-	-	-	-	-	6774

** See Figure 5-12

The reformer reaction is strongly endothermic. The heat addition to the reformer is supplied from combustion of the anode vent gas. Both the combustion air (ten percent excess air) and the anode vent gas are preheated to about 1200°F. The combustion zone and its heat exchanger train are pressurized to fuel-cell vent gas pressure, slightly less than 50 psia. The CO-shift reaction is moderately exothermic. This reaction heat is used to preheat the steam for steam generators in the fuel cell recirculating gas loop from about 375°F to about 600°F, and to preheat the feedwater to the steam generators in the fuel-cell recirculating gas loop. The CO-shift operates at 75 to 85 psia. A final processing of the exhaust gas from the CO-shift reactor removes the water from the fuel gas to a saturation temperature of 141°F in a recuperator/condenser unit. The dewatered fuel gas enters the fuel cell at 735°F and 50 psia. The water removed from the fuel gas is recirculated through the steam generator system, with additional water recovered from the cathode vent gas, with the steam-turbine condensate and with any required makeup water. A performance summary for this system is given in Table 5-22. The above performance is based on an average fuel cell temperature of 375°F, while all of the preceeding cases used 350°F for fuel cell temperature. For comparison purposes the heat rate for the Naphtha case at 350°F fuel cell temperature is 9450 Btu/kW-hr as shown in Table 5-23 which summarizes the performance of all the integrated systems.

5.2 DIRECT CAPITAL COST FOR PROTOTYPE SYSTEMS

The conceptual design layouts for each of the utility PAFC plants are presented in Section 5.1. Each concept considers one of the five fuels described in Section 4.1, namely, methanol, natural gas, ethanol, naphtha, and coal gas. Both nonintegrated and integrated fuel processing system (FPS) arrangements were evaluated for these fuels. The methanol system in particular also addressed a very low heat rate highly integrated FPS design with CO_2 clean-up. The direct capital cost evaluations were performed based on these ten cases.

The PAFC plants were designed, sized and costed to produce 7.5 M.We D.C. and to be self-sufficient. Cost estimates performed used 1980 cost levels. Guidelines were defined by the Federal Power Commission system of accounts (10) and the EPRI Technical Assessment Guide(11). Although detailed sizing was made for each of the system components, only the summary costs are given. For major components such as the fuel cell assembly a $326/kWe was used based upon Energy Research

Table 5-22

PERFORMANCE SUMMARY NAPHTHA FUELED INTEGRATED

Gross Electrical dc Output (kW)	7500
Gross Electrical ac Output (kW)	7200
Other Electrical Power Outputs (kW)	
Low Pressure Steam Turbine	365
Expander	911
Parasitic Losses	
Air Compressor	1374
Recirculator and Motor	444
Net Electrical ac Output (kW)	6658
Input Energy (hhv = 20593 Btu/lb) Btu/hr	60.469×10^6
Overall Plant Efficiency	0.376
Heat Rate (Btu/kW-hr)	9082*

* 375°F Average Fuel Cell Temperature

Table 5-23

PERFORMANCE COMPARISON INTEGRATED SYSTEMS

FUEL	METHANOL	METHANOL (CO_2 REMOVAL)	NATURAL GAS	COAL GAS	NAPHTHA
Gross electrical dc output (kW)	7500	7500	7500	7500	7500
Gross electrical ac output (kW)	7200	7200	7200	7200	8476
Parasitic Losses:					
Pumps	22	23	26	23	-
Fans (cooling tower)	150	135	200	205	-
Vacuum pump	3	3	3	3	-
Air Compressor	1	1	1	1	1374
Controls	15	15	15	15	-
Rotating group motor	93	318	0	0	444
Net electrical ac output (kW)	6916	6705	6955	6953	6658
High heating value (Btu/lb)	9758	9758	23890	4490	20593
Input energy (10^6 Btu/hr)	54.10	49.66	59.56	59.22	62.92
Overall plant efficiency	0.44	0.46	0.40	0.40	0.38
Heat rate (Btu/kW-hr)	7822	7406	8438	8520	9450
Fuel utilization	0.9	0.9	0.8	0.8	0.8

Corporation's estimated stack cost. A total fuel cell system including assembly, instrumentation, installation and fee is approximately $558/kWe. For the FPS the costs were established based upon vendor surveys discussed in Section 4.2.

Tables 5-24 through 5-33 show the estimates of direct capital cost for the seven prototype PAFC plants described above. These estimates include equipment, foundations, instrumentation and installation costs. All costs have been increased by 25 percent for IR&D, G&A and fee. Where ever possible, the cost of major equipment items was based on estimates quoted by equipment suppliers.

The capital equipment in these Tables includes structures and improvements which considers equipment such as the control trailer, HVAC, lighting, drainage systems, sewage systems, security fences, fire protection systems, communication systems, landscaping, paving, etc. Fuel handling and processing equipment such as storage of liquid fuels (as in the case for methanol), pumps, piping, fittings, insulation, vaporizer, (needed for liquid fuels), heat exchangers, reformer (not necessary for low methane coal gas), shift converter(s), combustor, blowers, instrumentation, foundations, and supports, etc. The rotating equipment and auxiliaries systems which account for the steam turbine and expander, condenser separators, CO_2 scrubber system (if required), induced draft cooling towers, air compressor system, filters, gas circulator, piping, fittings, pumps, instrumentation, foundations, and supports, etc. The electrical generating system, which has a low pressure boiler system, recirculation ducts, fuel cell assemblies, piping and fittings, supports, instrumentation and controls, foundations, etc. The accessory electric equipment which consists of a power conversion system, data acquisition system, instrumentation and controls, diesel generator system (if required) etc. Other miscellaneous power plant equipment such as the compressed air system for pneumatic valves, etc., water treatment and storage, inerting system, hydrogen system, sampling system, cranes, highlights, etc. Station equipment which includes the main transformer and any other site related equipment interfacing with the utility grid.

A summary of direct capital cost for the ten cases is given in Table 5-34. The significant cost differences between each of the plants is in the Fuel Handling and Processing and the Rotating Equipment and Auxiliaries accounts. All other accounts remain virtually the same except in those cases where the number of fuel cell modules must be adjusted. For the three methanol cases, the largest

Table 5-24

METHANOL PROTOTYPE 6.95 MW_e AC PAFC PLANT NONINTEGRATED DIRECT CAPITAL COST - 1980 BASIS (THOUSANDS OF DOLLARS)

ACCOUNT NO.(2)	$$10^3$
341 - STRUCTURES AND IMPROVEMENTS	163
342 - FUEL HANDLING AND PROCESSING(4)	2201
• FUEL HANDLING	64
• FUEL PROCESSING	2137
343 - ROTATING EQUIPMENT AND AUXILIARIES	812
• STEAM TURBINE/EXPANDER	177
• CONDENSER SYSTEM	32
• SEPARATOR SYSTEM	22
• COOLING TOWER	133
• AIR COMPRESSOR SYSTEM	295
• AIR FILTER/SILENCER SYSTEM	2
• CATHODE EXHAUST GAS CIRCULATOR	98
• MISCELLANEOUS AUXILIARIES	53
344 - ELECTRICAL GENERATING SYSTEM	4009
• LOW PRESSURE BOILER SYSTEM	96
• RECIRCULATION DUCTS	38
• FUEL CELL SYSTEM	3875
345 - ACCESSORY ELECTRIC EQUIPMENT	2455
• POWER CONVERSION SYSTEM	1712
• INSTRUMENTATION & CONTROL	390
• DAS SYSTEM	245
• DIESEL GENERATOR SYSTEM	108
346 - OTHER MISCELLANEOUS POWER PLANT EQUIPMENT	265
353 - STATION EQUIPMENT	548
• MAIN TRANSFORMER	548
TOTAL DIRECT CAPITAL COST(1) (3) (LAND NOT INCLUDED)	$10453 ($1504/kW)(5)

(1)IR&D, G&A, and Fee (25%) are included.

(2)Federal Power Commission Uniform Systems of Accounts for Public Utilities.

(3)EPRI Technical Assessment Guide

(4)Cost of initial catalyst is included in fuel processor cost ($250 K).

(5)Based on A.C. power to utility grid.

Table 5-25

METHANOL PROTOTYPE 6.92 MW_e AC PAFC PLANT INTEGRATED
DIRECT CAPITAL COST - 1980 BASIS
(THOUSANDS OF DOLLARS)

ACCOUNT NO.(2)	$\$10^3$
341 - STRUCTURES AND IMPROVEMENTS	163
342 - FUEL HANDLING AND PROCESSING(4)	3299
• FUEL HANDLING	64
• FUEL PROCESSING	3235
343 - ROTATING EQUIPMENT AND AUXILIARIES	812
• STEAM TURBINE/EXPANDER	177
• CONDENSER SYSTEM	32
• SEPARATOR SYSTEM	22
• COOLING TOWER	133
• AIR COMPRESSOR SYSTEM	295
• AIR FILTER/SILENCER SYSTEM	2
• CATHODE EXHAUST GAS CIRCULATOR	98
• MISCELLANEOUS AUXILIARIES	53
344 - ELECTRICAL GENERATING SYSTEM	4009
• LOW PRESSURE BOILER SYSTEM	96
• RECIRCULATION DUCTS	38
• FUEL CELL SYSTEM	3875
345 - ACCESSORY ELECTRIC EQUIPMENT	2455
• POWER CONVERSION SYSTEM	1712
• INSTRUMENTATION & CONTROL	390
• DAS SYSTEM	245
• DIESEL GENERATOR SYSTEM	108
346 - OTHER MISCELLANEOUS POWER PLANT EQUIPMENT	265
353 - STATION EQUIPMENT	548
• MAIN TRANSFORMER	548
TOTAL DIRECT CAPITAL COST (1) (3) (LAND NOT INCLUDED)	$11551 ($1669/kW)(5)

(1) IR&D, G&A, and Fee (25%) are included.

(2) Federal Power Commission Uniform Systems of Accounts for Public Utilities.

(3) EPRI Technical Assessment Guide

(4) Cost of initial catalyst is included in fuel processor cost ($250 K).

(5) Based on A.C. power to utility grid.

Table 5-26

METHANOL PROTOTYPE 6.71 MW_e AC PAFC PLANT
HIGHLY INTEGRATED
DIRECT CAPITAL COST - 1980 BASIS
(THOUSANDS OF DOLLARS)

ACCOUNT NO.(2)	$\$10^3$
341 - STRUCTURES AND IMPROVEMENTS	163
342 - FUEL HANDLING AND PROCESSING(4)	3299
• FUEL HANDLING	64
• FUEL PROCESSING	3235
343 - ROTATING EQUIPMENT AND AUXILIARIES	2249
• CO_2 SCRUBBER SYSTEM(6)	1437
• STEAM TURBINE/EXPANDER	177
• CONDENSER SYSTEM	32
• SEPARATOR SYSTEM	22
• COOLING TOWER	133
• AIR COMPRESSOR SYSTEM	295
• AIR FILTER/SILENCER SYSTEM	2
• CATHODE EXHAUST GAS CIRCULATOR	98
• MISCELLANEOUS AUXILIARIES	53
344 - ELECTRICAL GENERATING SYSTEM	4028
• LOW PRESSURE BOILER SYSTEM	115
• RECIRCULATION DUCTS	38
• FUEL CELL SYSTEM	3875
345 - ACCESSORY ELECTRIC EQUIPMENT	2455
• POWER CONVERSION SYSTEM	1712
• INSTRUMENTATION AND CONTROL	390
• DAS SYSTEM	245
• DIESEL GENERATOR SYSTEM	108
346 - OTHER MISCELLANEOUS POWER PLANT EQUIPMENT	265
353 - STATION EQUIPMENT	548
• MAIN TRANSFORMER	548
TOTAL DIRECT CAPITAL COST (1) (3) (LAND NOT INCLUDED)	\$13007 (\$1938/kW)(5)

(1) IR&D, G&A, and Fee (25%) are included.

(2) Federal Power Commission Uniform Systems of Accounts for Public Utilities.

(3) EPRI Technical Assessment Guide

(4) Cost of initial catalyst is included in fuel processor cost (\$250 K).

(5) Based on A.C. power to utility grid.

(6) Sized and costed based on a commercial K_2CO_3 system for CO_2 removable from stack gases.

Table 5-27

NATURAL GAS PROTOTYPE 6.95 MW_e AC PAFC PLANT
NONINTEGRATED
DIRECT CAPITAL COST - 1980 BASIS
(THOUSANDS OF DOLLARS)

ACCOUNT NO.(2)	$$10^3$
341 - STRUCTURES AND IMPROVEMENTS	163
342 - FUEL HANDLING AND PROCESSING(4)	2914
• FUEL HANDLING	64
• FUEL PROCESSING	2850
343 - ROTATING EQUIPMENT AND AUXILIARIES	812
• STEAM TURBINE/EXPANDER	177
• CONDENSER SYSTEM	32
• SEPARATOR SYSTEM	22
• COOLING TOWER	133
• AIR COMPRESSOR SYSTEM	295
• AIR FILTER/SILENCER SYSTEM	2
• CATHODE EXHAUST GAS CIRCULATOR	98
• MISCELLANEOUS AUXILIARIES	53
344 - ELECTRICAL GENERATING SYSTEM	3990
• LOW PRESSURE BOILER SYSTEM	96
• RECIRCULATION DUCTS	38
• FUEL CELL SYSTEM	3856
345 - ACCESSORY ELECTRIC EQUIPMENT	2455
• POWER CONVERSION SYSTEM	1712
• INSTRUMENTATION & CONTROL	390
• DAS SYSTEM	245
• DIESEL GENERATOR SYSTEM	108
346 - OTHER MISCELLANEOUS POWER PLANT EQUIPMENT	265
353 - STATION EQUIPMENT	548
• MAIN TRANSFORMER	548
TOTAL DIRECT CAPITAL COST (1) (3) (LAND NOT INCLUDED)	$11147 ($1604/kW)(5)

(1) IR&D, G&A, and Fee (25%) are included.

(2) Federal Power Commission Uniform Systems of Accounts for Public Utilities.

(3) EPRI Technical Assessment Guide

(4) Cost of initial catalyst is included in fuel processor cost ($250 K).

(5) Based on A.C. power to utility grid.

Table 5-28

NATURAL GAS PROTOTYPE 6.95 MW_e AC PAFC PLANT INTEGRATED DIRECT CAPITAL COST - 1980 BASIS (THOUSANDS OF DOLLARS)

ACCOUNT NO.(2)	$$10^3$
341 - STRUCTURES AND IMPROVEMENTS	163
342 - FUEL HANDLING AND PROCESSING(4)	4378
• FUEL HANDLING	64
• FUEL PROCESSING	4314
343 - ROTATING EQUIPMENT AND AUXILIARIES	812
• STEAM TURBINE/EXPANDER	177
• CONDENSER SYSTEM	32
• SEPARATOR SYSTEM	22
• COOLING TOWER	133
• AIR COMPRESSOR SYSTEM	295
• AIR FILTER/SILENCER SYSTEM	2
• CATHODE EXHAUST GAS CIRCULATOR	98
• MISCELLANEOUS AUXILIARIES	53
344 - ELECTRICAL GENERATING SYSTEM	3990
• LOW PRESSURE BOILER SYSTEM	96
• RECIRCULATION DUCTS	38
• FUEL CELL SYSTEM	3856
345 - ACCESSORY ELECTRIC EQUIPMENT	2455
• POWER CONVERSION SYSTEM	1712
• INSTRUMENTATION & CONTROL	390
• DAS SYSTEM	245
• DIESEL GENERATOR SYSTEM	108
346 - OTHER MISCELLANEOUS POWER PLANT EQUIPMENT	265
353 - STATION EQUIPMENT	548
• MAIN TRANSFORMER	548
TOTAL DIRECT CAPITAL COST (1) (3) (LAND NOT INCLUDED)	$12611 ($1815/kW)(5)

---

(1) IR&D, G&A, and Fee (25%) are included.

(2) Federal Power Commission Uniform Systems of Accounts for Public Utilities.

(3) EPRI Technical Assessment Guide

(4) Cost of initial catalyst is included in fuel processor cost ($250 K).

(5) Based on A.C. power to utility grid.

