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(NASA-CR-169912) HYBRID FUEL CELL/DIESEL GENERATION TOTAL ENERGY SYSTEM, PART 2 Final Report, Apr. - Sep. 1982 (Institute of Gas Technology) 88 p HC A05/MF A01 CSCL 10A
N83-19217
Unclas
G3/44 02829

Part II. HYBRID FUEL CELL/DIESEL GENERATION TOTAL ENERGY SYSTEM

Final Report
(April-September 1982)

by

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Project 65096

for

Jet Propulsion Laboratory

November 1982

This work was performed under JPL Subcontract No. 956210 for the Jet Propulsion Laboratory, California Institute of Technology, Sponsored by the National Aeronautics and Space Administration under Contract NAS7-100

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Summary

The objective of this study was to perform a technical and economic comparison between meeting the Goldstone Deep Space Communications Complex's (GDSCC) electrical and thermal requirements with the existing system and with fuel cells. Fuel cell technology selection was based on a 1985 time frame for installation. As part of this analysis the most cost-effective fuel feedstock for fuel cell application was identified. Fuels considered in this investigation included diesel oil, natural gas, methanol, and coal. These fuel feedstocks were considered not only on the cost and efficiency of the fuel conversion process, but also on complexity and integration of the fuel processor on system operation and thermal energy availability.

After a review of fuel processor technology, catalytic steam reformer technology was selected based on the ease of integration and the economics of hydrogen production. (See Appendix A.) The phosphoric acid fuel cell was selected for application at the GDSCC due to its commercial readiness for near term application. (See Appendix B.) With these technologies selected, fuel cell systems were analyzed for both natural gas and methanol feedstock. Although the methanol fueled system resulted in a lower capital cost, the subsequent economic analysis indicated that a natural gas fueled system was the most cost effective of the cases analyzed. Both the centralized and dispersed system, when sized for the peak load, did not meet the thermal energy requirements at the GDSCC.

Introduction

The market price of an energy source is an important criterion in determining the relative economics of utilization. But with fuel cells, hydrocarbon fuels must be converted to hydrogen. Therefore, the cost of hydrogen production becomes an overriding issue for fuel selection. In Appendix A of this report we determined the more economical fuel feedstock for conversion to hydrogen. Feedstocks that were considered for conversion to hydrogen included natural gas, methanol, diesel fuel, and coal. In a later section of this report the integration of the entire fuel cell system will be analyzed to consider the operational benefits of the different technologies.

Fuel Cell Systems Analysis[8]

The objective of this task is to determine the technical and economic characteristics of a fuel cell system that can meet the existing electrical and thermal requirements of GDSCC. A detailed analysis of the fuel processors analyzed for this study is presented in Appendix A. Based on the relative cost and complexity of integrating the fuel processor section with the fuel cell section, both diesel oil and coal were eliminated from further analysis. A more detailed analysis of catalytic steam reforming to produce hydrogen was performed to determine the system impact when operating on a natural gas or methanol feedstock.

Before the systems analysis task could be carried out, selection of the most appropriate fuel cell type for GDSCC was necessary. Based on the investigation in Appendix B, a phosphoric acid fuel cell was selected for use at GDSCC. This fuel cell was selected over the molten carbonate, alkaline, or solid oxide because of the projected commercial availibility of the phosphoric acid fuel cell in 1985.

The remainder of this section will focus on determining the system configuration and relative economics of a natural gas and methanol powered phosphoric acid fuel cell plant. Both of these systems are similar in configuration except for the fuel conversion section. Since electricity and thermal energy are both desirable products in this study, each fuel conversion section was designed as a nonintegrated component of the entire system. Rather than recovering a substantial portion of the preheating and vaporization energy from the cathode side of the fuel cell, the nonintegrated design uses energy derived from the fuel for these functions. This nonintegrated design has the advantage of lower capital costs and easier conversion to a different fuel if economically advantageous in the future. These advantages are offset by reduced system efficiency.

Another consideration in the design of this system was whether to use a pressurized fuel cell. In general, smaller fuel cells (less than 100 kW) have been designed to operate at near atmospheric pressure, whereas larger systems (greater than 4 MW) have been pressurized. The incentives for operating a fuel cell at elevated pressure include easier integration with the fuel processor, increased reactant partial pressure, and higher cell performance. Increases in cell performance can be considered as increases in current density, which reduce fuel cell size and cost. Problems are also associated with high-pressure operation, however, such as increased component corrosion at the resulting higher voltage.

Pressurized operation also involves the recovery of energy from pressurized gas for reuse in the system. The economy of recovering this energy is a matter of economy of scale. In general, pressurized operation becomes economically favorable at somewhere between 300 and 800 kW, but this is only opinion. A detailed study is required to determine the actual transition point. Since the exact determination of this transition point is beyond the scope of this study, all systems above 500 kW were assumed to benefit from pressurized operation.

The assumptions listed in Table 1 were used as a base in this analysis. A 2.6 $Mw_{(e)}$ (ac) design was chosen to compare the two fuel alternatives. This size corresponds to the peak load at the entire Goldstone Complex.* In both designs a forced-draft, dry, cooling tower was included to handle the fuel cell thermal discharge during low or no thermal demand periods at Goldstone. Incorporation of this cooling tower increases the thermal heat rate of both systems by 200 Btu/kWhr (e).

Table 1. ASSUMPTION USED IN SYSTEMS ANALYSIS

Design Fuel Cell Power Level	2.81 $Mw_{(e)}$ (DC)
Power Conditioner Efficiency	0.96
Ambient Temperature	26.7°C (80°F)
Make Up Water Requirements	none

Methanol Fueled System

A flow diagram of a nonintegrated methanol fueled fuel cell system is presented in Figure 1, and the accompanying mass balance is presented in Table 2. In this design, heat from the combustion of anode off gases at 1093°C (2000°F) is utilized in the fuel processing reformer subsystem. This

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* Includes Pioneer (DSS-11), Echo (DSS-12), Venus, (DSS-13), and Mars (DSS-14) only.

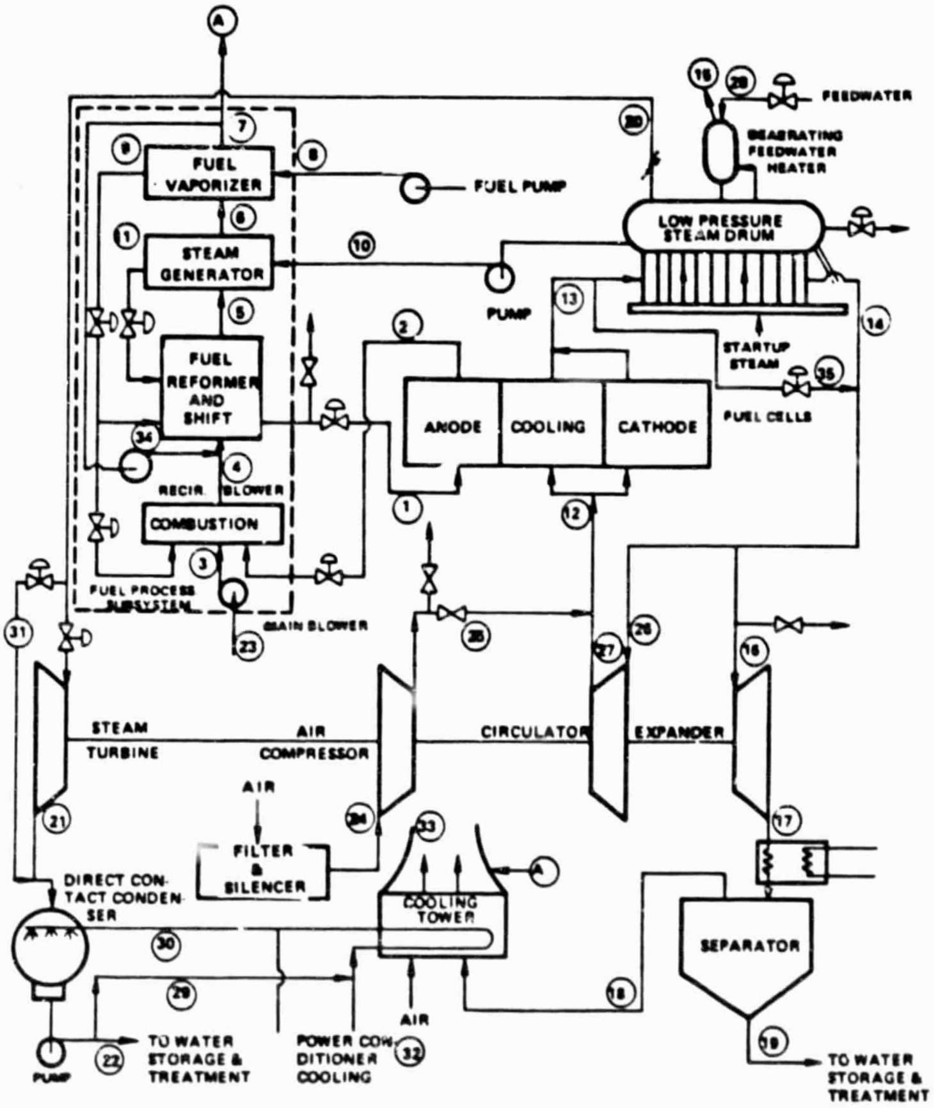


Figure 1. METHANOL FUELED NONINTEGRATED CONCEPTUAL FUEL CELL FLOW SCHEMATIC

energy drives the endothermic reforming reaction to generate steam and to vaporize the incoming methanol feed. In total, the fuel processing subsystem receives fuel, hot water, and an effluent from the anode side of the fuel cell to produce the hydrogen-rich gas required by the fuel cell. The hydrogen-rich gas enters the fuel cell anode where 80% of the hydrogen is consumed. Recirculation of exhaust gases may be necessary to control temperatures in the reformer and to avoid the formation of a vapor plume upon discharge.

The fuel cells are cooled by air and steam (stream 12) to an operating average temperature of 177°C (350°F). The 177°C (350°F) cathode effluent is cooled in the low-pressure steam generator. The steam generated is used in the steam turbine. A portion of the cathode effluent that is not recirculated is expanded to generate additional steam at a supersaturated state. A separator recovers input water, and the resulting supersaturated steam can be used to meet thermal requirements. A direct-contact condenser and a dry cooling tower are incorporated into the design to act as a thermal load when electrical and thermal load requirements differ.