Table 5-29

COAL GAS PROTOTYPE 8.10 MW_e AC PAFC PLANT NONINTEGRATED DIRECT CAPITAL COST - 1980 BASIS (THOUSANDS OF DOLLARS)

ACCOUNT NO.(2)	$$10^3$
341 - STRUCTURES AND IMPROVEMENTS	163
342 - FUEL HANDLING AND PROCESSING(4)	2171
• FUEL HANDLING	64
• FUEL PROCESSING	2107
343 - ROTATING EQUIPMENT AND AUXILIARIES	1074
• GAS TURBINE/GENERATOR SYSTEM	262
• STEAM TURBINE/EXPANDER	177
• CONDENSER SYSTEM	32
• SEPARATOR SYSTEM	22
• COOLING TOWER	133
• AIR COMPRESSOR SYSTEM	295
• AIR FILTER/SILENCER SYSTEM	2
• CATHODE EXHAUST GAS CIRCULATOR	98
• MISCELLANEOUS AUXILIARIES	53
344 - ELECTRICAL GENERATING SYSTEM	4067
• LOW PRESSURE BOILER SYSTEM	96
• RECIRCULATION DUCTS	38
• FUEL CELL SYSTEM	3933
345 - ACCESSORY ELECTRIC EQUIPMENT	2455
• POWER CONVERSION SYSTEM	1712
• INSTRUMENTATION AND CONTROL	390
• DAS SYSTEM	245
• DIESEL GENERATOR SYSTEM	108
346 - OTHER MISCELLANEOUS POWER PLANT EQUIPMENT	265
353 - STATION EQUIPMENT	548
• MAIN TRANSFORMER	548
TOTAL DIRECT CAPITAL COST(1)(3) (LAND NOT INCLUDED)	$10743 ($1326/kW)(5)

(1) IR&D, G&A, and Fee (25%) are included.

(2) Federal Power Commission Uniform Systems of Accounts for Public Utilities.

(3) EPRI Technical Assessment Guide.

(4) Cost of initial catalyst is included in fuel processor cost (≈$200 K).

(5) Based on A.C. power to utility grid.

Table 5-30

COAL GAS PROTOTYPE 6.95 MW_e AC PAFC PLANT
INTEGRATED
DIRECT CAPITAL COST - 1980 BASIS
(THOUSANDS OF DOLLARS)

ACCOUNT NO.(2)	$\$10^3$
341 - STRUCTURES AND IMPROVEMENTS	163
342 - FUEL HANDLING AND PROCESSING(4)	2171
● FUEL HANDLING	64
● FUEL PROCESSING	2107
343 - ROTATING EQUIPMENT AND AUXILIARIES	3106
● CO_2 SCRUBBER SYSTEM(6)	2294
● STEAM TURBINE/EXPANDER	177
● CONDENSER SYSTEM	32
● SEPARATOR SYSTEM	22
● COOLING TOWER	133
● AIR COMPRESSOR SYSTEM	295
● AIR FILTER/SILENCER SYSTEM	2
● CATHODE EXHAUST GAS CIRCULATOR	98
● MISCELLANEOUS AUXILIARIES	53
344 - ELECTRICAL GENERATING SYSTEM	4067
● LOW PRESSURE BOILER SYSTEM	96
● RECIRCULATION DUCTS	38
● FUEL CELL SYSTEM	3933
345 - ACCESSORY ELECTRIC EQUIPMENT	2455
● POWER CONVERSION SYSTEM	1712
● INSTRUMENTATION AND CONTROL	390
● DAS SYSTEM	245
● DIESEL GENERATOR SYSTEM	108
346 - OTHER MISCELLANEOUS POWER PLANT EQUIPMENT	265
353 - STATION EQUIPMENT	548
● MAIN TRANSFORMER	548
TOTAL DIRECT CAPITAL COST (1)(3) (LAND NOT INCLUDED)	\$12775 (\$1838/kW)(5)

(1) IR&D, G&A, and Fee (25%) are included.

(2) Federal Power Commission Uniform Systems of Accounts for Public Utilities.

(3) EPRI Technical Assessment Guide

(4) Cost of initial catalyst is included in fuel processor cost (≈\$200 K).

(5) Based on A.C. power to utility grid.

(6) Sized and costed based on a commercial K_2CO_3 System for CO_2 removable from stack gases.

Table 5-31

ETHANOL PROTOTYPE 6.95 MW_e AC PAFC PLANT
NONINTEGRATED
DIRECT CAPITAL COST - 1980 BASIS
(THOUSANDS OF DOLLARS)

ACCOUNT NO.(2)	$\$10^3$
341 - STRUCTURES AND IMPROVEMENTS	163
342 - FUEL HANDLING AND PROCESSING(4)	2201
• FUEL HANDLING	64
• FUEL PROCESSING	2137
343 - ROTATING EQUIPMENT AND AUXILIARIES	879
• STEAM TURBINE/EXPANDER	177
• CONDENSER SYSTEM	32
• SEPARATOR SYSTEM	22
• COOLING TOWER	133
• AIR COMPRESSOR SYSTEM	358
• AIR FILTER/SILENCER SYSTEM	2
• CATHODE EXHAUST GAS CIRCULATOR	98
• MISCELLANEOUS AUXILIARIES	57
344 - ELECTRICAL GENERATING SYSTEM	4048
• LOW PRESSURE BOILER SYSTEM	123
• RECIRCULATION DUCTS	38
• FUEL CELL SYSTEM	3887
345 - ACCESSORY ELECTRIC EQUIPMENT	2455
• POWER CONVERSION SYSTEM	1712
• INSTRUMENTATION & CONTROL	390
• DAS SYSTEM	245
• DIESEL GENERATOR SYSTEM	108
346 - OTHER MISCELLANEOUS POWER PLANT EQUIPMENT	265
353 - STATION EQUIPMENT	548
• MAIN TRANSFORMER	548
TOTAL DIRECT CAPITAL COST(1) (3) (LAND NOT INCLUDED)	\$10559 (\$1519/kW)(5)

(1) IR&D, G&A, and Fee (25%) are included.

(2) Federal Power Commission Uniform Systems of Accounts for Public Utilities.

(3) EPRI Technical Assessment Guide

(4) Cost of initial catalyst is included in fuel processor cost (\$250 K).

(5) Based on A.C. power to utility grid.

Table 5-32

NAPHTHA PROTOTYPE 6.95 MW_e AC PAFC PLANT
NONINTEGRATED
DIRECT CAPITAL COST - 1980 BASIS
(THOUSANDS OF DOLLARS)

ACCOUNT NO.(2)	$\$10^3$
341 - STRUCTURES AND IMPROVEMENTS	163
342 - FUEL HANDLING AND PROCESSING(4)	3199
• FUEL HANDLING	64
• FUEL PROCESSING	3135
343 - ROTATING EQUIPMENT AND AUXILIARIES	950
• STEAM TURBINE/EXPANDER	177
• CONDENSER SYSTEM	32
• SEPARATOR SYSTEM	22
• COOLING TOWER	133
• AIR COMPRESSOR SYSTEM	423
• AIR FILTER/SILENCER SYSTEM	3
• CATHODE EXHAUST GAS CIRCULATOR	98
• MISCELLANEOUS AUXILIARIES	62
344 - ELECTRICAL GENERATING SYSTEM	3997
• LOW PRESSURE BOILER SYSTEM	96
• RECIRCULATION DUCTS	38
• FUEL CELL SYSTEM	3863
345 - ACCESSORY ELECTRIC EQUIPMENT	2455
• POWER CONVERSION SYSTEM	1712
• INSTRUMENTATION & CONTROL	390
• DAS SYSTEM	245
• DIESEL GENERATOR SYSTEM	108
346 - OTHER MISCELLANEOUS POWER PLANT EQUIPMENT	265
353 - STATION EQUIPMENT	548
• MAIN TRANSFORMER	548
TOTAL DIRECT CAPITAL COST (1) (3) (LAND NOT INCLUDED)	$11577 ($1666/kW)(5)

(1)IR&D, G&A, and Fee (25%) are included.

(2)Federal Power Commission Uniform Systems of Accounts for Public Utilities.

(3)EPRI Technical Assessment Guide

(4)Cost of initial catalyst is included in fuel processor cost ($360 K)

(5)Based on A.C. power to utility grid.

Table 5-33

NAPHTHA PROTOTYPE 6.95 MW_e AC PAFC PLANT
INTEGRATED
DIRECT CAPITAL COST - 1980 BASIS
(THOUSANDS OF DOLLARS)

ACCOUNT NO.(2)	$10^3
341 - STRUCTURES AND IMPROVEMENTS	163
342 - FUEL HANDLING AND PROCESSING(4)	4767
● FUEL HANDLING	4703
● FUEL PROCESSING	4703
343 - ROTATING EQUIPMENT AND AUXILIARIES	812
● STEAM TURBINE/EXPANDER	177
● CONDENSER SYSTEM	32
● SEPARATOR SYSTEM	22
● COOLING TOWER	133
● AIR COMPRESSOR SYSTEM	295
● AIR FILTER/SILENCER SYSTEM	2
● CATHODE EXHAUST GAS CIRCULATOR	98
● MISCELLANEOUS AUXILIARIES	53
344 - ELECTRICAL GENERATING SYSTEM	3997
● LOW PRESSURE BOILER SYSTEM	96
● RECIRCULATION DUCTS	38
● FUEL CELL SYSTEM	3863
345 - ACCESSORY ELECTRIC EQUIPMENT	2455
● POWER CONVERSION SYSTEM	1712
● INSTRUMENTATION & CONTROL	390
● DAS SYSTEM	245
● DIESEL GENERATOR SYSTEM	108
346 - OTHER MISCELLANEOUS POWER PLANT EQUIPMENT	265
353 - STATION EQUIPMENT	548
● MAIN TRANSFORMER	548
TOTAL DIRECT CAPITAL COST (1) (3) (LAND NOT INCLUDED)	$13007 ($1872/kW)(5)

(1) IR&D, G&A, and Fee (25%) are included.

(2) Federal Power Commission Uniform Systems of Accounts for Public Utilities.

(3) EPRI Technical Assessment Guide

(4) Cost of initial catalyst is included in fuel processor cost ($360 K).

(5) Based on A.C. power to utility grid.

Table 5-34

SUMMARY OF PROTOTYPE PAFC PLANT DIRECT CAPITAL COST (1) - 1980 BASIS
(thousands of dollars)

ACCOUNT NO.(2)	METHANOL NON-INTEGRATED	METHANOL INTEGRATED	METHANOL HIGHLY INTEGRATED	NATURAL GAS NON-INTEGRATED	NATURAL GAS INTEGRATED	COAL GAS NON-INTEGRATED	COAL GAS INTEGRATED	ETHANOL NON-INTEGRATED	NAPHTHA NON-INTEGRATED	NAPHTHA INTEGRATED
341 Structure and Improvements	163	163	163	163	163	163	163	163	163	163
342 Fuel Handling and Processing(4)	2201	3299	3299	2914	4378	2171	2171	2201	3199	4767
343 Rotating Equipment and Auxiliaries	812	812	2249	812	812	1074	3106	879	950	812
344 Electrical Generating System	4009	4009	4028	3990	3990	4067	4067	4067	4048	3997
345 Accessory Electric Equipment	2455	2455	2455	2455	2455	2455	2455	2455	2455	2455
346 Other Miscellaneous Power Plant Equipment	265	265	265	265	265	265	265	265	265	265
353 Station Equipment	548	548	548	548	548	548	548	548	548	548
Total Direct Capital Cost(3)	10453	11551	13007	11156	12611	10743	12775	10559	11577	13007
$/kW$_e$, A.C.	1504	1669	1938	1604	1815	1326	1838	1519	1666	1872

(1) IR&D, G&A, and Fee (25% total) are included.

(2) Federal Power Commission Uniform Systems of Accounts for Public Utilities

(3) EPRI Technical Assessment Guide

(4) Cost of initial catalyst is included in fuel processor cost.

increases are the complex integration of the fuel processing system with the BOP equipment (nonintegrated vs. integrated) and the addition of a CO_2 clean-up system (integrated vs. highly integrated). These increased costs add an additional 1.1 million dollars for integrating the fuel processor and approximate 1.4 million dollars for CO_2 clean up. Details of how these capital cost impact the overall total plant cost, fuel costs and operation and maintenance of the PAFC plant are described in Section 5.3.

For the natural gas plants the difference is the 1.5 million dollars to integrate the fuel processor. The rotating group and auxiliaries are unaffected by this change. However, for the coal gas plant design the opposite is true. Here, the 2 million dollar difference in direct capital cost is due to a large auxiliary CO_2 removal system. As with the methanol plant, the total plant cost, fuel costs and operation and maintenance are given in Section 5.3.

The size and cost for CO_2 clean-up was based on a commercially available K_2CO_3 system designed for removing CO_2 from stack gases. As affected above in the increased capital costs this system is very expensive. Other removable systems should be considered. Preliminary evaluations to date indicate that a water scrubbing system will work but more detailed studies are needed to verify performance (i.e., sufficient CO_2 removal capability, system size, and capital cost).

5.3 ECONOMIC ANALYSIS

The economic studies described herein are based in part on the recommendations documented in Table 1 of the 1979 EPRI Technical Assessment Guide (Reference 10). These recommendations and the assumptions used are also described in the following paragraphs. The methods and results of calculating the that plant investment costs for each one of the plant configurations described in Section 5.2 are included. Cost of Electricity has been calculated for each one of the plant configurations with different fuel costs, operating conditions, and economic assumptions.

5.3.1 TOTAL PLANT INVESTMENT COSTS

The total plant investment costs are based on the direct capital cost estimates for each plant configuration described in Section 5.2. The following assumptions and ground rules were used:

- All capital and investment costs are stated in end of year 1980 dollars.
- The cost of land and land rights was neglected.
- A contingency of 10 percent was used for both 50th units and projected commercial units.
- The design and construction schedule for all plant configurations was two years.
- Interest during construction was eight percent simple interest per year.
- Other indirect costs consisted of the following
  - 2.1 percent for construction facilities, equipment, and services
  - 3.5 percent for engineering services
  - 0.9 percent for insurance and taxes.

The factors used to determine each of the above indirect costs are consistent with those used in previous EPRI cost studies.

The construction facilities, equipment, and services include general costs associated with the plant construction. These indirect costs include such items as field offices, warehouses, temporary power and utility lines, cost or rental of construction equipment and supplies, purchase of electric power, water and other utilities, security guards, training programs for the labor force, inspection and testing of construction materials, site cleanup, insurance and any other general cost associated with plant construction.

The engineering services under indirect costs include items such as preliminary investigations; site selection; air and water environmental studies; subsurface investigations; preparation of specifications and evaluation of proposals for major equipment packages, and preparation of preliminary and final design documents.

Other indirect costs include the owner's property and risk insurance, state and local property taxes on the site and improvements during construction, sales taxes on purchases materials and equipment, staff training, plant startup, and other general costs.

Interest during construction is calculated as simple interest, at an eight percent annual rate for each year of the two year design and construction schedule.

In addition, the effect of experience curve factors of 85, 90 and 95 percent on the 50th unit direct capital cost was calculated. The direct capital costs for all of the cases studied are summarized in Table 5-35. The cases include 50th units and projected commercial units. The 50th units are regarded as units that will be available for utility installations by 1990. Projected commercial units are regarded as a learned-out mature design that will be available later in the 1990's. Our estimate is that the experience curve effect on the direct capital cost will be 90 percent for the 50th units. The projected commercial unit direct capital costs reflect a mature design with an experience curve effect of 85 percent at 135 units.

5.3.2 RELATIVE COST OF ELECTRICITY

The cost of electricity (COE) calculations performed herein are stated in terms of constant 1980 dollars and are not levelized calculations. COE calculations were made for each one of the ten fuel and plant configuration combinations given in Table 5-35. The capital, fuel and operation/maintenance components of the COE were calculated using the equations summarized in Table 5-36.

Selected parameters were varied to analyze their effect on the COE of each plant configuration. These parameters were as follows:

Parameters	Study Range
Fixed Charge Rate	12% and 18%
Capacity Factor	30% and 70%
Fuel Cost	2-14 $/MBtu
Experience Curve (50th Units)	90 ± 5%
Experience Curve (Projected Commercial 135th Unit)	85%

The experience curve concept is based on observations that costs of production appear to go down as the cumulative number of units produced increases. The experience curve cost reductions are due to the combined effects of improved labor efficiency, technical improvements, economies of scale and volume purchasing. Experience curves encompass all such cost effects, while learning curves generally apply only to improved labor efficiency. Reference (16) provides a summary of learning and experience concepts.