The rotating group consists of a steam turbine, air compressor for cathode supply, recirculator, and expander. The steam turbine is small (approximately 261 kW, 350 hp) and is modeled with three stages. 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 447 kW, 600 hp), while the circulator requires only about 160 kW (215 hp). The expander produces approximately 377 kW (505 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 (about 6%) is bypassed (13) around the steam turbine. At partial power, the compressor, circulator, and expander powers are approximately proportional to flow rate cubed over pressure level. 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 characteristics of the gas) to produce an efficient, compact, heat exchanger. The pinch point temperature

Table 2. STATE POINTS METHANOL FUELED NONINTEGRATED SYSTEM[8]

Station	Temp. °C (°F)	Pressure kPa (psia)	Mole Fraction H_2	H_2O	CO	CO_2	O_2	H_2	CH_3OH	Flowrate kg/hr (lb/hr)
1	177 (350)	345 (50)	0.693	0.075	0.005	0.227	--	--	--	1842 (4060)
2	177 (350)	345 (50)	0.311	0.168	0.011	0.510	--	--	--	1683 (3710)
3	33 (92)	110 (16)	--	0.010	--	--	0.208	0.782	--	2164 (4770)
4	1102 (2015)	103 (15)	--	0.246	--	0.258	0.040	0.456	--	3847 (8480)
5	756 (1383)	103 (15)	--	0.246	--	0.258	0.040	0.456	--	3847 (8480)
6	391 (736)	103 (15)	--	0.246		0.258	0.040	0.456	--	3847 (8480)
7	88 (190)	103 (15)	--	0.246	--	0.258	0.040	0.456	--	3847 (8480)
8	27 (80)	552 (80)	--	--	--	--	--	--	1.000	1066 (2350)
9	149 (300)	517 (75)	--	--	--	--	--	--	1.000	1066 (2350)
10	116 (240)	724 (105)	--	1.00	--	--	--	--	--	780 (1720)
11	171 (340)	689 (100)	--	1.00	--	--	--	--	--	780 (1720)
12	136 (277)	345 (50)	--	0.187	--	--	0.100	0.713	--	187514 (413390)
13	178 (352)	345 (50)	--	0.197	--	--	0.094	0.709	--	187514 (413700)
14	132 (269)	338 (49)	--	0.197	--	--	0.094	0.709	--	187514 (413700)
15	121 (249)	207 (30)	--	1.000	--	--	--	--	--	41 (90)
16	132 (269)	338 (49)	--	0.197	--	--	0.094	0.709	--	11072 (24410)
17	59 (138)	103 (15)	--	0.175g 0.022ℓ	--	--	0.094	0.709	--	11072 (24410)
18	36 (96)	103 (15)	--	0.054g 0.042ℓ	--	--	0.106	0.798	--	10229 (22550)
19	36 (96)	103 (15)	--	1.000	--	--	--	--	--	844 (1860)
20	121 (249)	207 (30)	--	1.000	--	--	--	--	--	4096 (9030)
21	61 (141)	21 (3)	--	1.000	--	--	--	--	--	3842 (8470)
22	57 (135)	21 (3)	--	1.000	--	--	--	--	--	4096 (9030)
23	27 (80)	101 (14.6)	--	0.010	--	--	0.208	0.782	--	2164 (4770)
24	27 (80)	101 (14.6)	--	0.010	--	--	0.208	0.782	--	10932 (24100)
25	175 (348)	345 (50)	--	0.010	--	--	0.208	0.782	--	10932 (24100)
26	132 (269)	338 (49)	--	0.197	--	--	0.094	0.709	--	176582 (389290)
27	134 (273)	345 (50)	--	0.197	--	--	0.094	0.709	--	176582 (389290)
28	57 (135)	207 (30)	--	1.000	--	--	--	--	--	49170 (108400)
29	58 (136)	21 (3)	--	1.000	--	--	--	--	--	131530 (289970)
30	41 (106)	103 (15)	--	1.000	--	--	--	--	--	131530 (289970)
31	118 (244)	90 (13)	--	1.000	--	--	--	--	--	254 (560)
32	27 (80)	103 (15)	--	0.010	--	--	0.208	0.782	--	815754 (1798400)
33	38 (100)	103 (15)	--	0.010	--	--	0.208	0.782	--	815754 (1798400)

g = gas

ℓ = liquid

difference is 9°C (20°F), and the log mean temperature difference is 34°C (72°F) with a pressure-loss ratio ($\Delta P/P$) of 2%.

The conceptual design performance summary is given in Table 3. The power output at the dc bus is 2810 kW. The power conditioner efficiency is assumed to be 96% so that power available at the ac bus is 2697 kW. The parasitic loads are 97 kW so that the net output power is 2600 kW. The overall electrical efficiency is 39%.

Table 3. PERFORMANCE SUMMARY METHANOL FUELED NONINTEGRATED SYSTEM

Gross Electrical Output, kW_e (dc)	2810	
Gross Electrical Output, kW_e (ac)	2697	
Parasitic Losses, ($kW_{(e)}$)		
Pumps	7	
Fans (Cooling Tower)*	78	
Vacuum Pump	1	
Air Compressor & Dryer	1	
Controls	10	
Net Electrical Output, kW_e (ac)	2600	
Input Energy (hhv = 6.3 kW-hr/kg) (hhv = 9758 Btu/lb), kW-hr/hr (Btu/hr)	6693.41	(22.86×10^6)
Overall Plant Efficiency	0.39	
Heat Rate, kW-hr/kW-hr (Btu/kW-hr)**	2.57 (8791)	
Design Current Density, A/m^2 (A/ft^2)	3229 (300)	
Fuel Utilzation Factor	0.8	
Design Cell Voltage, V/cell)	0.66	

* Required when thermal load is not present.

** Includes 0.059 kWhr/kWhr (200 Btu/kWhr) heat rate loss to cooling tower.

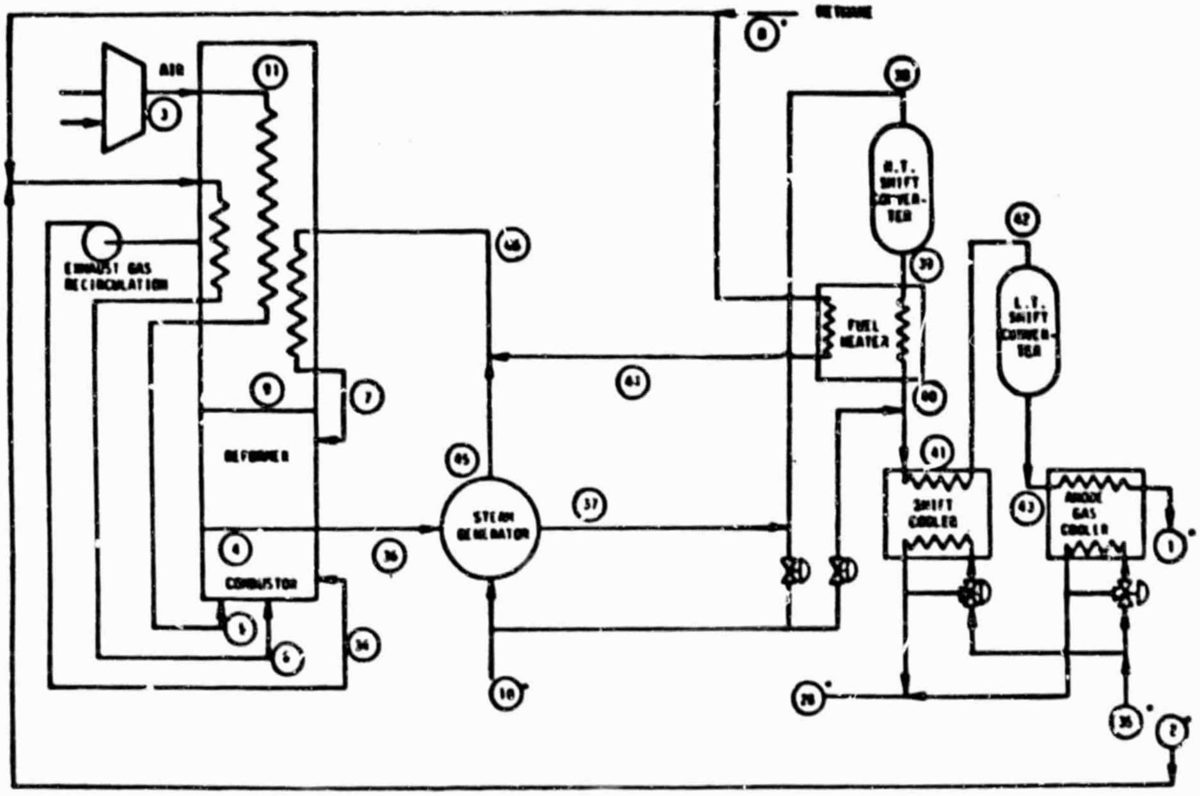
The power plant is started with 205 kPa (15 psig) steam from an auxiliary boiler. This steam is directed to flow up through the tubes of the steam generator. This heats the water to saturation and pressurizes the steam lines. The condensing of this steam requires that some water be removed to maintain the proper water level in the steam drum. Once the steam generator is pressurized, the air compressor is valved off at station (25) and vented to the atmosphere, and the valve at station (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 station (35) to ramp up the temperature in a controlled manner. After the fuel cells are heated to about 121°C (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 start-up steam is shut off. This puts the plant into the normal control mode.

Thermal energy can be taken from two places. A stream of hot water can be used from flows (31) and (21) before they enter the cooling tower system, or energy can be extracted from the warm flue gases in stream (7). However, this thermal energy is mostly available from 60° to 88°C (140° to 190°F). This is adequate for meeting hot water needs, but will not suffice where steam above atmospheric pressure is required. Higher temperature thermal energy is available at 132°C (269°F) and 338 kPa (49 psia) in stream (14). Tapping this stream for steam will divert energy from the turbo machinery that is required for compression. To increase the availability of thermal energy, the tubo machinery section must be replaced by electric motor drivers, thus lowering the overall electrical generation efficiency. A more detailed optimization study is required to determine the best approach.

Natural Gas Fueled System

A natural gas powered fuel cell is similar in design to the previously described methanol system. The major change is in the fuel processing subsystem where the dashed line segment in Figure 1 is replaced by the natural gas processing subsystem in Figure 2. The state points of this new system configuration that differ from the nonintegrated methanol design are shown in Table 4.



*Interface Points

Figure 2. NATURAL GAS (METHANE) FUELED NONINTEGRATED FUEL PROCESSING SUBSYSTEM

Table 4. STATE POINTS NATURAL GAS (METHANE) FUEL PROCESSING SUBSYSTEM

Station	Temp. °C (°F)	Mole Fraction H_2	Mole Fraction H_2O	Mole Fraction CO	Mole Fraction CO_2	Mole Fraction O_2	Mole Fraction N_2	Mole Fraction CH_4	Flowrate kg/hr (lb/hr)
1*	177 (350)	0.726	0.079	0.010	0.174	--	0.001	0.010	1501 (3308)
2*	177 (350)	0.346	0.139	0.023	0.415	--	0.004	0.023	1342 (2959)
3	27 (80)	--	0.016	--	--	0.204	0.780	--	2486 (5480)
4	1977 (3590)	--	0.0284	--	0.208	0.006	0.502	--	3858 (8505)
5	704 (1300)	--	0.016	--	--	0.204	0.780	--	2486 (5480)
6	704 (1300)	0.335	0.183	0.022	0.402	--	0.004	0.054	1373 (3026)
7	729 (1344)	--	0.609	--	--	--	0.003	0.388	1173 (2586)
8*	27 (80)	--	--	--	--	--	0.008	0.992	458 (1010)
9	1018 (1864)	--	0.284	--	0.208	0.006	0.502	--	3858 (8505)
10*	116 (240)	--	1.000	--	--	--	--	--	1073 (2365)
11	110 (230)	--	0.284	--	0.208	0.006	0.502	--	3858 (8505)
18	36 (96)	--	0.032g 0.030ℓ	--	--	0.110	0.828	--	9985 (22013)
19	36 (96)	--	1.000	--	--	--	--	--	1089 (2401)
28*	79 (175)	--	1.000	--	--	--	--	--	5170 (11398)
35*	827 (135)	--	1.000	--	--	--	--	--	5170 (11398)
36	827 (1520)	0.665	0.110	0.185	0.028	--	0.002	0.011	1173 (2586)
37	379 (714)	0.665	0.110	0.185	0.028	--	0.002	0.011	1173 (2586)
38	247 (476)	0.609	0.185	0.169	0.028	--	0.002	0.010	1369 (3019)
39	367 (693)	0.704	0.090	0.074	0.120	--	0.002	0.010	1369 (3019)
40	284 (544)	0.704	0.090	0.074	0.120	--	0.002	0.010	1369 (3019)
41	208 (407)	0.665	0.139	0.070	0.114	--	0.002	0.010	1500 (3307)
42	166 (330)	0.665	0.139	0.070	0.114	--	0.002	0.010	1500 (3307)
43	239 (462)	0.726	0.079	0.010	0.174	--	0.001	0.010	1500 (3307)
44	322 (611)	--	--	--	--	--	0.008	0.992	424 (934)
45	164 (328)	--	1.000	--	--	--	--	--	745 (1643)
46	227 (440)	--	0.609	--	--	--	0.003	0.388	1173 (2586)

* See Figures 7 and 8.

g = gas

ℓ = liquid

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This nonintegrated design uses about 7% of the incoming natural gas feed to supply the endothermic heat of reaction for the reforming step. The remaining 93% of the natural gas feedstock is heated to 322°C (611°F) and mixed with steam in a mole ratio of 1.57/1 steam/methane. This mixture is then heated to 729°C (1344°F) by combustion gases before it enters the reformer. Approximately 95% of the natural gas is reformed at 1520°F before it exits the reformer for cooling to 379°C (714°F). At this point additional steam is added to the synthesis gas mixture to adjust the mole ratios for the high-temperature shift reaction at 367°C (693°F). Two shift reactions convert the carbon monoxide to hydrogen. Before entering the second-shift reactor the gas is cooled to 166°C (330°F). After the exothermic reaction in the second shift converter the gas is again cooled to 177°C (350°F) before entering the anode of the fuel cell.

After shift conversion and cooling, the synthesis gas is composed of primarily hydrogen and carbon dioxide. Approximately 80% of the hydrogen entering the fuel cell anode is utilized. The anode exhaust gases are then used in a combustion reaction where a small amount of natural gas and 10% excess preheated air are used to produce a hot gas with a temperature of about 1977°C (3590°F). This gas is then routed to the reformer to supply the heat of reaction and to preheat the reformer feed and combustion air.

The natural gas and methanol systems are very similar in design. The water systems differ because natural gas reforming requires more steam at greater temperatures. This lowers system efficiency. In addition, the natural gas reformer subsystem is much more complex. Table 5 is a performance summary of the conceptual design of the nonintegrated natural gas system.

Thermal energy can be taken from the natural gas fueled system in two places. As shown in Table 5, thermal energy can be obtained from the 110°C (230°F) exhaust combustion products leaving the reformer or from the hot water leaving the low-pressure steam drum (stream 31). As in the methanol system, this thermal energy is of low quality and will probably only be used to meet hot water needs. Higher quality thermal energy can be obtained by replacing the turbo machinery with electric drivers, thus lowering the electrical generation efficiency, or by increasing the amount of natural gas that is combusted for use in the reformer. This can increase the availability of high-quality heat to the system. A more detailed analysis is required to optimize the production of both electricity and thermal energy depending on thermal energy quality requirements at Goldstone.

Table 5. PERFORMANCE SUMMARY NATURAL GAS (METHANE) NONINTEGRATED SYSTEM

Gross Electrical dc Output, kW	2810
Gross Electrical ac Output, kW	2697
Parasitic Losses, $kW_{(e)}$	
Pumps	8
Fans (Cooling Tower)*	78
Vacuum Pump	1
Air Compressor & Dryer	1
Controls	10
Net Electrical ac Output, kW	2600
Input Energy, kWr/hr (Btu/hr)	7056 (24.10×10^6)
Overall Plant Efficiency	0.368
Heat Rate, kWhr/kWhr (Btu/kW-hr)	2.71 (9270)

* Required when thermal load is not present.

** Includes 0.059 kWhr/kWhr (200 Btu/kWhr) heat rate loss to cooling towers.

Fuel Cell System Cost Comparisons[8]

As mentioned earlier, the methanol and natural gas powered fuel cell systems are technically very similar. For this reason a detailed economic evaluation is necessary to clearly determine the "best" fuel feedstock for fuel cell operation at Goldstone. This comparison will focus on a fuel cell size that is most appropriate to this analysis. Two sizes of the fuel cell system were determined by combining the commercial power load profile for the entire Goldstone complex and sizing the fuel cell to meet the peak electrical demand. The peak demand was calculated by adding the system segment load profile curves during the months of February and August. Figures 3 and 4 present the power load profiles for February and August 1978 for the Mars and Goldstone system. The Goldstone electrical load profile curves include the Echo, Venus, and Pioneer segments. The peak demand for the combined Goldstone and Mars segments during February 1978 occurred during the 13th day when demand reached nearly 2.6 MW as shown in Figure 3. Similarly, the peak load

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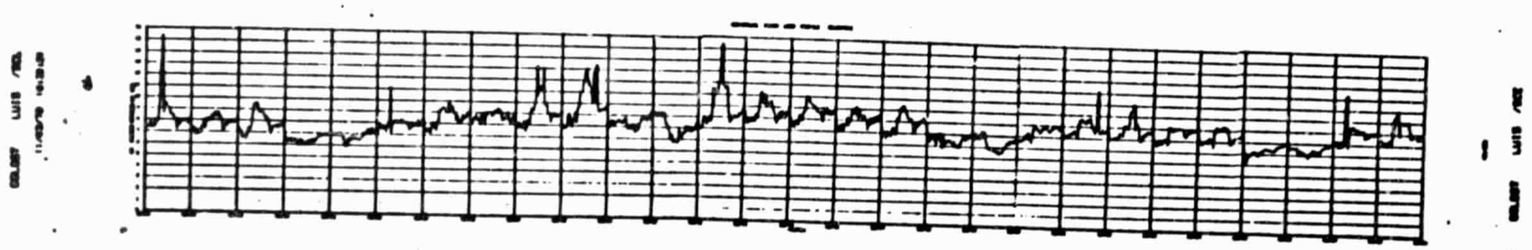
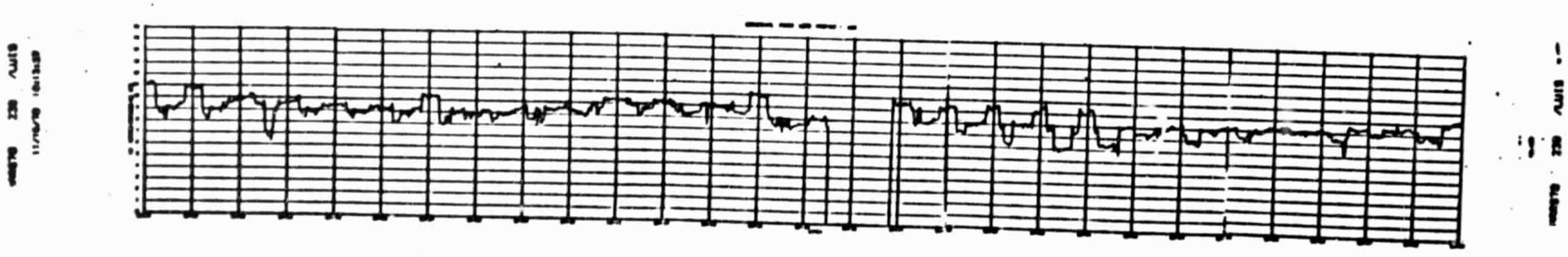
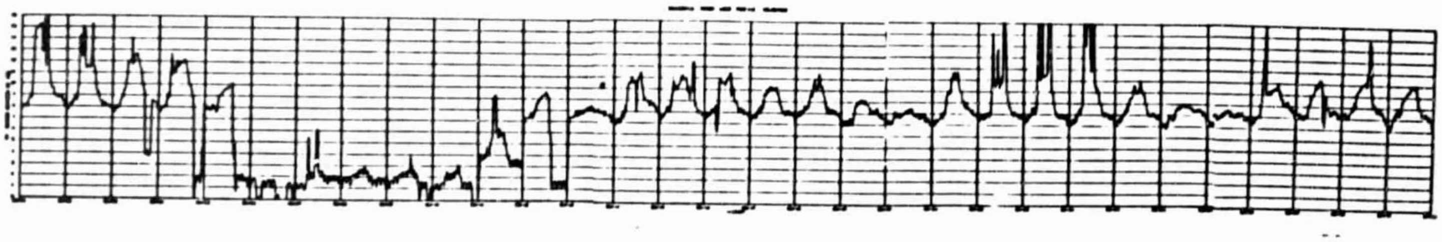


COMMERCIAL POWER LOAD PROFILE: GOLDSTONE — FEBRUARY 1978

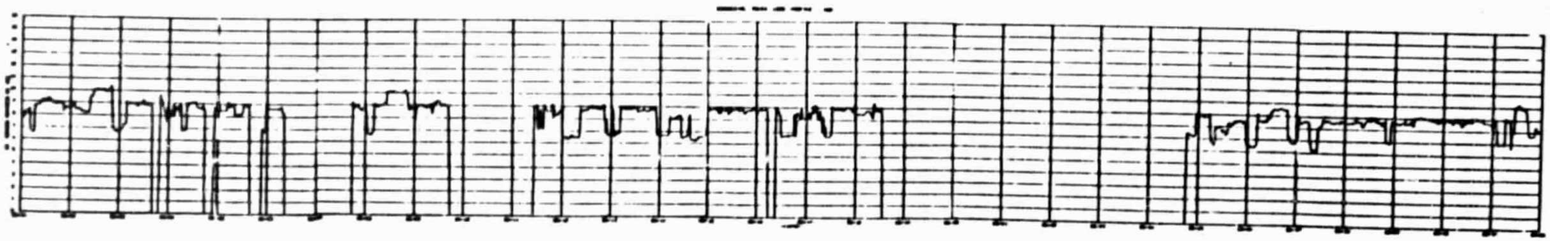


COMMERCIAL POWER LOAD PROFILE: MARS — FEBRUARY 1978

Figure 3. COMMERCIAL POWER LOAD PROFILE FOR GOLDSTONE COMPLEX DURING FEBRUARY 1978



COMMERCIAL POWER LOAD PROFILE: GOLDSTONE — AUGUST 1978



COMMERCIAL POWER LOAD PROFILE: MARS — AUGUST 1978

Figure 4. COMMERCIAL POWER LOAD PROFILE FOR GOLDSTONE COMPLEX DURING AUGUST 1978

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in August 1978 occurred on the 2nd day when the demand peaked at just over 2.5 MW as shown in Figure 4. Therefore, a system size of 2.6 MW (ac) was chosen for the comparison. The last fuel cell system analyzed differs from the other two systems in the sizing philosophy. This system was sized to meet the minimum monthly average electrical load. As shown in Table 9 this corresponds to December 1981 when the load was approximately 1.1 million kW-hrs. Averaged out over the month and using a 90% fuel cell utilization rate, an appropriate fuel cell size of 1.7 $MW_{(e)}$ was calculated for this case.