Table 5-35

SUMMARY OF PAFC PLANT DIRECT CAPITAL COSTS WITH EFFECT OF EXPERIENCE CURVE, 1980 DOLLARS

FUEL	PLANT TYPE	HEAT RATE (Btu/Kwh)	NET OUTPUT MWe (AC)	PROTOTYPE DIRECT CAPITAL COST, $/KWe	50TH UNIT DIRECT CAPITAL COST, $/KWe			PROJECTED COMMERCIAL CAPITAL COST, $/KWe
					@ 95% EXPERIENCE	@ 90% EXPERIENCE	@85% EXPERIENCE	
Methanol	Non Integrated	8790	6.95	1504	1126	830	601	476
	Integrated	7820	6.92	1669	1250	921	667	528
	Highly Integrated With CO_2 Clean Up	7390	6.71	1940	1452	1070	775	614
Natural Gas	Non Integrated	9270	6.95	1604	1203	887	642	508
	Integrated	8440	6.95	1815	1361	1003	726	575
Coal Gas	Non Integrated	9110	8.10	1326	988	728	527	420
	Integrated	8520	6.95	1838	1370	1010	731	582
Naphtha	Non Integrated	10748	6.95	1666	1247	919	666	527
	Integrated	9448	6.95	1872	1400	1033	748	593
Ethanol	Non Integrated	9058	6.95	1519	1136	838	607	481

Table 5-36

COE CALCULATION METHODOLOGY

$$\text{FIXED (CAPITAL) COST, (MILLS/kWh)} = \frac{\text{TOTAL PLANT INVESTMENT (\$/kWe) X FIXED CHARGE RATE X 1000}}{\text{8760 X CAPACITY FACTOR}}$$

$$\text{FUEL COST, (MILLS/kWh)} = \frac{\text{FUEL COST (\$/}10^{6}\text{ Btu) X HEAT RATE (Btu/kWh)}}{1000}$$

$$\text{O \& M COST, (MILLS/kWh)} = \text{VARIABLE O \& M COST (MILLS/kWh)} + \frac{\text{FIXED O \& M COST (\$/kW-yr) X 1000}}{\text{8760 X CAPACITY FACTOR}}$$

$$\text{COE (MILLS/kWh)} = \text{FIXED COST + FUEL COST = O \& M COST}$$

The O & M cost factors are based on the values given in Reference 10 but updated as follows to end-of-year 1980 dollars:

	1979 TAG (EOY 1978$'s)	PRESENT STUDY (EOY 1980 $'s)
Fixed O&M Cost, $/kW-YR	3.3	3.9
Variable O&M Cost, Mills/kWH	3.2	3.8

The following bar charts summarize the parametric calculations in relative COE terms. Since the basic capital and O&M costs included in the COE calculations are stated in terms of constant 1980 $'s, the effect on relative COE of any fuel cost (stated in terms of constant 1980 $'s) within the range of 2 - 14 $/MBtu may be determined from the bar charts. The bar charts are organized in the following categories:

- NonIntegrated 50th Units
- Integrated 50th Units
- NonIntegrated Projected Commercial Units
- Integrated Projected Commercial Units

COE results are given for all of the above in configurations that utilize natural gas, coal gas, methanol, naphtha, or ethanol as the source fuel.

Relative COE's* are used to facilitate the comparison of PAFC power plant configurations. As a result, the relative COE data presented herein should not be used to directly compare fuel cell power plant economic performance with that of alternate generation units.

Figure 5-12 presents the COE for the nonintegrated 50th units with the contribution to COE due to O&M, capital, and fuel costs shown. Figure 5-13 shows the effect of changing the fixed charge rate.

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*A relative COE of 1.0 is equal to 100 mills/kWh for all data presented herein.

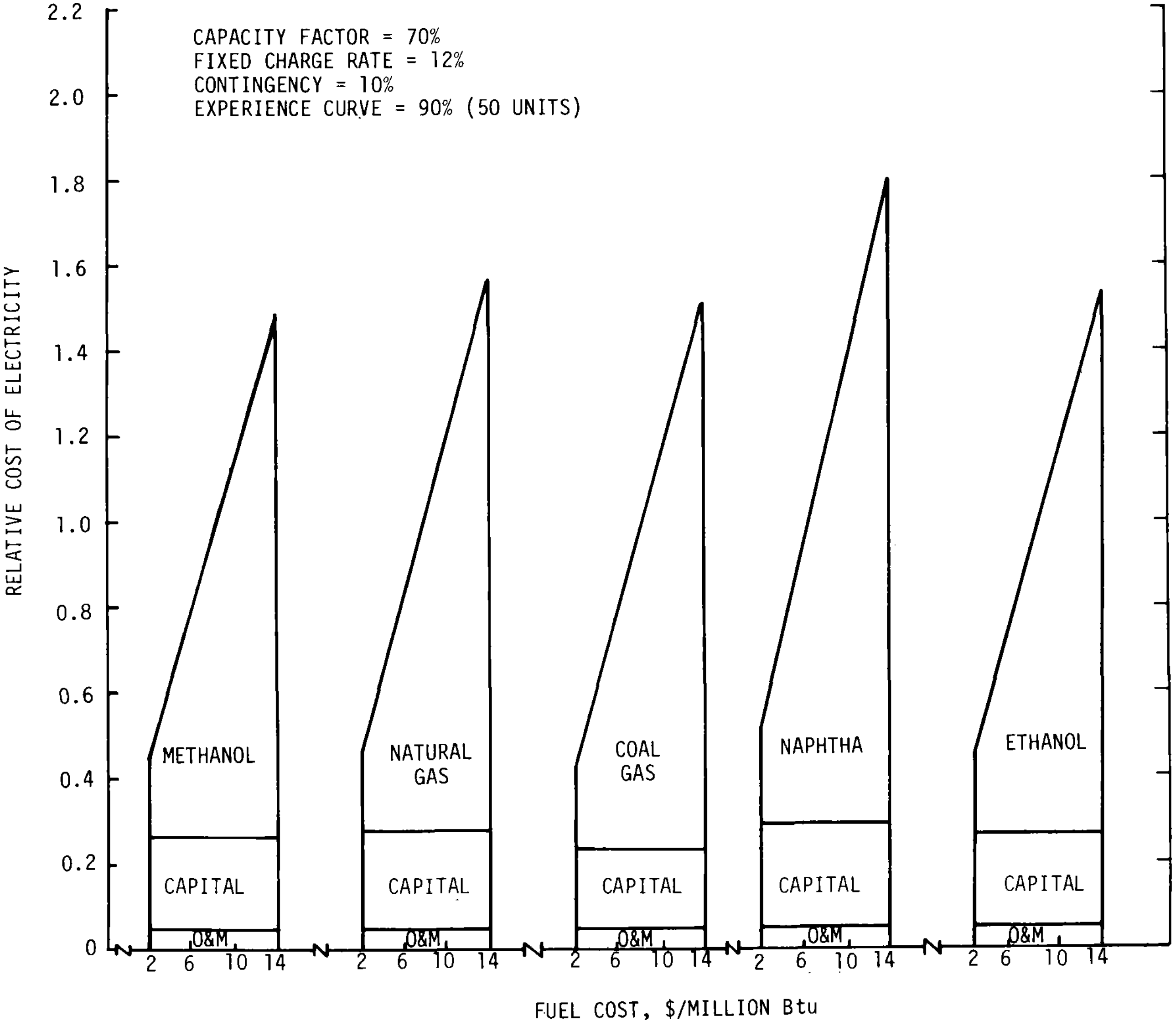

Figure 5-12. Relative COE Components for Nonintegrated 50th Units

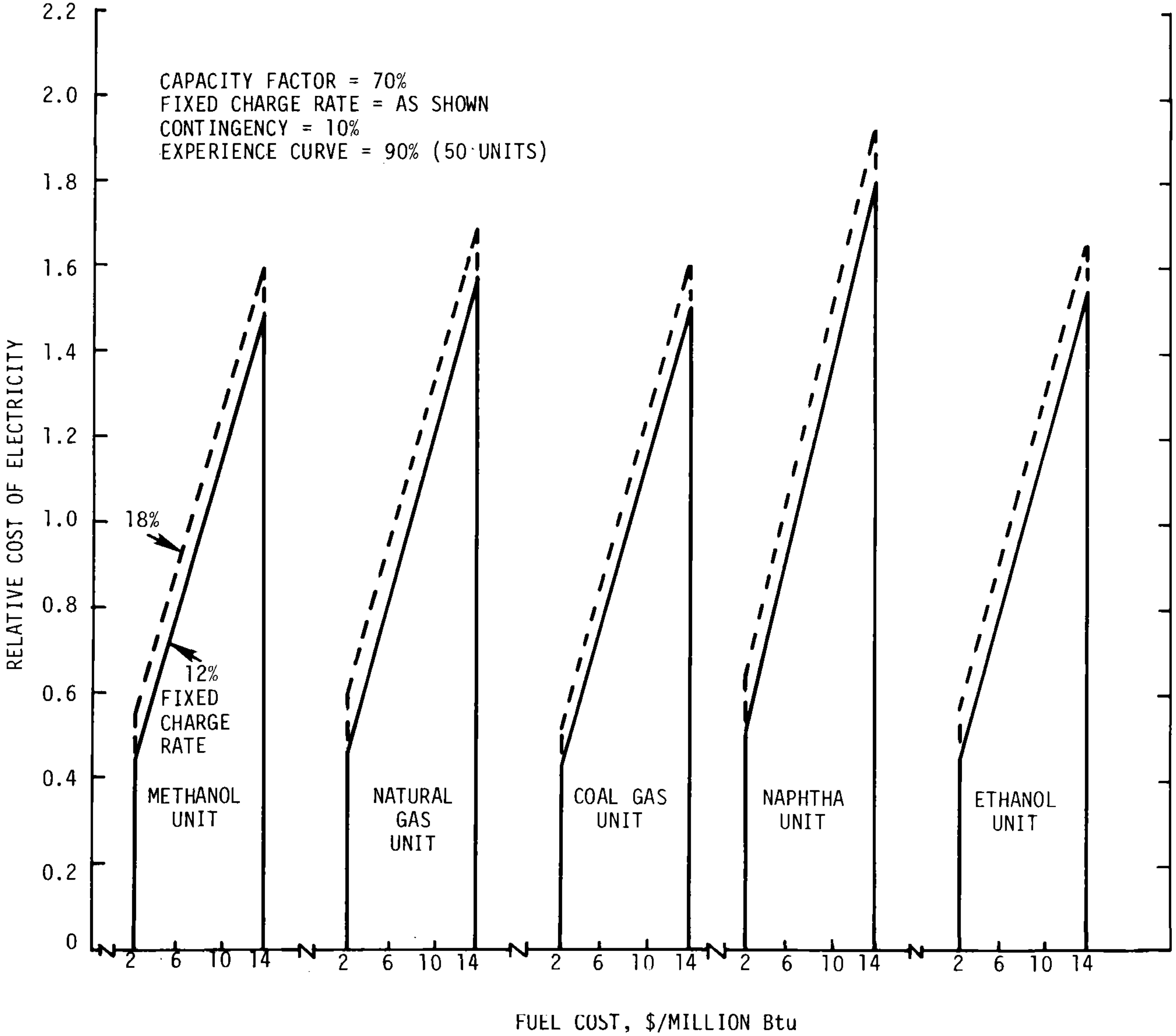

Figure 5-13. Effect of Fixed Charge Rate on Relative COE of Nonintegrated 50th Units

Figure 5-14 presents the COE for the integrated 50th units. While the highly integrated methanol unit with CO_2 clean up has the lowest heat rate, the associated increase in capital cost off sets the advantage when the 50th unit configurations are considered. Figure 5-15 shows the effect of changing the fixed change rate for the integrated 50th units.

Figures 5-16 presents the COE for projected commercial units in the nonintegrated configurations. The effect on COE of changing the capacity factor is shown in Figure 5-17 for these units.

Figures 5-18 and 5-19 present the COE and its compoents for the integrated commercial units and the COE effect of changing the capacity factor.

The decreasing effect of the capital cost component on relative COE as one moves from the 50th units to the projected commercial units can be seen in Figures 5-12 through 5-19. Fuel costs are by far the most significant contribution to the COE for the commercial units and this fact reinforces the need to design for minimum PAFC plant heat rates.

5.3.3 PLANT COST, PERFORMANCE AND COE SUMMARIES

The Fuel Cell Users Group Subcommittee on fuels published fuel price projections in Reference 13. This Subcommittee agreed on certain pricing trends which are:

1. The prices of commercial grade fuels will have converged in the 1985-90 time, and that synthetic fuels will roll in at similar prices as they become available in quantity.

2. There will be some price differentials between fuels reflecting differences in quality. For the fuels of interest for early commercial fuel cells, these differentials will be small fractions of the fuel cost. Possible differences in fuel cell heat rate and capital cost using various fuels would have a greater effect on cost of electricity, thus fuel quality price differentials will be ignored.

3. The price for No. 2 fuel oil can be used as the benchmark for other natural or synthetic commercial grade fuels that may be used in the initial fuel cells.