In the technical section the fuel cell systems were designed as nonintegrated, self-sufficient facilities. Cost estimates for this size system are based on prototype construction because these units are not generally commercially available. A summary of these prototype system costs is presented in Tables 6 and 7 for the methanol and natural gas powered concepts, respectively. These cost estimates have been updated to 1982 dollars. Based on the Energy Research Corporation estimates of fuel cell stack assembly costs, \$375/kW_e was used for the stack cost and \$641/kW_e was used for the installed fuel cell assembly. Costs for the fuel processing section and all other major equipment were based on vendor estimates where possible. All costs include a 25% increase for IR&D, G&A, and fee.

The capital equipment in these tables includes 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 including storage of liquid fuels (for methanol), pumps, piping, fittings, insulation, vaporizer, (for liquid fuels), heat exchangers, reformer, shift converters, combustor, blowers, instrumentation, foundations, supports, etc. The rotating equipment and auxiliaries systems account for the steam turbine and expander, condenser separators, induced-draft cooling towers, air compressor system, filters, gas circulator, piping, fittings, pumps, instrumentation, foundations, supports, etc. The electric generating system has a low-pressure boiler system, recirculation ducts, fuel cell assemblies, piping and fittings, supports, instrumentation and controls, foundations, etc. Accessory electric equipment consists of a power conversion system, data acquisition system, instrumentation and controls, diesel generator system (if required), etc. Other miscellaneous power plant equipment includes the compressed air system for

Table 6. METHANOL PROTOTYPE 2.6 MW_e AC PHOSPHORIC ACID FUEL CELL (PAFC) PLANT NONINTEGRATED
DIRECT CAPITAL COST — 1982 BASIS
(THOUSANDS OF DOLLARS)

Account No.(2)	10^3
341 — Structure and Improvements	77
342 — Fuel Handling and Processessing(4)	1045
• Fuel Handling	30
• Fuel Processing	1015
343 — Rotating Equipment and Auxiliaries	385
• Steam Turbine/Expander	84
• Condenser System	15
• Separator System	10
• Cooling Tower	63
• Air Compressor System	140
• Air Filter/Silencer System	1
• Cathode Exhaust Gas Circulator	46
• Miscellaneous Auxiliaries	26
344 — Electrical Generating System	1730
• Low Pressure Boiler System	46
• Recirculation Ducts	18
• Fuel Cell System	1666
345 — Accessory Electric Equipment	1165
• Power Conversion System	813
• Instrumentation and Control	185
• DAS System	116
• Diesel Generator System	51
346 — Other Miscellaneous Power Plant Equipment	126
353 — Station Equipment	260
• Main Transformer	260
Total Direct Capital Cost (1),(3) (Land Not Included)	4788 ($1842/kW)

[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).

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Table 7. GAS PROTOTYPE 2.6 MW_e AC PHOSPHORIC ACID FUEL CELL (PAFC) PLANT NONINTEGRATED DIRECT CAPITAL COST — 1982 BASIS (THOUSANDS OF DOLLARS)

Account No.(2)	10^3
341 — Structure and Improvements	77
342 — Fuel Handling and Processessing(4)	1383
● Fuel Handling	30
● Fuel Processing	1353
343 — Rotating Equipment and Auxiliaries	383
● Steam Turbine/Expander	84
● Condenser System	15
● Separator System	10
● Cooling Tower	63
● Air Compressor System	140
● Air Filter/Silencer System	1
● Cathode Exhaust Gas Circulator	46
● Miscellaneous Auxiliaries	26
344 — Electrical Generating System	1730
● Low Pressure Boiler System	46
● Recirculation Ducts	18
● Fuel Cell System	1666
345 — Accessory Electric Equipment	1165
● Power Conversion System	813
● Instrumentation and Control	185
● DAS System	116
● Diesel Generator System	51
346 — Other Miscellaneous Power Plant Equipment	126
353 — Station Equipment	260
● Main Transformer	260
Total Direct Capital Cost (1),(3) (Land Not Included)	5126 ($1971/kW)

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[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).

pneumatic valves, water treatment and storage, inerting system, hydrogen system, sampling system, cranes, highlights, etc. Station equipment includes the main transformer and any other site-related equipment interfacing with the utility grid.

These plant costs are based on the cost of producing a prototype because this type of facility is not commercially available. Actual capital costs could drop by as much as 69% when and if full-scale commercial production volumes are reached. This will probably not occur before 1985 or even 1990. In addition to the direct costs, a 10% contingency factor is included in the economic analysis to cover any indirect costs. Fixed operating and maintenance costs were assumed to be \$4.5/kW-yr, and variable operating and maintenance costs were estimated to be 4.4 mills/kWhr. These O&M costs are presented at 1982 dollars and do not include fuel costs.

Natural gas price estimates were obtained from Southwest Gas Corp., which supplies the Barstow area. Current commercial rates are \$0.024 kW-hr (\$0.70 per therm), which is about \$7.00 per million Btu. Methanol fuel, which is chiefly derived from natural gas, is estimated to cost 26.2¢/liter (\$0.99 per gallon), or \$0.05 per kW-hr (\$14.50 per million Btu) including taxes.

In addition to the above information, the criteria in Tables 8 through 10 were also used in the economic evaluation. Table 8 presents the technical constraints of the fuel cell system. The most important criterion in this table for the analysis is the projected 20 year project life, which is based on the replacement of the fuel cell stack every 40,000 hours of operation. Another recurring cost that is not shown in Table 8 but is included in the economic analysis is a \$250,000 charge every 3 years for catalyst and chemicals. Table 9 presents the energy consumption profile that was used in the analysis. Table 10, which is based on fiscal 1981 data, presents LPG, diesel fuel, and purchased electrical energy as well as a breakdown of end-use applications in the complex.

From these data we determined that a 2.6 MW fuel cell system (which was sized for the peak demand) could supply the Goldstone complex with electricity at a 68% yearly load factor. With this use pattern nearly 56% of the complex's thermal energy needs will be met by the fuel cell. This is based on utilizing all thermal energy above 100°F. This does not include distribution thermal losses, nor was any attempt made to see if the fuel cell thermal output matched the thermal quality requirements of the complex.

Table 8. CONSTRAINTS FOR FUEL CELL POWER PLANTS

Start up from cold (from standby to hold in 15 seconds)		~ 4 hours
Start/stop once per week		
Unattended operation		
Modular construction		
Minimumn to maximum	~ 1 second preferred ~ 1 minute acceptable	
Standby to maximum power	~ 15 second preferred ~ 2 minute acceptable	
Minimum Power		— 0.65 MW
Projected Life*		20 years
Fuel Consumpiton @ 0 Net power, kW-hr/hr (10^6 Btu/hr) (hold)		146.4 (0.5)
Reactive Power (±90° up to full MVAR rating)		0.2 seconds

* Cell stack replaced every 40,000 hrs.

For the life-cycle costing analysis the procedures of the National Bureau of Standards Handbook 135 were followed. The analysis determines the savings-to-investment ratio (SIR) for a retrofit-type project. Future expenses are adjusted to present values. First, the total life-cycle cost without retrofit is determined. A 20-year period is assumed (expected life of the conversion or retrofit). Savings with the retrofit are then calculated, and the SIR is determined. The SIR's are in Table 11; life-cycle cost calculations are in Appendix C.

The cost calculations in Appendix C include 12 different retrofit analyses. We reviewed three different system applications that use natural gas or methanol feedstock and assume a commercially available or prototype system cost. The first application assumes one centralized fuel cell to serve the entire complex and is sized for a 2.6 $MW_{(e)}$ peak load. The second system analyzed a two fuel cell dispersed arrangement, which will reduce thermal energy distribution costs. This dispersed fuel cell system consists of a 1.0 $MW_{(e)}$ fuel cell sized for the peak at the Mars busbar and a 1.6 $MW_{(e)}$ fuel cell sized for the peak at the Goldstone* busbar. The third system differs from the other two systems in the sizing philosophy. This system consists of the centralized fuel cell sized at 1.7 $MW_{(e)}$. This is of sufficient size to meet the minimum monthly electrical load. In all applications the cost of

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* Goldstone includes Echo, Pioneer, and Venus only.

Table 9. GOLDSTONE YEARLY ENERGY CONSUMPTION AND CATEGORIZED END USE CONSUMPTION[1]

	LPG(2)		Diesel(3) Fuel		Purchased(4) Electricity		Total Energy	Total Electric Energy
	(gal)	(MWHt)	(gal)	(MWHt)	(kwh)	(MWHt)	(MWHt)	(kWh)
FY/81								
October	5,101	143	38,030	1,546	963,706	3,276	4,965	1,420,066
November	4,900	137	65,250	2,652	645,566	2,194	4,983	1,428,566
December	6,974	195	21,250	864	853,074	2,900	3,959	1,108,074
January	6,069	170	25,230	1,025	1,050,006	3,569	4,764	1,352,766
February	6,275	176	21,750	884	972,026	3.304	4,364	1,233,026
March	6,221	174	24,160	982	952,034	3,236	4,392	1,241,954
April	3,572	100	22,890	930	878,960	2,988	4,108	1,153,640
May	2,950	83	25,800	1,049	899,147	3,056	4,188	1,208,747
June	1,044	29	27,270	1,108	924,556	3,143	4,280	1,251,796
July	1,025	29	26,350	1,071	1,074,557	3,652	4,752	1,390,757
August	1,568	44	53,830	2,188	754,119	2,563	4,795	1,400,079
September	1,000	28	46,620	1,895	716,509	2,435	4,358	1,275,949
Total	46,699	1,307	398,430	16,192	10,684,260	36,316	53,815	15,465,420

(1) For Pioneer (DSS-11), Echo (DSS-12), Venus (DSS-13), Mars (DSS-14) only.

(2) For Heating only; 95,500 Btu/gal.

(3) For on-site power generation without waste heat utilization; 138,700 Btu/gal; 12 KWH_e/gal approx.

(4) $KWH_e = 3.399 \times 10^{-3}\ MWH_{(t)}$

Table 10. ENERGY END-USE AT THE GDSCC

	KWH_e	$MHW_{(t)}$	%
Heating (LPG)	--	1,307	2.4
Heating (Elec.)	617,933	2,100	3.9
Cooling (V.C. Chiller)	4,634,502	15,753	29.3
Lighting	1,081,384	3,675	6.8
Electronic Equipment	4,634,502	15,753	29.3
Mech./Elec. Accessories	4,480,017	15,227	28.3

thermal energy distribution and storage was ignored. A scaling factor of 0.3 was used to calculate the smaller fuel cell capital costs.

For each of the three system configurations a different cost calculation was performed for methanol and natural gas. Because fuel systems are not yet a commercial reality, all the cost calculations were repeated to show the estimated capital costs for a prototype system and the expected cost of a commercially manufactured fuel cell system. This resulted in the 12 different calculations in Table 11. A savings-to-investment ratio greater than one indicates that the investment is cost effective; the higher the ratio, the greater the dollar savings per dollar spent.

As shown in Table 11 the only economically attractive case was the centralized commercially available system sized for minimum monthly load, which was fueled with natural gas. This system does not represent the optimum system because this was beyond the scope of this analysis.

Table 11. SAVINGS-TO-INVESTMENT RATIOS FOR COMMERCIAL AND PROTOTYPE NATURAL GAS AND METHANOL POWERED FUEL CELLS

	Centralized Peak Load Sized System	Dispersed Peak Load Sized System	Centralized Average Sized System
Natural Gas Powered Commercial Unit	0.941	0.914	1.309
Natural Gas Powered Prototype Unit	0.360	0.347	0.496
Methanol Powered Commercial Unit	−3.233	−3.148	−3.914
Methanol Powered Prototype Unit	−1.248	−1.207	−1.496

Conclusions and Recommendations

The major conclusions from this study are as follows:

- For the fuels analyzed in this study, natural gas is the preferred fuel when system sizing is considered.
- While there is extensive commercial experience with fuel processors that convert naphtha and natural gas into hydrogen, the complexity of a fuel processor subsystem 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. Production and manufacturing planning must be an inherent and early part of the fuel processing system and complete phosphoric acid fuel/cell power plant design process in order to achieve such benefits in actual practice.

- The fuel cost alone is by far the largest component in all the units considered.

- The phosphoric acid fuel cell (PAFC) may not have enough quality excess heat for thermal loads. A molten carbonate fuel cell is more appropriate for thermal energy generation.

The major recommendations resulting from this study are as follows:

- Determine the most cost-effective fuel cell size that would minimize the total life-cycle cost of a combined system of fuel cell, utility-purchased, and diesel-generated power.

- 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.

- The economical transition point between atmospheric and pressurized operation of the PAFC should be determined.

- A detailed study should be conducted to determine the required location, consumption, and quality of heat at Goldstone.

- The cost of distributing thermal energy throughout the Goldstone complex should be determined.

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

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

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APPENDIX A.

Fuel Conversion — Hydrogen Production

For the four energy feedstocks considered; natural gas, methanol, diesel fuel, and coal, three different technologies are required for the conversion to hydrogen. These technologies are: 1) catalytic steam reforming, partial oxidation, and coal gasification. For the natural gas and methanol feedstocks, catalytic steam reforming technology must be used to produce hydrogen. The catalytic steam reforming process, commercially available since the 1930's, is the most widely used and the most economical process for producing hydrogen from light hydrocarbon gases or from naphthas. In this process light hydrocarbons, (including natural gas, methanol, ethanol, to light naphtha, and heavy naphtha,*) are converted to a synthesis gas containing hydrogen, carbon monoxide, and carbon dioxide by reaction with steam over a catalyst.

With methane as a feedstock, the reaction is —

$$CH_4 + H_2O \rightarrow CO + 3H_2 \qquad (1)$$

This highly endothermic reaction is carried out at 650°C to 1000°C (1200°F to 1830°F) at (100 to 700 psig) in a reformer tube furnace fueled by the feedstock as shown in Figure 1.

Because the nickel-based catalysts used in this process are sensitive to sulfur, the hydrocarbon feed stream must be desulfurized before it enters the steam reformer. In the first step of the desulfurization process, the feed stream is passed over a colbalt/molybdenum catalyst in the presence of 5% hydrogen; the reaction of sulfur-containing compounds and hydrogen produces hydrogen sulfide (H_2S). The H_2S containing feedstock is then cooled and scrubbed with monoethanolamine (MEA) to reduce the H_2S concentration to 25 ppm. Final polishing of the feedstock is performed in a zinc oxide (ZnO) bed at 340° to 370°C (644°F to 698°F), where the H_2S is further reduced.

The cleaned gas is then sent to the primary reformer, a direct-fired chamber containing high nickel alloy (20% to 35% Ni) tubes that are normally 10.8 cm OD by 7.3 cm ID (4-1/4 inches OD by 2-7/8 inches ID). These tubes

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* Advances in materials, design, and operation will allow the use of distillate fuels as a feedstock after the 1985 timeframe.

contain the nickel-based catalyst (12% to 25% Ni) which is supported as NiO on alumina. The feedstock is mixed with steam in a steam/carbon ratio of 2.5 to 5.0 and is passed over the catalyst with a space velocity of 5000 to 8000 vol/hr^{-1}. The reacted mixture exits the reformer at temperatures of around 105°C to 870°C (1300°F to 1600°F) and a pressure usually in the range of 325 to 500 psig. The flue gas temperature exiting the furnace section is 980°C to 1040°C (1800°F to 1900°F.) These hot flue gases are used to generate superheated steam for the hydrocarbon-steam reaction, feedstock preheating, and to drive steam turbomachinery. When methane is the feedstock, the conversion rate in the reformer approaches 95%.

The synthesis gas exiting the reformer is comprised of approximately 76% H_2, 12% CO, 10% CO_2, and 1.3% CH_4. This gas is further processed in two downstream reactors, where the water-gas shift reaction takes place:

$$CO + H_2O \rightarrow H_2 + CO_2 \qquad (2)$$

The first shift reaction takes place at 343°C to 455°C (650°F to 850°F) over a chromium-promoted iron oxide catalyst. Before entering the second shift reaction (which is very sensitive to sulfur poisoning) the gas may again be desulfurized in a Zinc Oxide (ZnO) bed. The second shift reaction is then carried out over a copper-zinc catalyst at a temperature of 200°C to 230°C (390°F to 450°F) to produce a gas with 85% H_2, 22% CO_2, 0.25% CO, and 1.3% CH_4. This gas can then scrubbed to remove the CO_2 by a process such as the MEA process or by pressure swing absorption. At this point the gas contains 98.2% H_2, 0.3% CO, 0.01% CO_2, and 2.5% CH_4. Depending on purity requirements, the remaining carbon oxides can be further reduced by passing the gas over a methanation catalyst (nickel oxide) to promote methane formation from the remaining carbon oxides and hydrogen.

A flow diagram of the catalytic steam reforming process is presented in Figure 5. The utility and feedstock requirement for a large natural gas reforming facility producing hydrogen with a purity in excess of 97% is shown in Table 12.

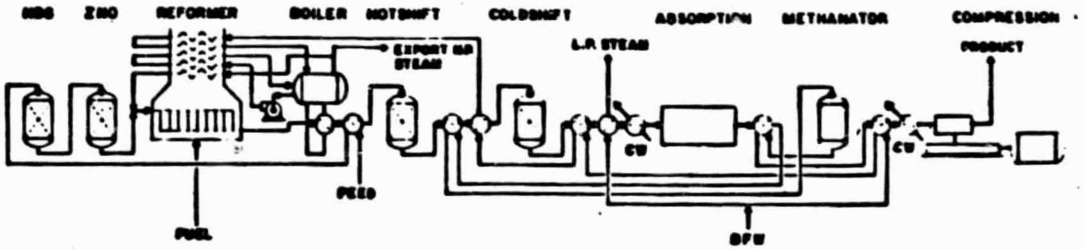

Figure 5. LOW PRESSURE STEAM-HYDROCARBON REFORMING HYDROGEN PROCESS

Table 12. EFFICIENCY, PROCESS, AND UTILITY REQUIREMENTS OF STEAM REFORMING NATURAL GAS FOR THE PRODUCTION OF 97%-PURE HYDROGEN
Per 28.32 m^3 (Per 1000 SCF) of Hydrogen[1]

Process Feed, kg of methane	(11.9 lbs.)
Fuel, kg of methane	(7.9 lbs.)
Electric Power, kWhr	0.4
Cooling Water, m^3	(400 gal)
Boiler Feedwater, m^3	(10 gal)
Condensate Returned, m^3	(6 gal)

Input 19.8 lb of methane X 23,800 Btu/lb = 138.02 kWhr 472,000
0.4 kWhr electricity X 3413 Btu/kWhr = 0.4 kWhr 1,400

Total 8.98 kg of methane X 15.37 kWhr/kg
0.4 kWhr of electricity 138.42 kWhr (474,200)

28.32 m^3 of Hydrogen X 3.36 $kWhr/M^3$
Output = (1000 SCF of Hydrogen X 325 Btu/SCF) = 95.22 kWhr (325,000)

$$\text{Efficiency} = \frac{\text{Output}}{\text{Input}} = \frac{95.22}{138.42} \times 100 \quad \frac{325{,}000}{474{,}200} \times 100$$

$$= 68.5\%$$

Partial Oxidation Processes[2,3]

As the molecular weight of the carbonaceous fossil fuel continues to increase beyond the naphtha range, the nickel based catalysts lose their effectiveness in promoting the reaction between the hydrocarbon and steam. To convert the heavier hydrocarbons to hydrogen it is necessary to subject them to partial oxidation using gasifier technology. Two partial-oxidation processes have been extensively commercialized: the Texaco Process (since 1954) and the Shell Gasification Process (since 1956). These two noncatalytic pressurized partial-oxidation processes produce a gas consisting primarily of hydrogen and carbon monoxide by burning hydrocarbons with high-purity oxygen or an oxygen-rich stream.