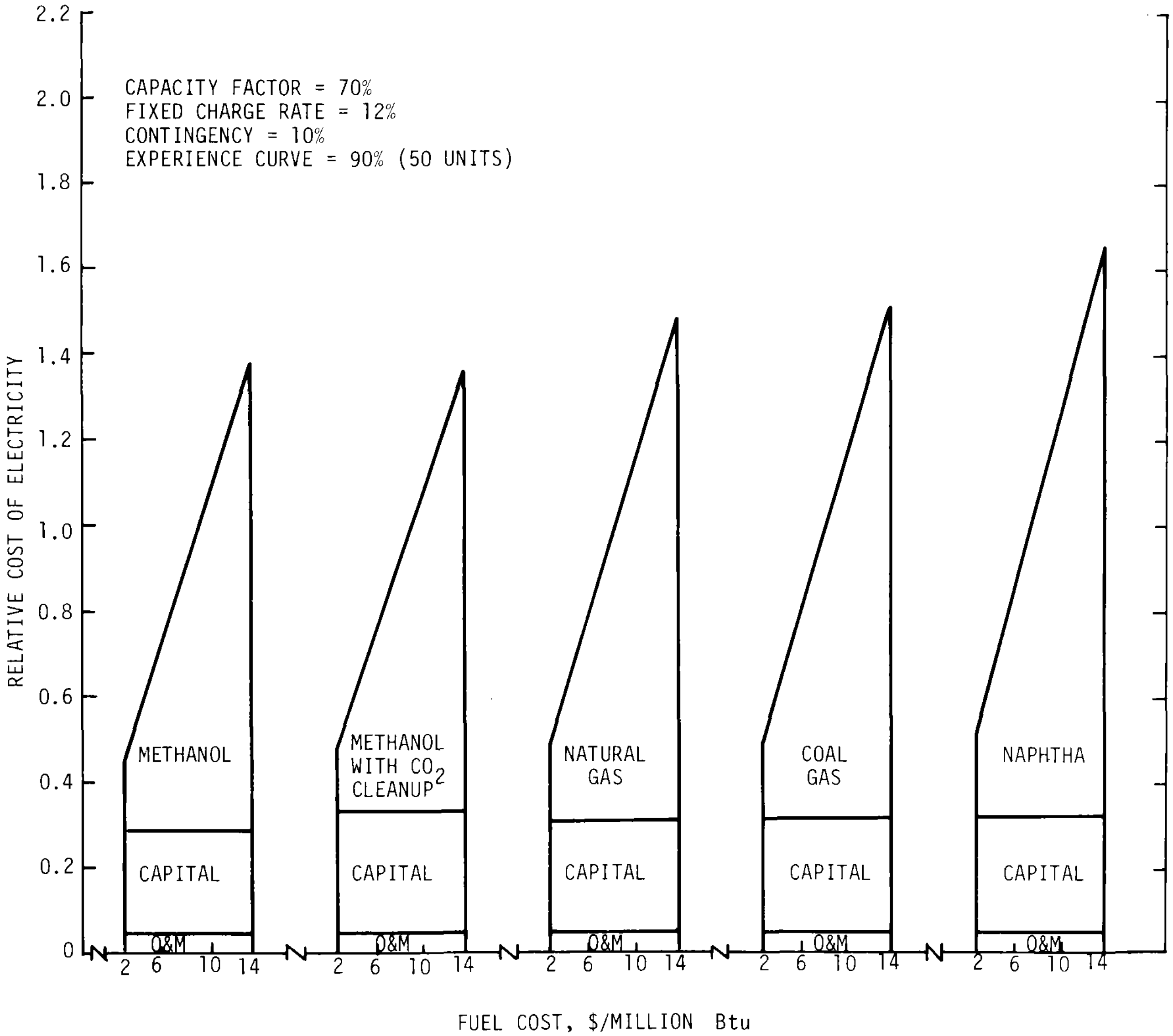

Figure 5-14. Relative COE Components for Integrated 50th Units

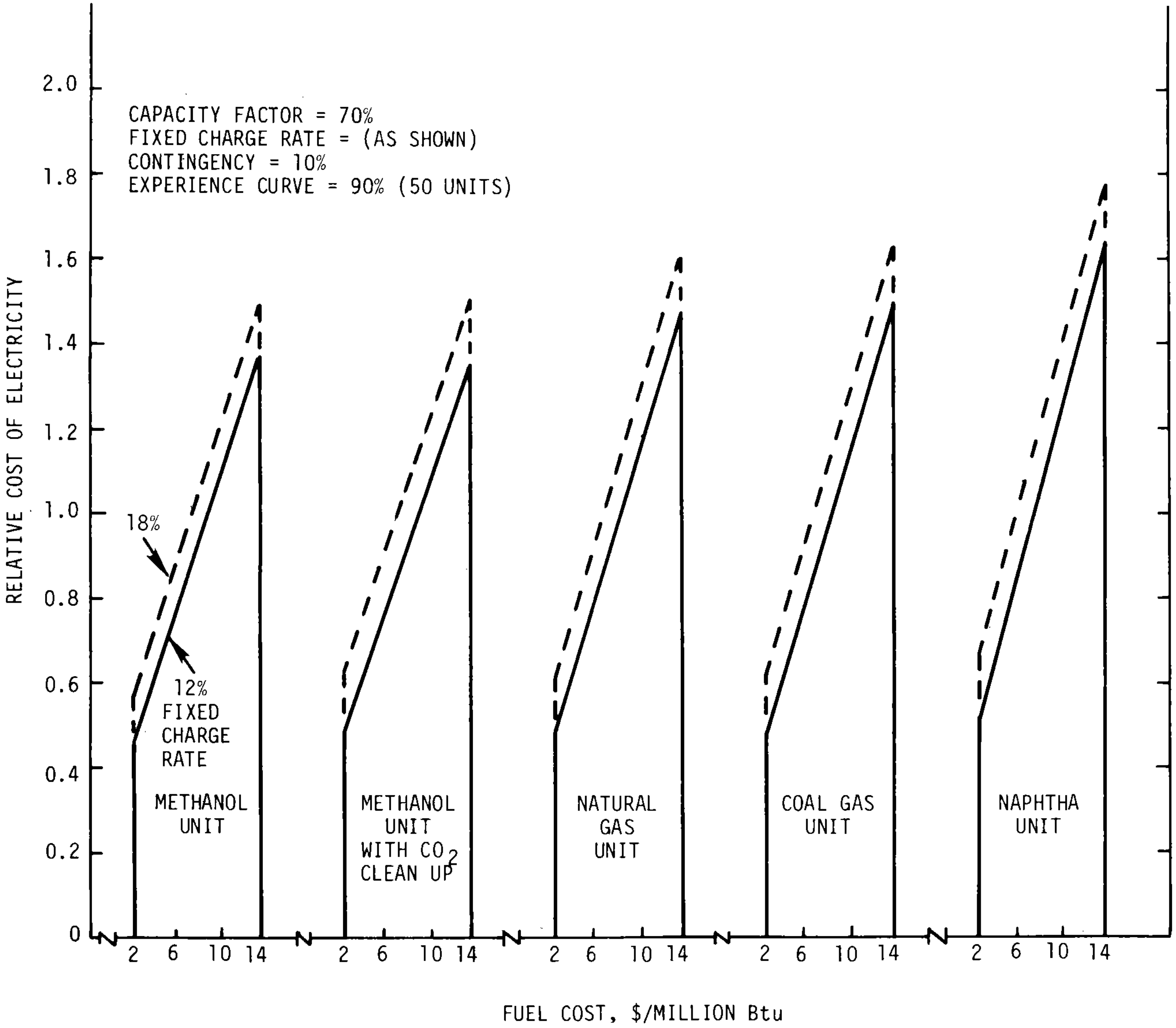

Figure 5-15. Effect of Fixed Charge Rate on Relative COE of Integrated 50th Units

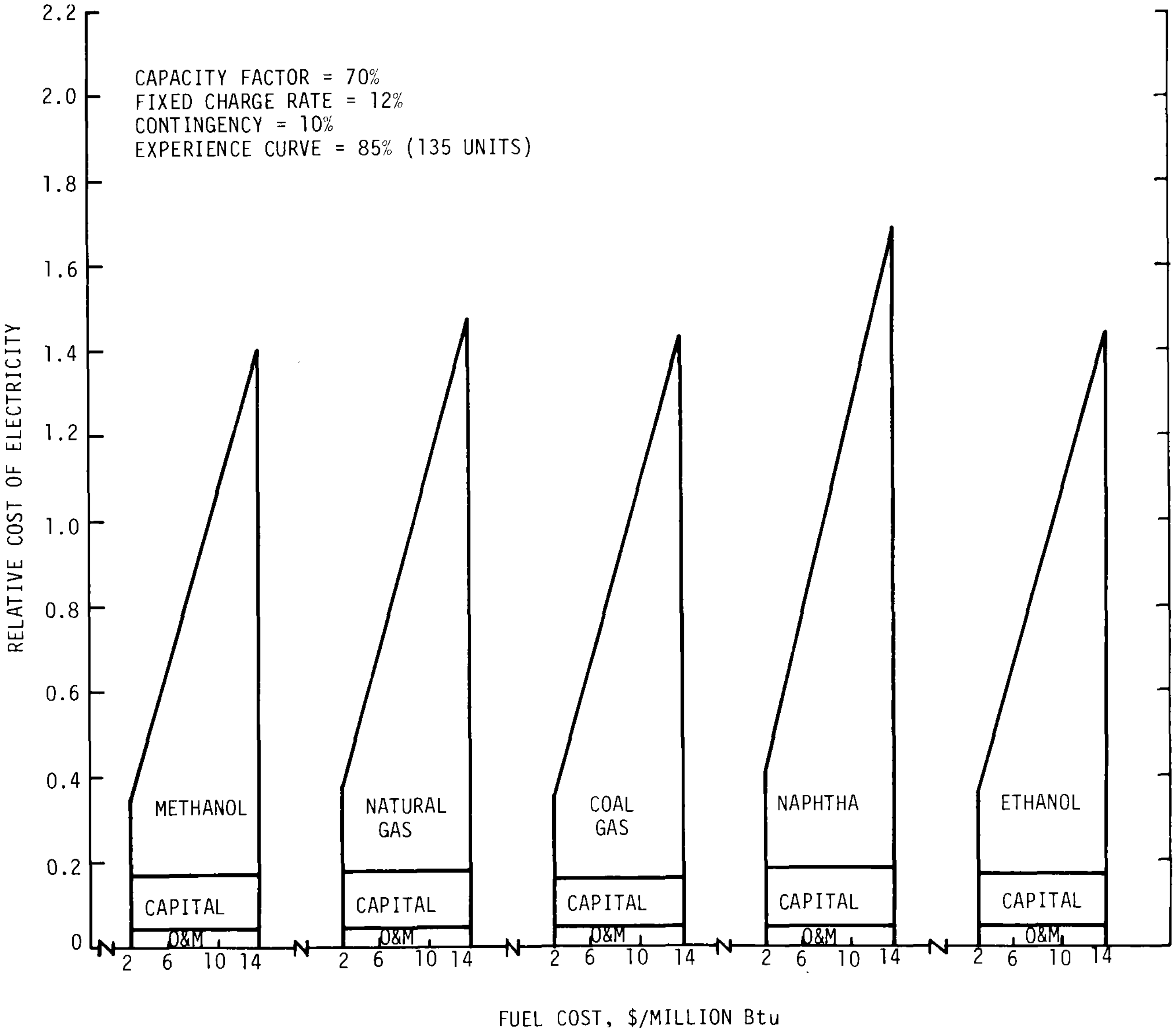

Figure 5-16. Relative COE Components for Nonintegrated Commercial Units

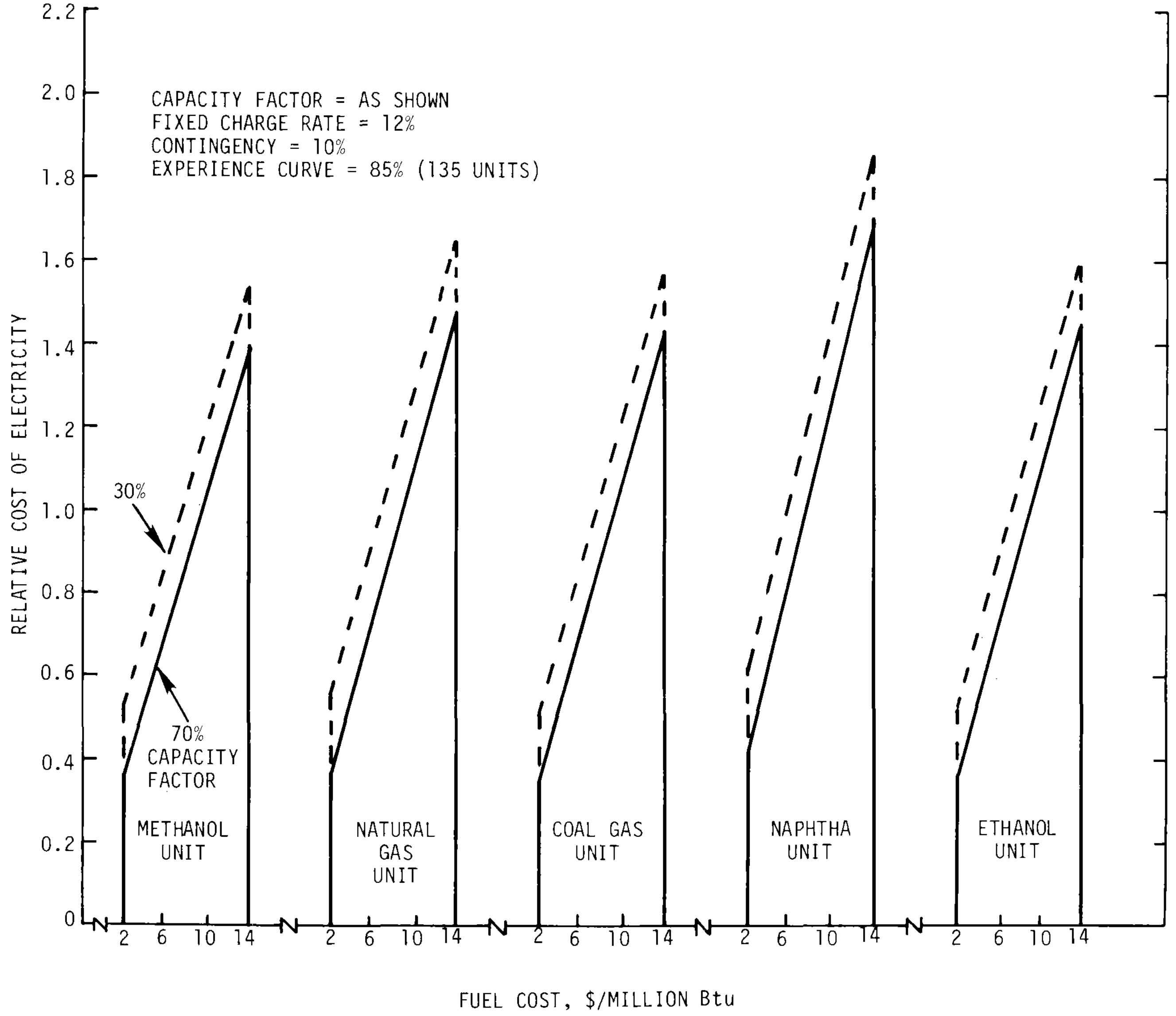

Figure 5-17. Effect of Capacity Factor on Relative COE of Nonintegrated Commercial Units

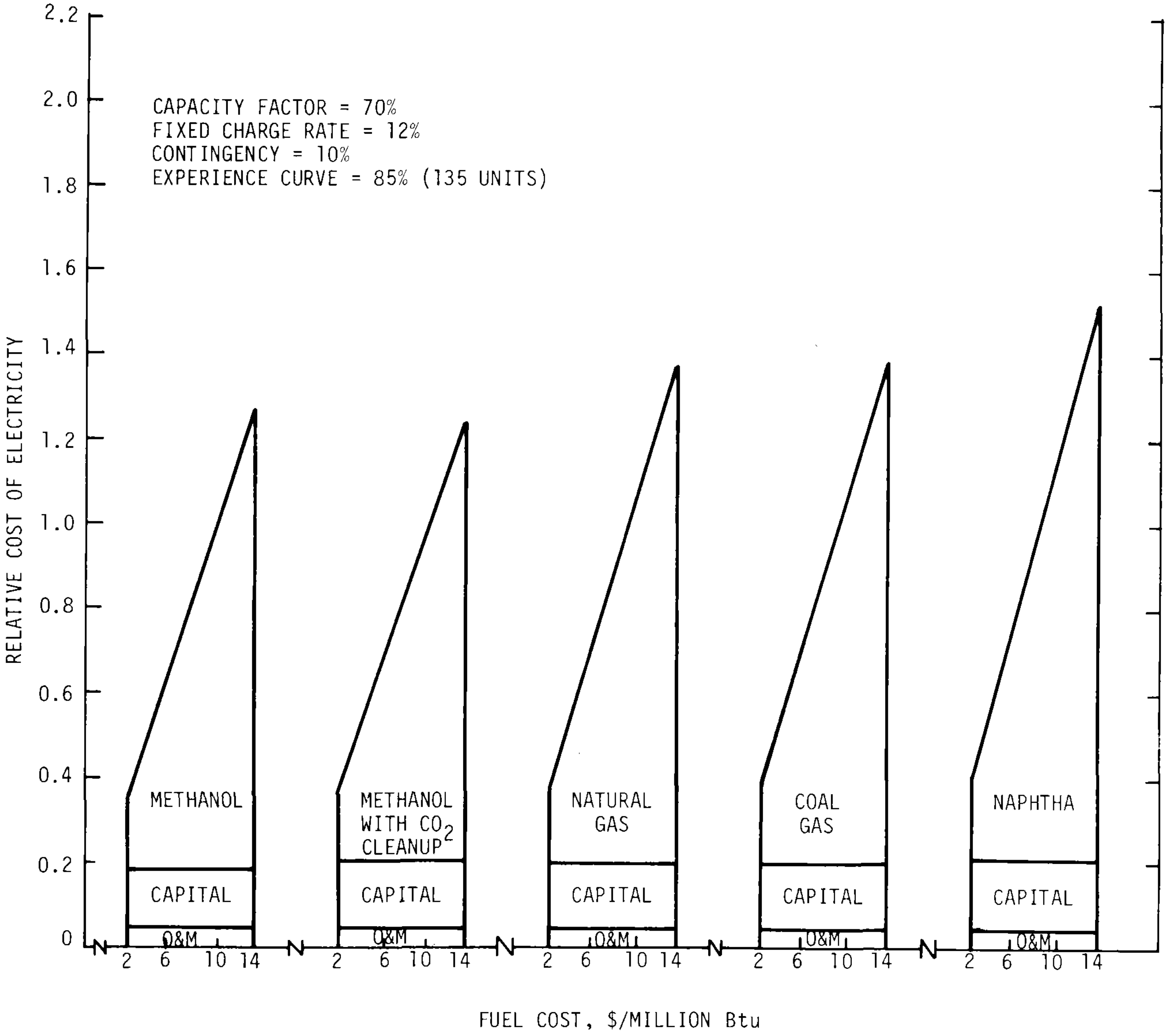

Figure 5-18. Relative COE Components for Integrated Commercial Units

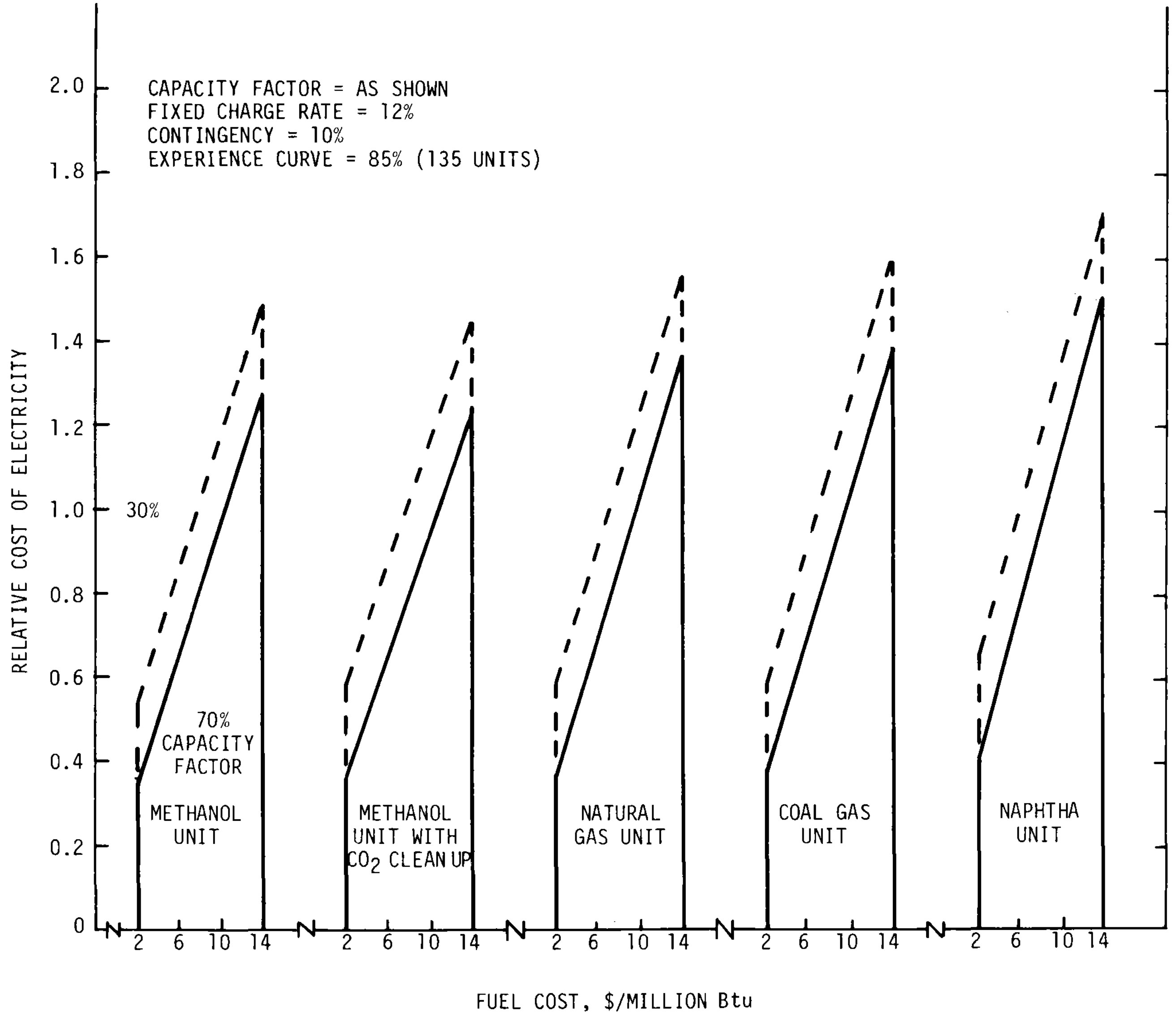

Figure 5-19. Effect of Capacity Factor on Relative COE of Integrated Commercial Units

Table 5-37 shows the subcommittee's high and low price projections based on comparisons with No. 2 fuel oil, natural gas and LNG. No. 2 prices were escalated at four percent per year in "real" dollars to 1985. The low price case beyond 1985 does not include any real price escalation. The corresponding high price case, however, does include a real dollar increase of three percent per year. 1980 natural gas/LNG prices are currently highly volatile and are only generally represented in Table 5-37. 1985 natural gas prices are based on EPRI Supply Program estimates for LNG in Los Angeles and appear to be converging on the 1985 oil prices. A fuel price range is shown for 1990 and 2000; it is expected that gas and fuel oil will be competitively priced (as will synfuel commodities) over this time period as explained above.