Unlike the catalytic steam reforming process, the partial-oxidation process can operate on nearly any type of pumpable or compressible hydrocarbon feedstock, from methane gas through crude and residual oils to asphalts, regardless of their sulfur content. This process's oxygen feedstock requirements make it both more capital and operating cost intensive than the catalytic steam-reforming process. For this reason the partial-oxidation process has been used in areas where light hydrocarbon feeds are either unavailable or less economical than heavy hydrocarbons such as crude oil.

In the partial-oxidation process as shown in Figure 6 a preheated hydrocarbon feed, preheated oxygen, and steam are injected into a pressurized combustion vessel through specially designed burners. By controlling the amount of oxygen entering the vessel, complete combustion does not occur; instead, the following overall partial oxidation reaction occurs at 1290°C to 1400°C (2350°F to 2550°F):

$$C_n H_M + (N/2)\ O_2 \rightarrow (N)CO + (M/2)\ H_2 \qquad (3)$$

Reaction 3 occurs in three distinct phases within given regions in the gasification vessel. The first-phase reactions occur when the preheated hydrocarbons come into contact with the oxygen and steam mixture at the burner tip. At this point the mixture is rapidly heated and vaporized by radiation from the flame front and reactor walls. This rapid heating cracks the heavy hydrocarbons to carbon, methane, and hydrocarbon radicals. These first-phase reaction products enter the second phase, where partial oxidation occurs in the highly exothermic combustion reaction —

$$C_NH_M + (N+M/4)O_2 \quad (N)CO_2 + (M/2)\ H_2O \qquad (4)$$

In the combustion reaction nearly all of the available oxygen is consumed. The heat from the combustion reaction drives the endothermic reaction of the remaining hydrocarbons with steam and the combustion products. The principal reaction in this phase is the reforming of the hydrocarbons with steam in the following endothermic reaction:

$$C_NH_M + (N)H_2O \rightleftarrows (N)CO + (N + M/2)\ H_2 \qquad (5)$$

The thermal equilibrium of Reaction 5 occurs around 1340°C (2450°F).

In the final phase occurring within the gasification vessel, part of the high-temperature gases and unreacted carbon react with the carbon dioxide (CO_2) and steam. Nearly 97% to 99% of the carbon entering as feed is reacted before exiting the reactor, compared with a 99.98% carbon conversion efficiency for the catalytic steam-reforming process. The hot gases that exit the partial oxidation reactor undergo a water-gas shift reaction ($CO + H_2O + CO_2 + H_2$), which increases the concentration of hydrogen gas.

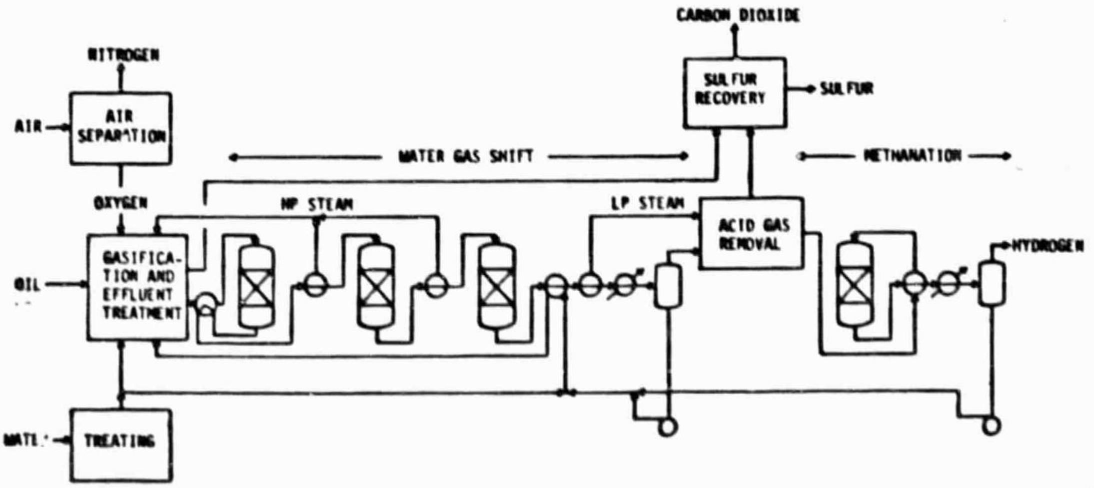

Figure 6. OIL PARTIAL OXIDATION HYDROGEN PROCESS

The reducing atmosphere within the reactor promotes the formation of sulfur compounds. The sulfur that enters with the feed is converted to hydrogen sulfide (H_2S) and small amounts of carbonyl sulfide (COS) that are removed in downstream processing equipment.

The Shell and Texaco partial oxidation processes are very similar. In the Shell process five basic units are used to gasify an oil feedstock. These five units are the gasification reactor, the waste-heat exchanger, the economizer heat exchanger, the carbon removal system, and the carbon recovery system.

The vertical steel pressure gasification vessel is multiple-layer refractory lined. The preheated feedstock and oxidant are fed to the combustor, which is located at the top of the vessel. The steam is mixed with the oxygen in sufficient quantity to moderate the flame front. The hydrocarbon feed and oxidant are sprayed as a rotating vortex into the combustion zone to promote mixing. The reactor can be designed for pressures from as low as 207 kPa (15 psig) to as high as 5858 kPa (835 psig) and still provide adequate residence time to permit the partial oxidation reaction to approach equilibrium conditions.

The hot gas exiting the reactor flows directly to a helical coil waste-heat exchanger where process steam is generated. Using the helical coil tubes and proper gas velocity minimizes soot deposit within the heat exchanger. The steam produced in the waste-heat exchangers is generated at least 1034 kPa (150 psi) greater than the reactor pressure so that it can be used directly in the reactor. Waste-heat exchanger designs exist for generating steam at pressures up to 10,443 kPa (1500 psig).

The synthesis gas exiting the waste-heat exchanger has a temperature somewhat greater than the generated steam temperature. An economizer is used to further cool the synthesis gas and preheat the feedwater for the waste-heat recovery unit. The heat duties of the waste-heat exchanger and the economizer are functions of gasifier operating conditions and end-product production.

The soot-laden gas that exits the heat recovery equipment is then scrubbed to remove most of the soot. Some of the soot, which was formed in the gasification vessel, does not exit with the synthesis gas; it is deposited at the bottom of the gasification vessel and remove during periodic shutdown inspections. The synthesis gas that exits the scrubber section contains less than 5 ppm(v) soot. The carbon (soot) removed in the scrubber section is recovered and recycled into the gasifier.

The cooled synthesis gas leaving the scrubber section contains H_2S, CO_2, and COS, which must be removed. The Shell Sulfinol process can be used to remove these gases. This process uses the organic solvents sulfolante (tetrahydrothiophene dioxide) and aqueous alkanolamine to both physically and chemically absorb the unwanted gases. The striped acid gases are then processed to convert the sulfur-containing gases to elemental sulfur by the Claus and SCOT processes. The relatively sulfur-free synthesis gas that exits the Sulfinol reactor is then processed in much the same way as the gases in the catalytic steam-reforming process. The synthesis gas is first sent to the water-gas shift reactor to produce more hydrogen, the resulting CO_2 can be removed by the MEA process, and the gas is finally sent to a methanation unit. A typical feedstock requirement for the partial oxidation process is shown in Table 13.

Table 13. GASIFICATION OF HEAVY OIL
(Shell Gasification Process)
Basis: Production of 2.83×10^6 m^3 (100×10^6 SCF)
H_2/stream day[4]

Feedstock Type		Heavy Fuel Oil
Feedstock Properties		
Gravity, °API		14.2
Specific Gravity, @ 15.5°C		0.97
C/H Weight Ratio		7.49
C/H Atomic Ratio		1.59
Sulfur, wt %		3.50
Ash, wt %		0.07
Feedstock, kg/Stream hr @ 65.55°C (150°F)	39939	(88,050 lbs)
Oxygen Feed,* kg/Stream hr @ 37.77°C (100°F)	42670	(94,070 lbs)
Naphtha Net Usage For Carbon Recovery,** kg/Stream hr	200	440
Fuel Oil For Auxiliary Boiler, kg/Stream hr	4205	(9,270 lbs)

* Expressed as 100% O_2; actual O_2 purity is 99.50 vol %. Pressure is 920 psig.

** Constitutes feed to gasification reactor over and above the heavy fuel oil feedstock. Soot production is recycled 100% to gasification reactor.

Coal Gasification[5,6]

When a solid carbonaceous feedstock such as coal is used, the conversion to hydrogen is similar to the partial oxidation of heavy oils. However, major equipment changes are required to handle the solid feedstock and the resulting unwanted residue such as ash. As in the partial oxidation process, coal gasification also requires an oxygen and steam feed. Unlike methane gas or methanol, coal is a heterogeneous substance which contains trace elements such as sulfur, nitrogen and metals which are detremental to fuel cell operation. These trace elements, which are also found in the heavy oils but to a lesser extent, react with the hydrogen, oxygen, and carbon to form other unwanted substances such as hydrogen sulfide and carbonyl sulfide. Since coal gasification kinetics depend exclusively on temperature and pressure, these elements do not affect the overall conversion to hydrogen. The raw gas from the gasifier contains hydrogen, carbon dioxide, carbon monoxide, water, soot, methane, sulfur compounds and nitrogen. These raw gases must undergo further processing to become an acceptable fuel cell feedstock.

In the first step of gas processing, the raw gas is usually sent to a cyclone separator where unreacted carbon particles and ash are removed. This is generally followed by a waste-heat recovery boiler which generates part of the steam feedstock requirement by heat-exchanger with the hot raw gas. Due to the catalytic process which are also in the downstream processing section, a sulfur removal process is required after the heat recovery step. The recovered sulfur compounds undergo further treatment to recover elemental sulfur.

The gas which exits the sulfur removal step contains predominately hydrogen, carbon monoxide and carbon dioxide. This gas is then sent to the water-gas shift reactor to produce additional hydrogen. In the next step carbon dioxide is removed by chemical or physical solvent washing and finally the gas is sent to a methanation unit where the remaining carbon monoxide and carbon dioxide are reacted to form methane.