TABLE 5-37

DELIVERED FUEL PRICES IN

LOS ANGELES (Constant 1980 $/Million Btu)

	1980	1985	1990	2000
High Price Projection	5.79	7.04	8.16 High	10.97 High
Low Price Projection	3.11	6.14	7.04 Low	7.04 Low

The 1985 high and the 1990 low fuel cost of 7.04 $/MBtu fuel cost estimates is of particular interest. An integrated methanol unit operating with a fuel cost of 7.04 $/MBtu was therefore selected as a baseline case for comparison with the other types of units, other operating conditions, and the higher fuel cost of 8.16 $/MBtu. The COE breakdown for the integrated methanol case is:

• Fuel Cost Contribution (@ 7.04 $/MBtu)	53.5 Mills/kWh
• O & M Contribution	4.5
• Capital Investment Contribution	15.5
• Total COE	73.5 Mills/kWh

The assumptions used for the above baseline case are:

- Projected Commercial 135th unit
- 85 Percent Experience

- 12 Percent Fixed Charge Rate
- 10 Percent Contingency
- 70 Percent Capacity Factor

The effect of the different plant types, capacity factor, and fuel cost on COE compared to the above baseline case is shown in Table 5-38. The COE increments, particularly with the commercial units at 70 percent capacity factor, illustrate the COE advantage that is due to the lower heat rates that can be achieved with methanol units. This advantage is maintained at the lower 30 percent capacity factor and the high fuel cost of 8.16 $/MBtu. At equal but higher fuel cost levels, the lower heat rate methanol units will show a greater incremental COE advantage. It should be noted that heat rate for each fuel and plant configuration was assumed constant and did not vary with capacity factor. A detailed yearly load profile as a function of capacity factor would be necessary to more precisely determine relative COE under part load conditions. Since, in general, PAFC plant efficiency tends to increase as loading is decreased, this COE data can only be used to compare PAFC plant configurations at different capacity factors. They cannot be used to compare PAFC plants to alternate generation units that decrease in efficiency when loading (or capacity factor) is reduced.

5.3.4 BREAKEVEN FUEL COSTS

The relative COE for integrated and nonintegrated 50th and commercial units as a function of fuel cost is shown in Figure 5-20 through 5-23. Breakeven fuel costs, to yield the same relative COE from each type of unit, can be derived from these Figures. For the projected commercial units the approximate breakeven fuel costs are as follows:

	Non Integrated Commercial Units		Integrated Commercial Units	
Fuel/Plant Type	$/MBtu	Normalized*	$/MBtu	Normalized*
Methanol	7.04	1.08	7.04	1.08
Methanol/with CO_2 clean up	NA	NA	7.15	1.13
Natural Gas	6.50	1.00	6.30	1.00
Coal Gas	6.90	1.06	6.15	0.98
Naphtha	5.65	0.87	5.60	0.89
Ethanol	6.80	1.05	NA	NA

*Normalized to the indicated natural gas cost.

Table 5-38

INCREMENTAL EFFECT ON COE* OF UNIT TYPE, CAPACITY FACTOR, AND FUEL COSTS

FUEL	PLANT TYPE	50TH UNIT @ 70% CAPACITY FACTOR	COMMERCIAL UNIT @ 70% CAPACITY FACTOR	COMMERCIAL UNIT @ 30% CAPACITY FACTOR	COMMERCIAL UNIT @ 70% CAPACITY FACTOR & HIGH FUEL COST
METHANOL	Non Integrated	+15	+ 6	+23	+16
	Integrated	+10	0	+20	+ 9
	Highly Integrated With CO_2 Clean Up	+11	- 1	+22	+ 7
NATURAL GAS	Non Integrated	+20	+10	+29	+21
	Integrated	+19	+ 6	+27	+15
COAL GAS	Non Integrated	+15	+ 7	+22	+17
	Integrated With CO_2 Clean Up	+18	+ 7	+28	+17
NAPHTHA	Non Integrated	+32	+21	+40	+33
	Integrated	+25	+14	+35	+24
ETHANOL	Non Integrated	+17	+ 8	+26	+18
Fuel Cost, $/MBtu		7.04	7.04	7.04	8.16

Fixed Charge Rate = 12%, Contingency = 10%, Experience Curve = 90% for 50th Units, 85% for Commercial Units

*COE Increment Above (+) or Below (-) the Integrated Methanol Unit at 70 Percent Capacity Factor and a Fuel Cost of 7.04 $/MBtu

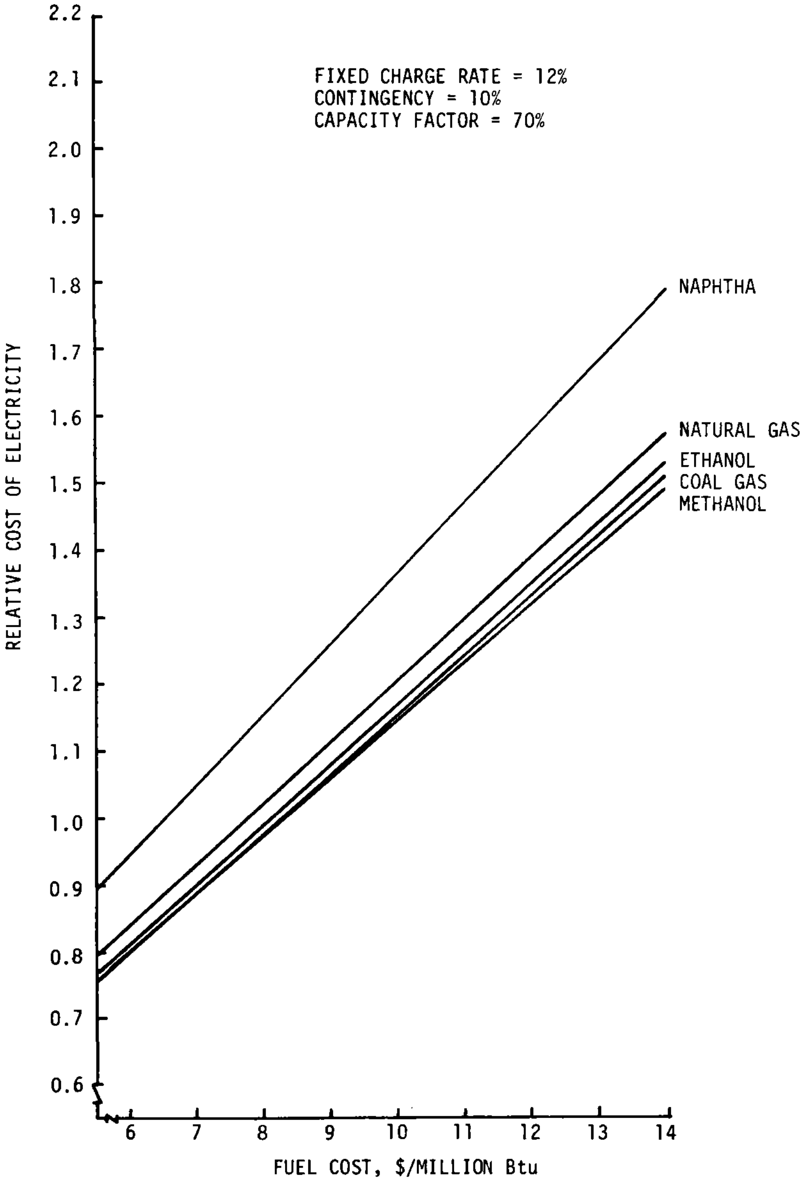

Figure 5-20. Relative Cost of Electricity as a Function of Fuel Cost for Nonintegrated 50th Units

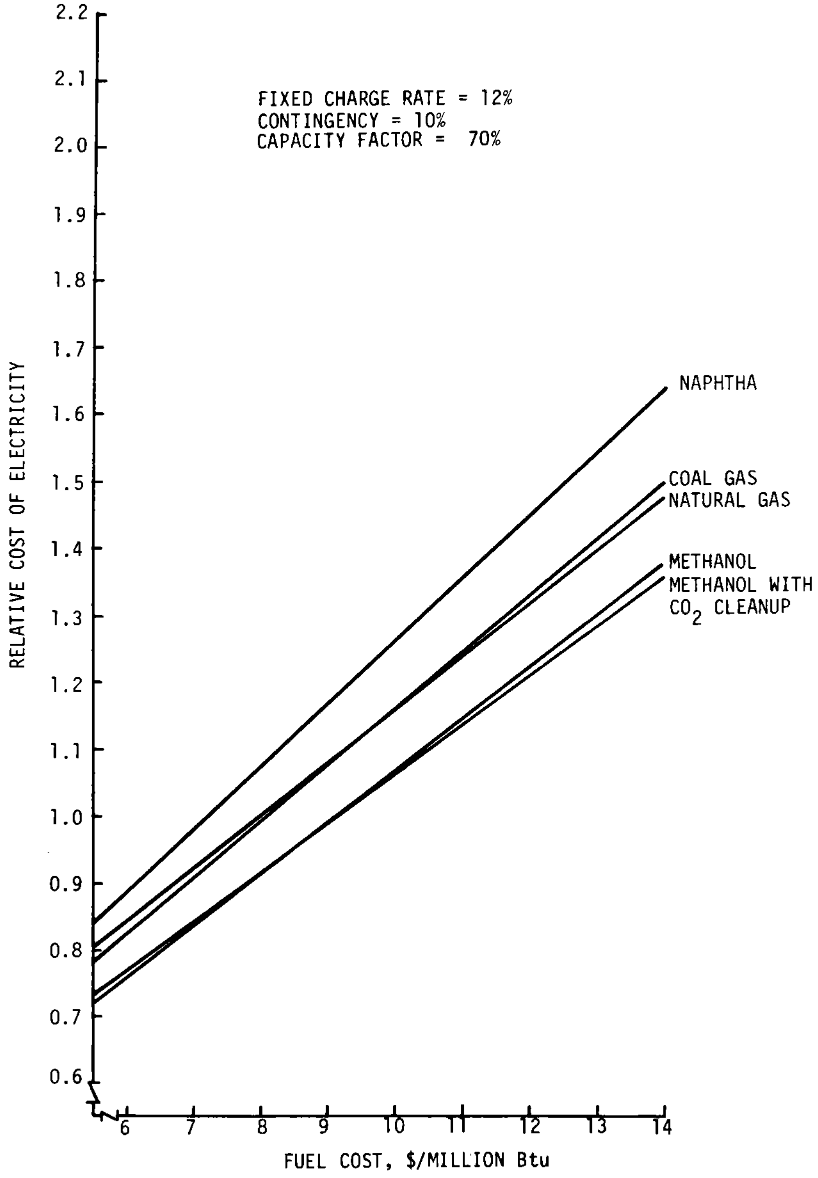

Figure 5-21. Relative Cost of Electricity as a Function of Fuel Cost for Integrated 50th Units

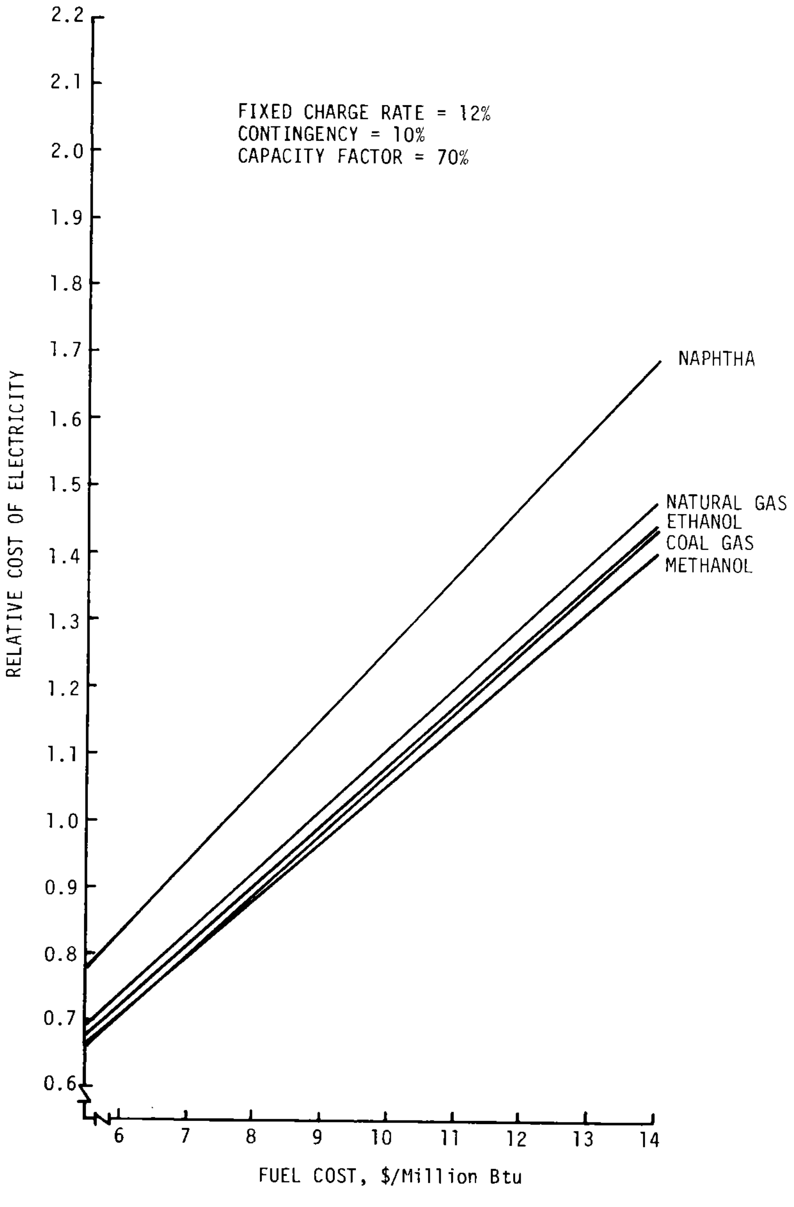

Figure 5-22. Relative Cost of Electricity as a Function of Fuel Cost for Nonintegrated Commercial Units

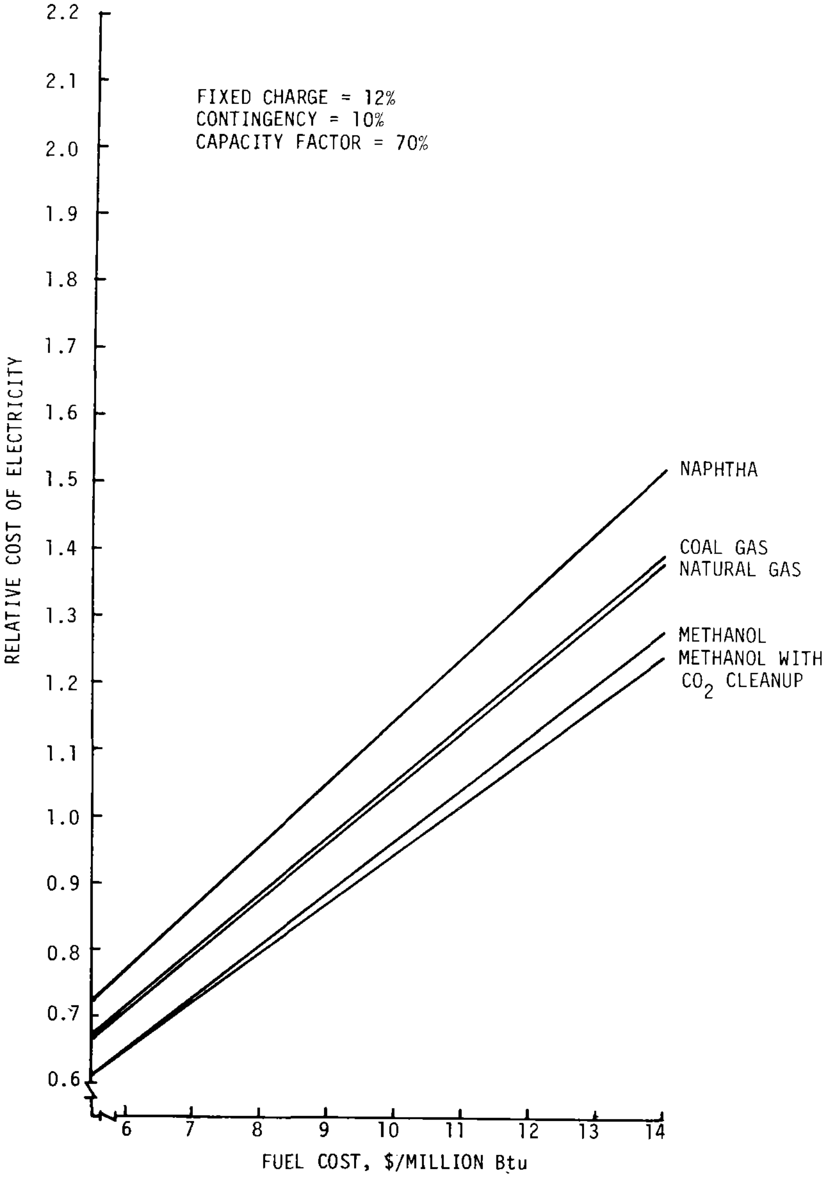

Figure 5-23. Relative Cost of Electricity as a Function of Fuel Cost for Integrated Commercial Units

The above data indicate that, for nonintegrated commercial units, an eight percent premium can be paid for methanol while still generating electricity at the same COE as the natural gas unit. A premium of five to six percent can be paid for ethanol and coal gas. Naphtha will have to be priced significantly below natural gas to yield the same COE. For the integrated commercial units the methanol premium is 8 - 13 percent. Coal gas must be priced approximately two percent below natural gas to yield the same COE, but this small differential will most likely be overshadowed by other considerations. As in the nonintegrated case, naphtha must be priced significantly below natural gas to yield the same COE.

As with the types of generation units which use gaseous or liquid fuels, the fuel cost component will be the dominant factor in the COE of PAFC power plants. Figure 5-24 shows the COE for the methanol integrated commercial unit as a function of direct capital costs for several levels of fuel cost. At a fuel cost of 7.04 $/MBtu, the COE of the integrated methanol unit will still be 59 mills/kWh* even if the direct capital cost of the PAFC plant is zero.

5.3.5 RELATIVE COE ESTIMATE UNCERTAINTIES

Based on the information provided by the fuel processing system (FPS) vendors and previous Westinghouse purchasing experience, it is estimated that the cost uncertainty range of an FPS of a given design is ± 30 percent. This estimate assumes that a detailed specification for the FPS is used and that a competitive situation exists among the potential FPS vendors. When a FPS cost uncertainty of ± 30 percent is included in the total plant direct cost and the resulting COE uncertainty range calculated, the results are as shown in Figure 5-25 for 50th units with a fuel cost of 7.04 $/MBtu.

For integrated 50th units at a 70 percent capacity factor, there is very little overlap between the COE uncertainty bands for the methanol units and natural gas. The conclusion is, that even with a ± 30 percent uncertainty in the FPS capital cost, the COE of the methanol units will be lower than that of the natural gas, coal gas, or naphtha units. At a 30 percent capacity factor, the COE uncertainty bands show considerable overlap between the natural gas, coal gas, and the methanol unit with CO_2 clean up.

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*Relative COE of 0.59

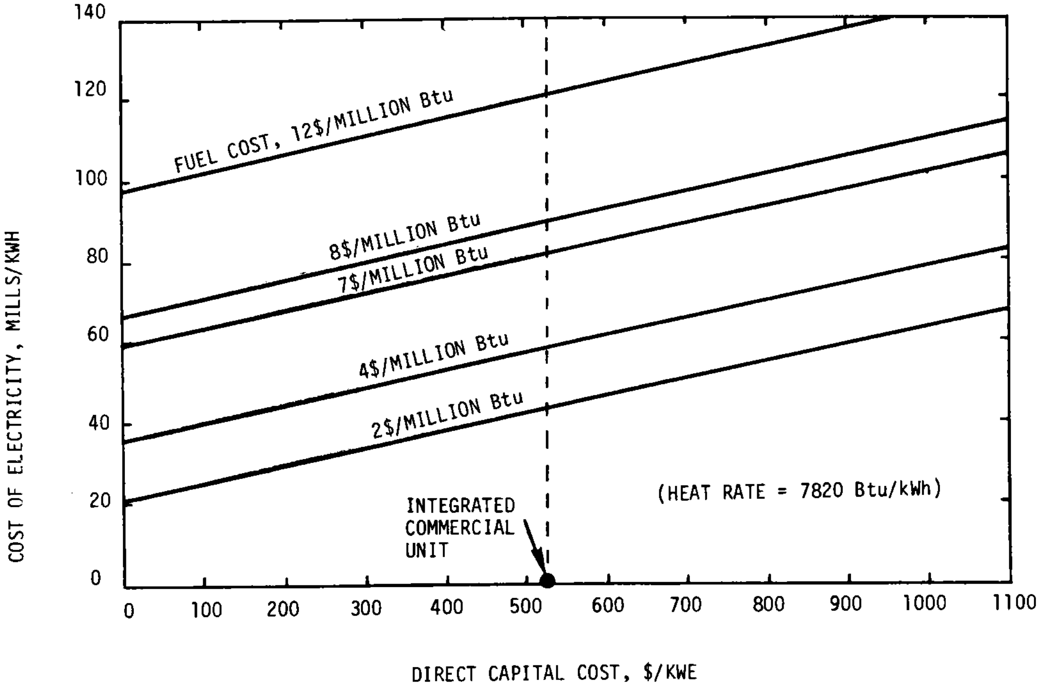

Figure 5-24. Cost of Electricity Versus Direct Capital Cost for Integrated Methanol Commercial Unit

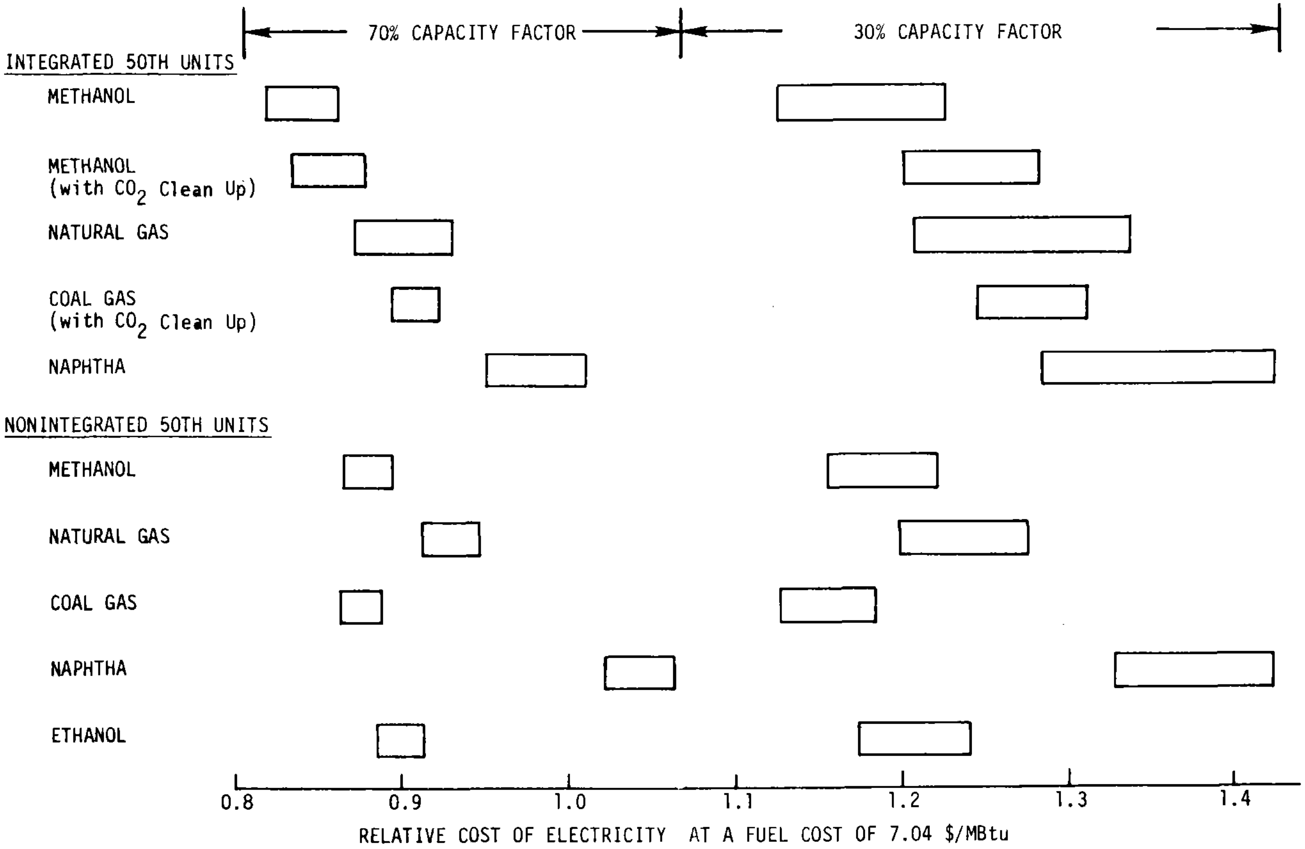

Figure 5-25. Effect of ± 30% Fuel Processing System Cost Uncertainty

When the nonintegrated units at 70% capacity factor are considered, there is no overlap between natural gas and the methanol, coal gas and ethanol units which yield a lower COE. The nonintegrated coal gas unit also shows a significant COE advantage over the natural gas unit at 30 percent capacity factor.

5.3.6 COST OF ELECTRICITY SENSITIVITY

The sensitivity of the cost of electricity to changes in several parameters was investigated for one "base case" which was the integrated 50th unit using natural gas as the source fuel. This base case is defined as follows:

Parameter		Base Case Value
Direct Capital Cost (50th Unit)	-	1003 \$/kWE
Capacity Factor	-	70%
Heat Rate	-	8440 Btu/kWh
Fuel Cost	-	7.04 \$/Million Btu
Experience (50th Unit)	-	90%
Fixed Charge Rate	-	18%
Contingency (50th Unit)	-	10%
Interest During Construction	-	8%/Year

Each one of the above parameters was varied individually and the effect on COE observed. The effect on the base case COE as the parameters are varied is shown in Figure 5-26. The COE effect of the parameters in order of decreasing sensitivity is as follows:

Experience Curve Effects (over a production run of units)
Capacity Factor (of a specific unit)
Heat Rate or Fuel Cost (of a specific unit)
Fixed Charge Rate or Direct Capital Cost (of a specific unit)
Interest During Construction (of a specific unit)
Contingency (of a specific unit)

While the relative sensitivity of the above parameters may change somewhat if a different base case were selected, certain qualitative conclusions have resulted from this analysis. First, it is clear that changes in either the contingency allowance or the interest during construction have very little impact on COE. The fixed charge rate and the fuel cost can each significantly affect COE but they are economic parameters not under the control of a fuel cell power plant designer.

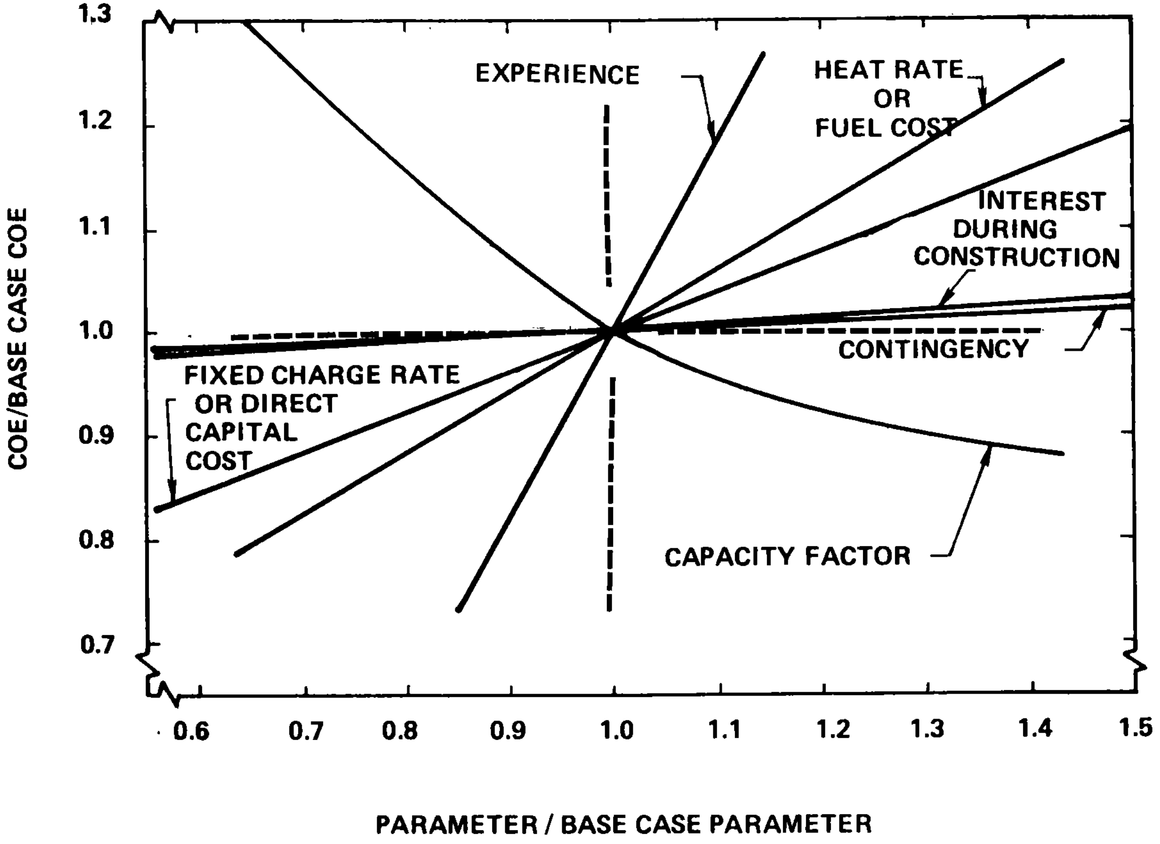

Figure 5-26. Cost of Electricity Sensitivity

Capacity factor, assuming it is not limited by an extremely high forced outage rate, will be determined by the utility owner's operating requirements. Therefore, there are only three parameters (heat rate, direct capital cost, and experience) affecting COE that are controllable to a certain extent by the designer. But there is a trade off between heat rate and direct capital cost that constrains the designer. Since COE is more sensitive to heat rate than to direct capital cost, the conclusion is that there is an incentive for improving the heat rate even at the expense of increased capital cost. However, COE is most sensitive to the experience curve factor when a run of production units is considered. The sensitivity of the COE to the experience curve factor reflects the fact that consistent achievement of even small experience curve benefits over a production run is more important than a one-time reduction in the capital cost of a specific unit.

Therefore production and manufacturing planning must be an inherent and early part of the fuel cell power plant design process if experience curve benefits are to be achieved in actual practice. The basic designs of the systems and components should be developed in a manner that maximizes benefits due to the following:

- Modular designs with in-shop manufacturing assembly, and checkout.
- Components and materials selection based on the potential for cost savings due to volume purchasing.
- Automated and standardized factory tests and checkout.
- Standardized equipment and procedures for on-site assembly and checkout.
- Standardized manufacturing processes and tooling specifically designed for volume production.
- Design features that minimize on-site construction labor and material costs.
- Standardized equipment, subsystem and system sizes.
- An effective design control and documentation process that allows staged improvements when justified but which avoids ad hoc or frequent design changes.