Hydrogen Production Costs

The complexity of the above described processes becomes significantly greater as one goes from natural gas or methanol to coal. This complexity is also represented in the higher capital cost of the oil and coal based hydrogen production processes. Figure 7 presents the capital cost range for hydrogen production versus plant size for steam reforming, partial oxidation and coal gasification. As can be seen coal gasification and partial oxidation of oil are significantly more expensive than steam reforming of natural gas or methanol. This is even more so when small scale facilities as those required by the fuel cell are involved. The economy of scale affect can be seen in

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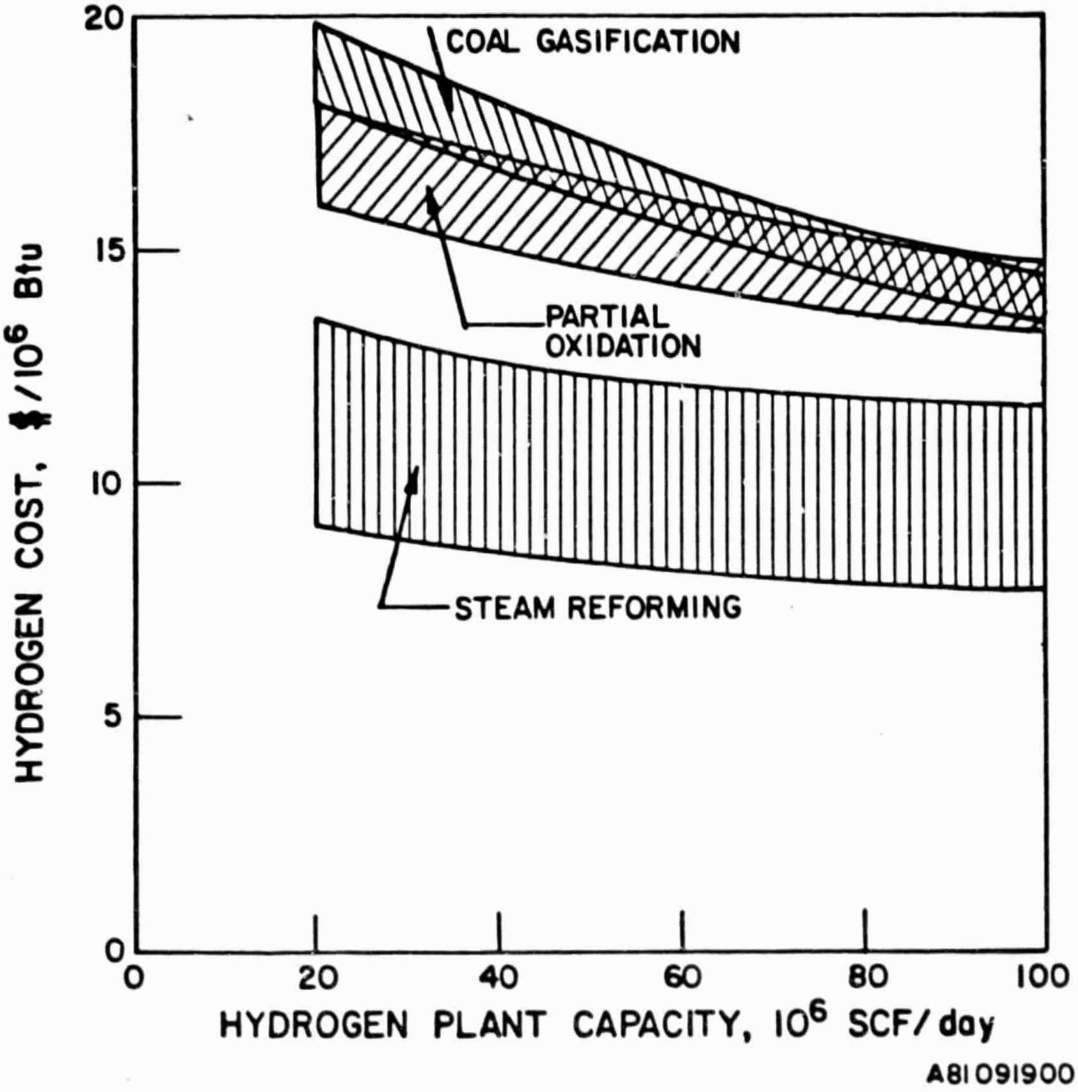

Figure 7. HYDROGEN PRODUCTION COST vs. PLANT CAPACITY, 1981 DOLLARS

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Table 14 for a natural gas steam reformer producing 2832 m^3 to 67,968 m^3/day (100,000 to 2.4 X 10^6 SCF/day) of hydrogen. For the smaller scale plant hydrogen production costs are nearly four times that of the largest plant. Economics of scale would be even more dramatic for the oil based partial oxidation process and the coal gasification case. For this reason the remainder of this study will focus on natural gas and methanol as the primary fuel source. At this point a more detailed analysis (which will appear in the systems section of this report) will need to be made before the final fuel section can be made.

Table 14. ECONOMICS OF HYDROGEN MANUFACTURE IN SMALL REFORMERS, 1982 $

Hydrogen capacity, m^3/day	(2832)	(13,594)	(67,968)
Operating hours per year	7920	7920	7920
Operators per shift	0.15	0.20	0.26
Investment, $ Millions			
Onsite	0.690	1.20	2.39
Offsite	0.173	0.30	0.60
Total	0.8630	1.50	2.99
% contingency in investments	10	10	10
$/MBtu/yr $/$10^3$ kWhr/yr	27.03	9.74	3.90
Working capital, $ thousands	20.7	65.5	277.2
Costs & Charges, $/$10^3$ kWhr product			
Natural gas @ $1.24/$10^3$ kWhr	2.00	2.00	2.00
Utilities	0.04	0.04	0.04
Labor and supervision	1.84	0.52	0.13
Maintenance (4% of onsites)	0.86	0.31	0.13
Plant overhead (2.6% of onsites)	0.58	0.20	0.08
Insurance, property taxes (1.5% of total)	0.40	0.15	0.06
Depreciation (10% onsites, 5% offsites)	2.45	0.87	0.35
Interest on working capital (10%/yr)	0.07	0.04	0.03
Return on investment (20%/yr)	5.40	1.95	0.77
Total Cost, $/$10^3$ kW-hr	13.64	6.08	3.59

APPENDIX B.

Types of Fuel Cells

Fuel cells usually are classified according to the type of electrolyte or ion-conducting media used and the temperature of operation, as shown in Figure 8. The figure also shows fuel quality requirements. A brief introduction to the four fuel cell types follows. These four types of fuel cells include the 1) solid oxide, 2) molten carbonate, 3) acid and 4) alkaline fuel cells.

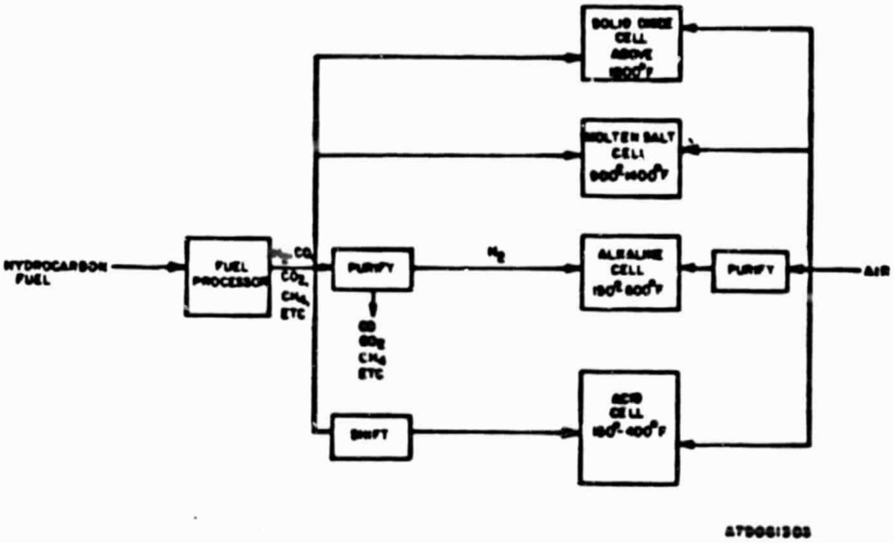

Figure 8. FUEL CELL TYPES

Solid Oxide Fuel Cells. Fuel cells with solid oxide electrolytes operate at temperatures above 800°C (1472°F) using non-noble metal electrodes. The electrolyte is usually yttria- or calcia-stabilized zirconia (ZrO_2), which conducts oxygen ions at the operating temperature. Because of their high operating temperature, solid oxide fuel cells will offer high-quality waste heat that can be utilized either in cogeneration or in a bottoming cycle, hence giving the potential for increasing overall system efficiency. Solid oxide fuel cell technology is still at the research stage. Three of the major problems yet to be solved are as follows:

1. Efficienct contact of electrodes with electrolytes, involving a solid/solid interface, is difficult to achieve.
2. An extremely large number of cells are needed because of the brittle and fragile nature of the solid oxide electrolytes, which limits cell size to a few square centimeters.

3. Materials that are stable at high temperatures are needed, such as cell interconnecting materials and stable electrolytes.

Molten Carbonate Fuel Cells. This type of cell operates at temperatures of 600° to 700°C (1112°F to 1292°F), using impure hydrogen and air, non-noble metal electrodes, and an electrolyte of molten alkali-metal carbonates in a porous ceramic carrier (tile). As with solid oxide cells, high-quality waste heat is available for use in cogeneration or in a bottoming cycle to increase overall system efficiency. Molten salt fuel cells are entering the pilot plant technology stage. The problems associated with molten carbonate fuel cells are as follows:

1. Molten carbonate electrolytes are lost by evaporation, creepage, and corrosion.

2. High-temperature heat exchangers capable of efficient heat transfer at about 1093°C (2000°F) need to be developed.

3. Sulfur removal to about 1 ppm or less is required to prevent damaging of the electrodes.

4. Electrolyte tile integrity and corrosion resistance of cell hardware must be improved for successful long-term (approximately 40,000 hours) operation.

Acid Fuel Cells. Many acid electrolytes have been considered for use in fuel cells because acids do not react with carbon dioxide, which allows the use of impure hydrogen and air. Phosphoric acid (H_3PO_4) fuel cells are the most advanced of all the fuel cell technologies, and they operate at about 190°C (375°F). Pilot plants up to 1 MW have been successfully operated, and a 4.8 MW unit is in startup. The following factors are important in considering the use of H_3PO_4 cells for commercial application —

1. Platinum is poisoned by carbon monoxide (CO), reducing its electrocatalytic activity for hydrogen oxidation. Therefore, CO concentrations of 4% or less in the fuel are desirable.

2. Cathode performance must be improved to improve cell performance.

3. The acid electrolyte is lost because of evaporation and corrosion of cell components. Electrolyte carryover can also cause corrosion problems downstream of the cell.

Alkaline Fuel Cells. Aqueous KOH electrolyte fuel cells operate at lower temperature, 60° to 120°C (140°F to 248°F) and have demonstrated reliability in space applications using pure H_2 and O_2. The major drawback to commercialization of alkaline fuel cells is that the electrolyte reacts with CO_2, limiting the fuel to noncarbonaceous compounds. Even the 0.04% CO_2 in air must be removed. This cannot be accomplished by present technology in a cost-effective manner.

All four fuel cell types are affected to various degrees by impurities such as H_2S, COS, SO_2, HCl, NO_x, and NH_3 in the fuel. Thus, gas cleanup schemes are necessary for the systems.