5.4 RANKING OF FUEL CANDIDATES

Table 5-39 summarizes the characteristics of PAFC power plant configuration for the fuel alternatives and for variations in the thermal integration of the FPS. Included in this table are operational factors, performance factors, capital costs and incremental COE. This data is presented based on the following assumptions:

Table 5-39

COMPARISON OF FUEL ALTERNATIVES FOR THE PAFC POWER PLANT
(At a 70% capacity factor and a fuel cost of 7.04 $/million Btu)

		FUEL PROCESSING SYSTEM								
		REFORMER TEMPERATURE (°F)	TECHNOLOGY STATUS	OPERATIONAL FACTORS	PROTOTYPE SYSTEM DIRECT CAPITAL COST $ MILLIONS	HEAT RATE (BTU/KWHR)	NET PLANT POWER OUTPUT MW_e (AC)	PROTOTYPE DIRECT CAPITAL COST ($/KW)	PROJECTED COMMERCIAL DIRECT CAPITAL COST $/KW	INCREMENTAL COST OF ELECTRICITY Mills/KWH
METHANOL	**Nonintegrated**	500-1000	Near mid-term/ developmental.	More rapid than natural gas.	2.2	8790	6.95	1500	480	+6
	Integrated	500-1000	Less complex than natural gas.	High availability and reliability.	3.3	7820	6.92	1670	530	0
	Integrated with CO_2 removal	500-1000	Same		3.3	7390	6.71	1940	610	-1
NATURAL GAS	**Nonintegrated**	1600-1800	Near term/ commercial.	Very slow response. Massive complex system.	2.9	9270	6.95	1600	510	+10
	Integrated	1600-1800	Very complex system.	Lowest availability.	4.4	8440	6.95	1820	580	+6
COAL GAS	**Nonintegrated**	None	Near term.	Fastest response. Least problems visualized.	2.2	9110	8.1	1330	420	+7
	Integrated with CO_2 removal	None	Least complex.	Highest availability.	2.2	8520	6.95	1840	580	+7
NAPTHA	**Nonintegrated**	1600-1800	Near term/ commercial.	Slow response.	3.2	10750	6.41	1670	530	+21
	Integrated	1000-1800	Most complex.		4.8	9450	6.94	1870	590	+14
ETHANOL	**Nonintegrated**	500-1000	Developmental	Similar to methanol.	2.2	9060	6.95	1520	480	+8

- 7.5 MWe (DC) fuel cell power
- projected commerical unit
- 1990 start date with low fuel price of 7.04 $/Million Btu
- 70% capacity factor

Comparison of the alternate plant configurations illustrate several of the trade offs in performance, capital costs, and cost of electricity.

5.4.1 PERFORMANCE

Comparison of the five fuels in the nonintegrated configuration indicates that methanol has the lowest heat rate followed closely by ethanol, coal gas and natural gas. The total span of difference between these four is five percent which is relatively close. The heat rate for Naphtha is considerably greater than the others (17-22 percent). It should be pointed out that for the Naphtha case a 4/1 steam/carbon ratio was used (as compared with 2.3/1 for natural gas) based on recommendation by vendors and current commercial practice for existing hydrogen plants. Dependent on the operating temperatures, selection of catalyst, etc. there is a potential for reducing this to 3/1. At this ratio the heat rate for Naphtha would be reduced to 10,000 Btu/kWh for the nonintegrated configuration which would still be 14 percent greater than methanol.

Integrating the systems result in the potential for reducing the heat rate for all the fuels by 600-1400 Btu/kWh (6-16 percent). In an integrated configuration, methanol can have a significantly lower heat rate (600-1000 Btu/kWh than the other fuels. On a percent basis this is 12 percent lower.

This is attributed to the lower reforming temperature in combination with the latter thermal integration offering essentially a constant temperature operation (350-750°F) throughout the fuel processing/power generation cycle. The basis for this is the relatively low steam/carbon ratio (1.6) required for the methanol purity considered. In an integrated configuration ethanol, natural gas and coal gas have essential the same heat rates of 8450-8500 Btu/kWh. The heat rate for the Naphtha, however, was 9500 Btu/kWh which is another 1000 Btu/kWh greater than the second cluster of fuels. For a steam/carbon ratio of 3/1 this can be reduced by 500 Btu/kWh. It should also be pointed out that the above analyses were performed for a fuel cell temperature of 350°F and for a dry cooling tower. By raising the fuel cell temperature to 375°F the heat rates for all fuels can be reduced by roughly 300-400 Btu/kWh. In addition, use of a wet cooling tower can reduce the plant heat rate by 250 Btu/kWh. On the basis of plant heat rate, therefore the fuels can be ranked as follows:

1. Methanol
2. Toss up between ethanol, natural gas, and coal derived gas.
3. Naphtha

5.4.2 TECHNOLOGY STATUS

For this study differences in technology status were assessed only in respect to the Fuel Processing system. Response from the four vendors indicated that all four have commercial experience with processing of natural gas and naphtha for hydrogen production. Limitations and design considerations are well understood with these two fuels. This does not, however, preclude a significant design effort to tailor the fuel processing system to a specific PAFC power plant. This kind of design effort could, of course, be included in the development of projected commercial PAFC power plants. An FPS for this application would be specifically designed to meet the transient, startup/shutdown, modularity, and other requirements of PAFC power plants and would be significantly different from current FPS technology. However, even an FPS specifically for PAFC power plant applications would be influenced by the inherent characteristics of the fuel that is being reformed. Therefore, the following discussion based on current FPS technology will still provide a qualitative assessment of the effect of different fuels.

The four vendors indicated that they did not have direct commercial experience with methanol, ethanol, and coal gas. It was their opinion, however, that since methanol and ethanol can be reformed at lower temperatures, that these systems could be developed in the near term. Furthermore, some methanol reforming work is ongoing for smaller scale systems by other organizations.

The question of the schedule and cost needed to develop a methanol and ethanol fuel processing system was beyond the scope of this study; however, follow up discussion with these potential commercial suppliers should be carried out to better assess the technology of methanol and ethanol. Since low methane coal gas will require shift conversion but not reforming, the technology status of processing coal gas is thought to be near term and of lower risk than methanol and ethanol. Based on the above considerations, relative ranking of the five systems on technology status is:

1. Natural Gas and Naphtha
3. Coal Gas
4. Methanol and Ethanol

5.4.3 OPERATIONAL CONSIDERATIONS

Since low methane coal gas will only require intermediate and low temperature shift conversion the fuel processing system will be less complex, smaller and have much less thermal capacitance than any of other fuel processing systems in either the integrated or nonintegrated configurations. Start up and shut down of a plant with a coal gas fuel processor should be much faster than with any of the other systems. The amount of catalyst required for shift conversion only should be less than the others; therefore, maintenance time and cost should be reduced. Maintenance time and costs will also be less due to having fewer components. The environmental effect of fuel processing maintenance relative to total plant maintenance was not addressed in this study; therefore, the magnitude or importance of this reduced maintenance effort can not be ascertained but should be considered in further studies.

Commercial natural gas fuel processing systems have relatively thick and long tubes in their reformers. Furthermore the high temperature reforming process dictates the need for more heat exchange equipment. The start up and shut down times for commercial natural gas reformers are quite long (8-24 hours) which precludes consideration of rapid response peaking applications given current commercially available technology. It is perceived that these fuel processors will be more sluggish than for methanol, ethanol, and coal gas. Because of the added equipment and high temperature operation it is anticipated that maintenance costs will also be greater.

Naphtha reformers should be similar to natural gas reformers. Because of the need to have a higher steam/carbon ratio to prevent coking due to the presence of larger amounts of aromatics the fuel processor will be more massive than for natural gas. Furthermore degradation of the high temperature catalysts due to coking will be more severe thus adding to maintenance time and costs.

Response time and maintenance for ethanol and methanol are judged to be intermediate to coal gas and natural gas based soley on the fact that lower temperature reforming should result in a simpler, less massive system than that for natural gas. Thus response times and maintenance should be less severe.

A relative ranking of the five systems on operational considerations is therefore:

1. Coal Gas
2. Methanol and Ethanol
4. Natural Gas
5. Naphtha

5.4.4 ECONOMIC CONSIDERATIONS

From the designer standpoint the primary considerations in the PAFC plant economic performance are:

- Heat Rate
- Direct Capital Cost
- Fuel Cost
- Experience Curve Factor

Fuel cost is not controllable by the designer, but the level of the fuel cost does affect the trade off between lower heat rate and higher capital cost.

The design and manufacturing methods necessary to achieve experience curve benefits are equally applicable to all of the source fuel-plant configurations examined. If such benefits are achieved in practice, the capital cost contribution to the COE of a projected commercial unit will be about 16-21 percent of the total COE. The contribution to total COE due to fuel cost will be much higher. When one considers the projected commercial units, there is a strong economic incentive to design for the lowest possible heat rate even at the expense of higher capital costs. While the methods of achieving low capital costs from experience curve benefits are generally applicable, achievement of such benefits in actual practice will be relatively more difficult with the integrated natural gas and naphtha fuel processor systems due to their increased complexity. The above considerations suggest that the methanol, coal gas, and ethanol units are preferable to the natural gas and naphtha units on the basis of actual experience curve benefits factors and the direct capital cost of projected commercial units.

Consideration of the estimated COE of the alternate unit configurations, (with the assumption of equal fuel costs) shows that methanol, coal gas, and ethanol will retain a COE advantage over natural gas and naphtha even when uncertainties in fuel processor costs are considered as in Section 5.3.5. There is some overlap between the estimated COE's when the integrated natural gas unit is compared with the nonintegrated ethanol unit at different fuel cost capacity factors and stage of commercial development. In all cases the unit configurations with the lowest heat rates (methanol, integrated natural gas, and integrated coal gas units) will have an increasing COE advantage over the other configurations if fuel costs increase beyond 7-8 $/Million Btu.

On the basis of this economic analysis the potential fuels for the commercial PAFC unit are ranked as follows in order of decreasing preference:

- Methanol
- Coal Gas (or Natural Gas)
- Ethanol
- Naphtha

This analysis suggests that methanol should be considered the primary liquid fuel with ethanol as the alternate liquid fuel. Coal gas should be the primary gaseous fuel with natural gas as the alternative.

5.4.5 SELECTION OF REFERENCE FUELS

The ranking of fuels by criteria is summarized in Table 5-40. Of the three liquid fuels considered, methanol is preferred over ethanol and naphtha. Comparison of methanol versus ethanol indicates that it is a toss up in terms of technology and operation factors; however, methanol should have a lower heat rate and COE. Naphtha rates over methanol in technology status; however, its projected heat rate and COE are considerably poorer.

Natural gas and coal gas rate equally in terms of performance and COE. The fuel process technology is more mature than coal gas; however, operationally coal gas should be superior. Ultimately coal gas should be a preferred gas over natural gas. For near term applications the supply of coal gas is tied to the development and construction of large coal gasification plants and the associated risks of that development technology reaching commercial maturity. Therefore for near term application, natural gas is preferred over coal gas and vice versa for far term applications.

Table 5-40

RANKING OF FUELS BY HEAT RATE, TECHNOLOGY, OPERATION, AND COE CRITERIA

RANK	HEAT RATE	TECHNOLOGY	OPERATION	COE
1st	Methanol	Natural Gas, Naphtha	Coal Gas	Methanol
2nd	Ethanol, Natural Gas, Coal Gas		Methanol, Ethanol	Ethanol, Natural Gas, Coal Gas
3rd		Coal Gas		
4th		Methanol, Ethanol	Natural Gas	
5th	Naphtha		Naphtha	Naphtha

Methanol is preferred over natural gas and coal gas due to slight advantages in performance and COE. Its technology status is less certain than natural gas and the production of methanol in large quantities is also tied to producing methanol from coal which is not mature. Therefore in near term applications natural gas is preferred over methanol. Although operationally coal gas may be better than methanol, the performance and cost advantages particularly in an integrated plant, favor methanol. The following recommendations are therefore proposed.

	Near Term Applications	Future Commercial Applications
Primary Fuel	Natural Gas	Methanol
Alternate Fuel	Methanol	Coal Gas

One of the factors not considered in this study is multifuel capability of the Fuel Processor System. Obviously natural gas and naphtha have similar fuel processor systems whereas natural gas and methanol fuel processors may be significantly different. This may also be true for methanol and coal gas. The impact of multifuel capability on fuel processor designs also needs to be considered in future studies.

6.0 CONCLUSIONS AND RECOMMENDATIONS

The major conclusions from this study are as follows:

- Based on considerations of heat rate, COE, and operational factors, methanol is selected as the preferred fuel for future commercial PAFC plant applications with coal gas as the second choice.
- For near term applications where current technology status is an important consideration, natural gas is the preferred fuel with methanol as the second choice.
- While there is extensive commercial experience with fuel processors which convert naphtha and natural gas into hydrogen, the complexity of an FPS for either naphtha or natural gas will make it relatively more difficult to design a commercial utility unit with the desired operation and maintenance characteristics.
- Experience curve benefits in a production run of commercial units will be important in minimizing capital costs and COE. Production and manufacturing planning must be an inherent and early part of the fuel processing system and complete PAFC power plant design process in order to achieve such benefits in actual practice.
- The high cost of the CO_2 clean up system tends to offset the COE and heat rate advantages of the extensively integrated methanol system when fuel costs are low. At the highest fuel costs the integrated methanol system with CO_2 clean up will yield the lowest COE.
- The fuel cost alone is by far the largest component in the COE of all the units considered. For this reason, it is significant that methanol units have the potential for the lowest heat rates. A heat rate in the range of 7000 Btu/kWh is a credible objective with methanol.

The major recommendations resulting from this study are as follows:

- The effects of PAFC plant operating pressures and temperatures on plant heat rate at full and part-load conditions and costs should be evaluated. The degree of technical risk associated with projected improvements in performance and costs should also be evaluated.

- Operability and performance features such as varying degrees of part-load, transient, and startup/shutdown capabilities should be examined in detail.

- Less expensive methods of CO_2 removal and resulting effects on heat rate and complexity, should be examined.

- A conceptual design and cost estimate of a fuel processing system that is integrated thermally with a PAFC plant should be developed with close attention paid to negative affects on availability due to more difficult maintenance.

- The cost, performance, operational, and COE impacts of adding multifuel capabilities to the fuel processing system needs to be considered in a further study.

- A study of multifuel capability should include the cost, performance, operational and COE aspects for each potential fuel that are described herein. The future availability of the potential fuels and utility-owner fuel preference should be inputs to such a future study. An evaluation of current projections for the future availability of methanol and coal gas should also be considered.

- The developmental costs and technical risks associated with bringing methanol and coal gas fuel processing systems from their present status to that required for commercial PAFC plants should be evaluated through appropriate vendors.

The importance of achieving experience curve benefits and successful commercialization of the PAFC plant also lead to the following recommendations:

- Standard sizes for the major systems of the PAFC power plant should be defined early to permit manufacturing and production planning for volume production to become an early and inherent part of the design process.

- Overall power plant specifications should be developed early in the design process to: facilitate production and manufacturing planning; to allow the selection of standard sizes; to permit the maximum use of common components; and to allow the definition of modular, factory assembled systems.

- The PAFC power plant specifications should be complete and comparable to the specifications used by utilities for the purchase of alternate types of generating units that will be competing with PAFC power plants. Such specifications should be carefully prepared in order to avoid imposing performance or other requirements on fuel cell power plants that either exceed those imposed on competing generating units or unnecessarily increase PAFC power plant cost or complexity.

- Conversely, the PAFC specifications should emphasize the unique benefits of fuel cells such as low part-load heat rates, low emissions, modularity, and siting flexibility.

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APPENDIX A

FUEL PROCESSOR SPECIFICATION (PRELIMINARY)

FUEL PROCESSOR SPECIFICATION (PRELIMINARY)

For Fuel Cell Power Plants
in Electric Utility Applications

I. INTRODUCTION

Westinghouse is developing Fuel Cell Power Plants for electrical power generation at dispersed utility company sites. These power plants will incorporate a fuel processor subsystem (FPS) which converts a source fuel into a hydrogen rich gas stream that feeds the fuel cell power subsystem. The candidate source fuels can potentially include methanol, methane (natural gas), synthesis gas from off-site coal processing plants, petroleum derivatives, and other fuels.

Early in 1980, Westinghouse corresponded with suppliers of hydrogen generations systems concerning the application of such systems to fuel cell power plants. As a result, a number of hydrogen generation system suppliers with appropriate capabilities and an interest in participating in the development of such applications were identified.

One of the concerns of potential utility users is the effect of different types of source fuels on fuel cell power plants. Accordingly, Westinghouse has been asked to assess the effects of alternative source fuels on the economics and performance of fuel cell power plants. This assessment has the objective of selecting reference source fuel(s) and identifying backup fuel(s) prior to further design of a commercial utility fuel cell power plant. The future availability of alternative source fuels is also of interest to utility users. However, this aspect is being treated separately and availability questions should not influence the objectivity of the FPS performance/cost assessments that are addressed herein.

In order to perform the above assessment, Westinghouse is asking potential suppliers to provide the following type of information for each source fuel of interest:

1) The price of the fuel processor subsystem (FPS).

2) Operation and maintenance cost estimates for the FPS. The fraction of the annual O&M cost due to the replacement of catalysts, absorbents, etc., should be estimated.

3) The utility requirements (feed water, electrical power, etc.) to operate the FPS.

4) An indication of the FPS technology status.

5) The impact of designing the FPS with multiple source fuel capability.

6) Other information as indicated in Table III-1.

Westinghouse will then combine the above information with the fuel cell power subsystem performance, the capital cost of the fuel cell subsystem and auxiliary systems, and other plant subsystem O&M requirements in order to compute the overall fuel cell power plant heat rate, capital cost, and cost of electricity produced as a function of each source fuel type. Consistent adherence to the same set of assumptions and specifications is necessary in order to arrive at an objective comparison of the effects of alternate source fuels on the fuel cell power plant performance and economics.

The FPS Requirements are outlined in Section II. The format of the requested information is given in Section III.

II. REQUIREMENTS

1.0 FUEL PROCESSOR SUBSYSTEM

1.1 FUNCTIONAL REQUIREMENTS

The Fuel Processor Subsystem (FPS) shall convert the specified source fuel(s) into a hydrogen rich feedstream for the fuel cell anode.

A high efficiency FPS design is preferred in order to maximize the overall efficiency of the fuel cell power plant. Therefore it is desirable that:

- the total source fuel input to the FPS per unit of hydrogen output be minimized.
- the export steam from the FPS be minimized to the extent possible given standard FPS industry practices.

Given the type and nature of the source fuel, the FPS shall perform source fuel cleanup including desulfurization if required.

Cooling water, boiler feed water, electrical power, control air, and inerting gas (if required) will be supplied to the FPS at rates and conditions to be specified by the FPS designer.

1.2 PROCESS REQUIREMENTS

1.2.1 FUEL CELL ANODE FEED STREAM

The FPS will produce a hydrogen rich anode feed stream for the fuel cell at the following rates and conditions:

H_2 production rate = 227,000 SCFH*
Feed Stream Pressure = 200 psia
Feed Stream Temperature = 300 - 400° F

The anode feed stream hydrogen content shall be a minimum of 50 percent by volume.

The diluent content of the anode feed stream shall not exceed 50 percent by volume. Diluents may include CO_2, N_2, He, Ar, H_2O, and CH_4. These diluents are inert in terms of their effect of fuel cell operation and are not regarded as impurities.

Impurities which have an adverse effect on fuel cell operation shall be limited to the following maximum allowable values:

*SCFH (Standard Cubic Feet per Hour) as measured at 60° F and 760 MM of Mercury

IMPURITY		MAXIMUM ALLOWABLE
CO	-	1.0% by volume
H_2S	-	100 PPM by volume
COS	-	100 PPM by volume
C_2^+	-	100 PPM by volume
NH_3	-	1 PPM by volume
Cl^-	-	1 PPM by volume
Metal Ions	-	Less than 1 PPM by weight

The FPS designer is asked to indicate if the above impurity levels have a significant impact on the FPS price or configuration.

With the exception of the above impurities, an extremely pure hydrogen product is not needed for the fuel cell anode feed stream.

A hydrogen product purity level as low as 50 percent by volume is acceptable for fuel cell operation provided the above impurity limits are observed. Since the removal of the diluents CO_2, N_2, He, Ar, H_2O and CH_4 is not necessary to product on a high hydrogen product purity level, the FPS design complexity and price should be reduced by eliminating unnecessary absorption and purification features.

1.2.2 FUEL CELL ANODE EXHAUST STREAM

The fuel cell utilizes about 85 percent (by mass) of the hydrogen contained in the anode feed stream. Therefore, nominal anode exhaust stream conditions as compared with the input anode feed stream conditions are as follows:

	ANODE FEED STREAM (FROM FPS)	ANODE EXHAUST STREAM (RETURN TO FPS)
H_2 Contained in Feed or Exhaust Stream, SCFH	227,000	34,000
Pressure, psia	200	150
Temperature, ° F	300-400	350-425

Note that in addition to hydrogen, both the feed and exhaust streams contain the diluents (CO_2, N_2, He, Ar, H_2O, CH_4) which are unchanged by passage through fuel cell anode. For purposes of illustration, representative compositions of the anode feed/exhaust streams could be as follows:

	ANODE FEED STREAM	ANODE EXHAUST STREAM
Mole Fraction		
H_2	0.726	0.346
H_2O	0.079	0.189
CO	0.010	0.023
CO_2	0.174	0.415
N_2	0.001	0.004
CH_4	0.010	0.023

The fuel cell anode exhaust stream should be returned to the FPS with the objective of achieving high FPS efficiency. Methods for using the anode exhaust stream in the FPS are FPS designer options.

1.3 SOURCE FUEL(S) SPECIFICATIONS

The types and compositions of source fuels under consideration for Fuel Cell Power Plant applications are listed in Table II-1. Further general information on fuels is given in Attachment I.

TABLE II - 1, FUEL CELL POWER PLANT SOURCE FUEL COMPOSITIONS (VOLUME PERCENT RANGES)

	COMPOSITION						
Type of Source Fuel	CH_4	C_2^+	H_2	CO	CO_2	N_2	Notes
1. Natural Gas (Includes Natural Gas, SNG and LNG)	75→100	0→18	-	-	0-2	0-17	Small amounts of Mercaptans, Methyl Sulfides or Hydrothiophenes.
2. Naphtha	—	—	Fuel Grade		—	—	—
3. Methanol	—	—	Fuel Grade		—	—	Contains 96 Vol. % Methanol, higher alcohols, acetone, formic acid and water.
4. Fuel Oil (#2)	—	—	Fuel Grade		—	—	Contains up to 8000 ppm Sulfur and 25 ppm Nitrogen
5. Medium BTU Coal Gas	0→12	—	36→40	20→53	9→28	0→1	
6. Ethanol	—	—	Fuel Grade		—	—	

III. REQUESTED INFORMATION

The information requested for each alternative source fuel and the resulting FPS configurations is summarized in Table III-1.

Alternative FPS configurations numbered 1 through 6 are to be designed for the processing of the indicated primary source fuel only. Features whose sole functions is to provide multiple source fuel capability should be deleted from these FPS configurations. The purpose of these FPS alternatives is to identify FPS price and performance differences that are caused solely by changing the selected source fuel.

Alternative FPS configurations numbered 7 through 13 are to be designed for the indicated multiple source fuel capability. The purpose of these FPS alternatives is to identify FPS price and performance differences caused by adding multiple source fuel capabilities.

In addition to the information requested in Table III-1, the FPS designer is asked to indicate if any of the requirements stated herein are unusual in terms of standard industry practices and if they significantly impact FPS price and performance.

TABLE III-1. SUMMARY OF FPS DATA REQUESTED FROM SUPPLIERS

NOTE: FPS Price and Estimates (Parameters No. 7 through No. 13) should be Based on Standard Industry Practices to the Extent Possible

		FPS ALTERNATE CONFIGURATIONS												
PARAMETER	REQUIREMENT OR ACCEPTABLE PARAMETER RANGE	#1	#2	#3	#4	#5	#6	#7	#8	#9	#10	#11	#12	#13
1. Primary Source Fuel - Supply Pressure for all Gaseous Fuels is 500 psig. See Table II-1 and Attachment 1 for Fuel Data	As Indicated per FPS Alternate	Natural Gas	Naphtha	#2 Fuel Oil	Methanol	Ethanol	Medium BTU Coal Gas	Natural Gas	Medium BTU Coal Gas	Natural Gas	Medium BTU Coal Gas	Natural Gas	Medium BTU Coal Gas	Methanol
2. Backup Source Fuel See Table II-1 and Attachment 1 for Fuel Data	As Indicated per FPS Alternate	None	None	None	None	None	None	Naphtha	Naphtha	Methanol	Methanol	#2 Fuel Oil	#2 Fuel Oil	#2 Fuel Oil
3. FPS Hydrogen Production SCFH	227,000													
4. Utilities Per 1000 SCF of Hydrogen Produced: - Total Source Fuel Feed - Source Fuel Feed Used for Combustion - Power, KWH - Cooling Water (Gal)/(°F Rise) - Boiler Feed Water - Export Steam (psig/lbs) - Other	TBD by FPS Designer													
5. FPS Price (in 1980 $'s, <u>Without</u> Escalation Reflecting Anticipated Future Delivery or Installation Dates - <u>Not</u> Including Item 6.)	TBD by FPS Designer													

TABLE III-1, SUMMARY OF FPS DATA REQUESTED FROM SUPPLIERS (Continued)

NOTE: FPS Price and Estimates (Parameters No. 7 through No. 13) should be Based on Standard Industry Practices to the Extent Possible

		FPS ALTERNATE CONFIGURATIONS												
PARAMETER	REQUIREMENT OF ACCEPTABLE PARAMETER RANGE	#1	#2	#3	#4	#5	#6	#7	#8	#9	#10	#11	#12	#13
6. Price of Initial Charge of Catalysts and Absorbents (in 1980 $'s)	TBD by FPS Designer													
7. Estimated Field Installation Cost (in 1980 $'s, Gulf Coast Basis)	TBD by FPS Designer													
8. Annual Operation and Maintenance - Total Estimated Annual O&M Cost (1980 $'s) - Estimated Annual Cost of Catalysts and Absorbents (1980 $'s) - Total Estimated O&M Labor (Hours per Year) - Estimated Schedule Outage for O&M (Hours per Year) - Major Maintenance Items	TBD by FPS Designer													
9. Estimated Useful Operating Lifetime of FPS Without Replacement of Major Components, Years	TBD by FPS Designer													

TABLE III-1, SUMMARY OF FPS DATA REQUESTED FROM SUPPLIERS (Continued)

NOTE: FPS Price and Estimates (Parameters No. 7 through No. 13) should be Based on Standard Industry Practices to the Extent Possible

		FPS ALTERNATE CONFIGURATIONS												
PARAMETER	REQUIREMENT OF ACCEPTABLE PARAMETER RANGE	#1	#2	#3	#4	#5	#6	#7	#8	#9	#10	#11	#12	#13
10. Estimated Lead Time from Receipt of Order to FPS Operation	TBD by FPS Designer													
11. Estimated Overall Envelope Size of Installed FPS.	TBD by FPS Designer													
12. Estimated Emissions from FPS: NO$_x$, LB/Million BTU SO$_x$, LB/Million BTU Particulates, LB/Million BTU	TBD by FPS Designer													
13. Estimated Operational Factors: Turn Down Capability, % Load Change Capability, %/Hour Start Up/Shut Down Time, Hours Start Up Requirements: Inerting Gas? Start Up on Source Fuel?	TBD by FPS Designer													
14. Assessment of Technical Status (Given Source Fuel and Present Status of FPS Technology) (a) "Commercial" Routine Application (b) "Near Term" - Minimal Development (c) "Developmental" - Considerable Development Required (d) "Speculative" - Extensive R&D is Required	TBD by FPS Designer, Identify as (a), (b), (c) or (d)													

ATTACHMENT I

GENERAL INFORMATION ON FUELS FOR UTILITY INTERMEDIATE AND PEAKING DUTY

Natural Gas, LNG

Natural gas is typically 90-95% methane with some heavier hydrocarbon gases, nitrogen and/or carbon dioxide. The light hydrocarbon liquids and sulfur compounds present in the gas from the well are removed at the source before transmission through the pipeline. The heating value ranges from 950-1050 Btu/SCF. A small amount of sulfur compounds are added as an odorant in the local distribution system.

The liquified natural gas (LNG) is natural gas that has been liquified by cooling for storage or oversea transportation. It is revaporized before use.

SNG

Synthetic natural gas (SNG) can be made by gasifying coal, oils, or biomass. The raw coal gas is cleaned to remove contaminants. The clean gas is converted to methane by removal of carbon dioxide followed by methanation. The industry requires that SNG be interchangeable with natural gas so it can be distributed in the present system without changing the pilots or burners in home appliances and furnaces.

Naphtha and Selected Kerosenes and Jet Fuels

A variety of light, low sulfur hydrocarbon fuels derived from petroleum that generally conform to the specification summarized in Table 1 below. The fuel can contain paraffin type compounds, and a moderate level of aromatics and unsaturated compounds. The fuel may be a simple light distillate from crude oil, one of the many solvents prepared from oil, or a byproduct (raffinate) left after the aromatics have been removed.

Table 1 - Summary of Fuel Specification

End Boiling Point	420°F	Max
Aromatics	12 Vol %	Max
Olefins	3 Vol %	Max
Sulfur	.05 wt.%	Max

Light Coal Liquids

A variety of light, low sulfur hydrocarbon fuels from coal-derived synthetic crude oil that generally conforms to the liquid fuel specification (Table 1). The synthetic crude oil may be produced from coal by direct liquefaction using the EDS, SRC, or H-coal process, or by indirect liquefaction using the Fisher-

Tropsch or M-gasoline process. The light liquid fuel is derived from the synthetic crude oil by a combination of distillation and extraction techniques with hydrotreating to reduce the nitrogen and sulfur levels.

Medium and Low Btu Gas

Medium Btu gas is produced by gasifying coal or oil. The gas, after cleaning for pipeline use, is primarily hydrogen and carbon monoxide with some carbon dioxide and methane. The heating value is 250-300 Btu/SCF. The exact composition varies with the type of coal gasifier used; entrained flow, fluidized bed or moving bed, Table 2. Low Btu gas (100 Btu/SCF) is produced by using air to gasify the fuel. The heating value is lowered due to the dilution effect of the nitrogen introduced in the air.

Table 2 - Typical Gas Compositions

	Medium Btu Gas			Low Btu Gas
Process	Entrained Flow	Fluidized Bed	Moving Bed	Entrained Flow
Composition (Vol %)				
H_2	36	42	40	12
CO	53	33	20	20
CO_2	9	20	28	7
CH_4	0.1	3	12	0
N_2	0.8	1	-	61
Heating Value (Btu/SCF)	290	275	315	100

Methanol and Ethanol

Methanol (methyl alcohol) and Ethanol (ethyl alcohol) are made primarily from petroleum at the present time. They can be made from gasified coal or by fermentation of grains and other agricultural products. The primary impurities are higher alcohols and water. They are normally produced and distributed at relatively high purity. Fuel grade methanol is methanol that will likely be made from coal, and will contain higher than normal levels of ethanol, higher alcohols, and water to reduce the cost of the fuel.

Hydrogen

Commercial grade hydrogen contains a minimum of 99.6% hydrogen. The heating value is 325 Btu/SCF. It is produced from natural gas or light liquid fuels by steam reforming, or from coal and residual oil by partial oxidation. It can also be produced by electrolysis of water and as a byproduct from chlorine production. Processes for producing hydrogen from thermal energy or solar energy are being developed.

LPG (Propane, Butane)

Liquified petroleum gas consists primarily of propane and butane which are gaseous at ambient conditions but are liquids at moderate pressures. The majority of the LPG is a byproduct separated from natural gas before entering the pipeline. A modest amount is a byproduct from refining operations. A small fraction is produced during coal liquefaction. The LPG from petroleum refining or coal liquefaction may contain some unsaturated compounds (propylene, butene).

Heavy Coal Liquids

A variety of middle and heavy distillate hydrocarbon fuels from coal derived synthetic crude oil. The synthetic crude oil may be produced from coal by several alternative processes, examples of which are listed in the table below. Utilization of these fuels in fuel cells would either require reconfiguration of the system's fuel processor, or conversion to a lighter liquid in an intermediate facility.

Coal Liquefaction Process	Coal Liuqid Products may be Replacements for:
Fisher - Tropsch	No. 2 fuel oil No. 6 fuel oil
H-Coal	No. 6 fuel oil
Exxon Donor Solvent	No. 6 fuel oil
Solvent Refined Coal (SRC-II)	No. 2 fuel oil No. 6 fuel oil

EPRI EM-1936

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Final Report
July 1981

Effect of Alternate Fuels on the Performance and Economics of Dispersed Fuel Cells

Contractor: Westinghouse Electric Corporation

An assessment of the effects of alternative fuels—natural gas, coal-derived medium-Btu gas, methanol, ethanol, and naphtha—on the performance and economics of conceptual phosphoric acid fuel cell (PAFC) power plants is presented. Analysis of (1) properties of typical examples of these fuels; (2) PAFC requirements for preliminary specification for fuel processing systems; and (3) vendor-supplied cost, performance, and operational data on fuel processing systems to define ten PAFC power plant configurations utilizing the five alternative fuels is described. 190 pp.

EPRI Project Manager: E. A. Gillis

Cross-References:
1. EPRI EM-1936 2. RP1041-7 3. Fuel Cells and Chemical Energy Conversion Program
4. Fuel Cells

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Contractor: Westinghouse Electric Corporation

An assessment of the effects of alternative fuels—natural gas, coal-derived medium-Btu gas, methanol, ethanol, and naphtha—on the performance and economics of conceptual phosphoric acid fuel cell (PAFC) power plants is presented. Analysis of (1) properties of typical examples of these fuels; (2) PAFC requirements for preliminary specification for fuel processing systems; and (3) vendor-supplied cost, performance, and operational data on fuel processing systems to define ten PAFC power plant configurations utilizing the five alternative fuels is described. 190 pp.

EPRI Project Manager: E. A. Gillis

Cross-References:
1. EPRI EM-1936 2. RP1041-7 3. Fuel Cells and Chemical Energy Conversion Program
4. Fuel Cells

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FUEL CELLS

EPRI

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