Basic Principles of Operation

Figures 9 and 10 show a schematic diagram of one repeating element in a stack of fuel cells. Each element is made up of an anode, in which oxidation of the fuel occurs; an electrolyte, to separate the anode and cathode and to conduct the ions between them; and a cathode, in which reduction of the oxidant occurs. The operation of the cell involves many complex mechanisms, which are conceptually simplified in the following discussions.

Phosphoric Acid. Gaseous hydrogen in the fuel diffuses through the porous anode to a reaction site at the electrode (solid)/electrolyte (liquid) interface where it is electrochemically oxidized:

$$H_2 \rightarrow 2H^+ + 2e^- \quad (6)$$

The electrons are transported through the external circuit, and the hydrogen ions are conducted through the electrolyte to the cathode reaction sites. Oxygen, which has diffused through the cathode, reacts with the hydrogen ions and electrons (Equation 7), and the product water diffuses back out of the cathode:

$$1/2\ O_2 + 2H^+ + 2e^- \rightarrow H_2O(g) \quad (7)$$

The net reaction is:

$$H_2 + 1/2\ O_2 \rightarrow H_2O(g) \quad (8)$$

Molten Carbonate. The electrolyte is a mixture of alkali metal carbonates (a molten ionic conductor at the cell operating temperatures of 600° to 700°C) (1112°F to 1292°F) and ceramic particulates that retain the liquid.

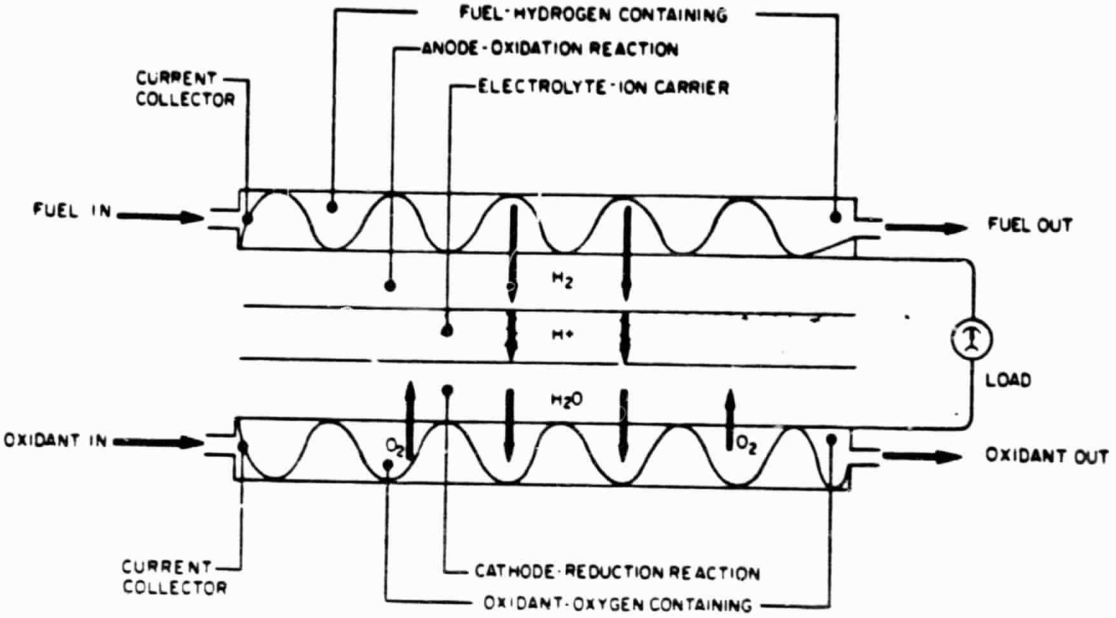

Figure 9. CONCEPTUAL DIAGRAM OF A PHOSPHORIC ACID FUEL CELL

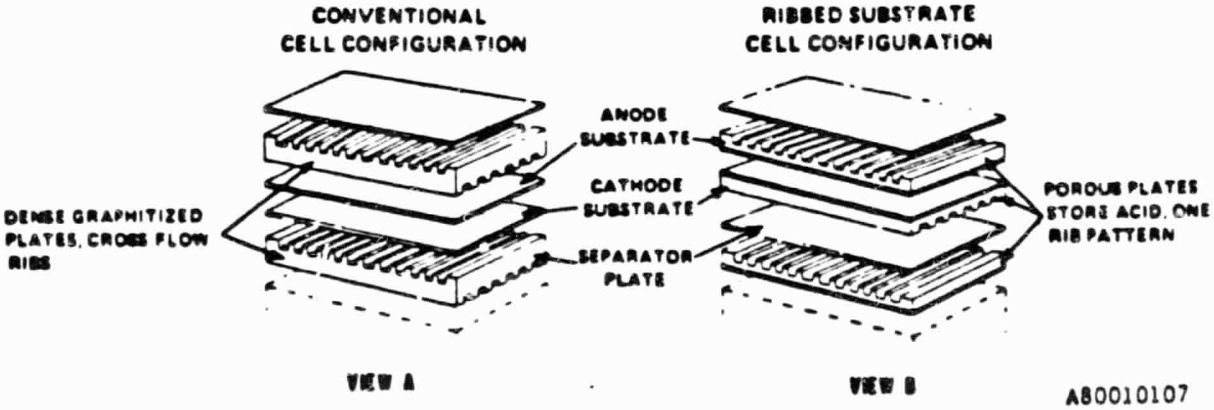

Figure 10. ILLUSTRATIVE CELL CONFIGURATIONS

Fuel enters the anode, and the H_2 reacts* at the anode/electrolyte interface according to —

$$H_2 + CO_3^{=} \rightarrow H_2O + CO_2 + 2e^- \qquad (9)$$

The product H_2O and CO_2 diffuse out of the electrode. At the cathode, Equation 10 takes place:

$$CO_2 + 1/2\ O_2 + 2e^- \rightarrow CO_3^{=} \qquad (10)$$

Note that CO_2 is formed at the anode and consumed at the cathode. For a cost-effective molten carbonate fuel cell system, the CO_2 in the anode effluent must be recycled to the cathode. This can be accomplished by catalytic combustion of the anode effluent to CO_2 and H_2O (plus N_2 in the combustion air), followed by mixing with air to produce the cathode feed.

Applications

The fuel cell is quiet because it has no moving parts. Because the fuel cell is not a combustion device, emissions such as NO_x, CO, and unburned hydrocarbons are not a serious problem. SO_x emissions are not a problem because the sulfur content of the fuel stream to the power plant must be reduced to attain long fuel-cell life. Such low levels of emission, couples with the fuel cell's quiet, water-conserving operation, result in environmental acceptability and siting flexibility.

A single fuel cell normally generates power at approximately 0.5 to 1 volt and can be connected in series stacked with other cells to obtain almost any desired voltage. The current produced is a function of the size (the area) of a single cell. The range of sizes, the modularity, and the load-following capabilities make the fuel-cell system an attractive candidate for power generation in a variety of applications, including on-site and central plants for commercial, industrial, and residential uses.

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[*] Carbon monoxide in the fuel also can be oxidized electrochemically, but the more rapid reaction for the utilization of CO_2 is via conversion to hydrogen by the shift reaction, $CO + H_2O \rightarrow CO_2 + H_2$. The hydrogen formed is then consumed according to Equation 7.

Because of the above-mentioned advantages of fuel cells in addition to their high operating efficiency, they are likely to find applications in the commercial, industrial, and residential areas and in military installations. Fuel cells are also used extensively in the space program. The high-quality heat rejected by molten carbonate and solid oxide fuel cells also makes them potential candidates for cogeneration applications.

State-of-the-Art

The phosphoric acid fuel-cell system is being tested in realistic utility situations. A 4.8-MW power plant is scheduled to operate in New York City as a joint venture between industry and the Department of Energy.

The Department of Energy is also sponsoring a 3-year program at IGT-GE and UTC to test a 1-MW molten carbonate fuel-cell stack by 1983. Demonstration of this technology in realistic utility situations is projected for 1985.

Both of these systems are expected to succeed during the 1980's. The molten carbonate fuel cell system is less developed and will require more work to solve high-temperature-related material problems. For this reason the remainder of this analysis will focus on the phosphoric acid fuel cell. Additional experience on small scale phosphoric acid fuel cells is being gained from a joint Gas Research Institute/DOE/Utilities testing program which will involve 48 field test units at various industrial and commercial installations. This testing program is limited to a 40 kW fuel cell size.

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APPENDIX C.

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BaseCase. Goldstone Complex Annual Energy Consumption - F/Y 81*

RETROFIT LCC WORKSHEET

A. Calculating the Present Value of Energy Costs Without the Retrofit

TYPE	(1) ANNUAL UNITS OF ENERGY PURCHASED	(2) BASE-YEAR ENERGY PRICE PER UNIT	(3) BASE-YEAR ENERGY COSTS	(4) UPW FACTOR	(5) PRESENT VALUE OF ENERGY COSTS
ELECTRICITY	10,684,260 kWhr	$0.084/kWhr	$897,478 BASE CHARGE $____ DEMAND CHARGE $____ TIME OF DAY CHARGE $____ CONTRACT CAPACITY CHARGE $____ OTHER CHARGE COMPONENT	10.33	$9,270,946 $____ $____ $____ $____
OIL	398,430 gal	$7.21/MMBTU	$398,441	14.37	$5,725,594
GAS					
OTHER LPG	46,699 gal	$8.80/MMBTU	$39,246	14.37	$563,963
TOTAL					$15,560,503

B. Calculating Investment Costs for the Existing System Without the Retrofit

✓ (1) Base-Year Resale, Salvage, or Reuse Value of the Existing System to be Replaced $0

(2) Base-Year Renovation Costs for the Existing System if the Retrofit Project is Not Implemented $0

C. Calculating Annually Recurring Nonfuel Operation and Maintenance (O&M) Costs Without the Retrofit

(1) Amount of Annually Recurring Costs in Base Year	(2) UPW Factor	(3) Present Value of Annually Recurring Costs
$18,000	10.59	$190,620

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* Goldstone Complex = Pioneer (DSS-11), Echo (DSS-12), Venus (DSS-13) And Mars (DSS-14) only.

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RETROFIT LCC WORKSHEETS (Continued) Base Case.

D. Calculating Non-Annually Recurring O&M (Non-fuel) Costs, Replacement Costs, and Salvage Value Without the Retrofit.

(1) YEAR IN WHICH EXPENDITURE IS EXPECTED TO OCCUR	(2) AMOUNT OF NON-ANNUALLY RECURRING O&M COSTS (IN BASE-YEAR $)[1]	(3) AMOUNT OF REPLACEMENT COSTS (IN BASE-YEAR $)[1]	(4) AMOUNT OF SALVAGE VALUE (IN BASE-YEAR $)[1]	(5) SPW FACTORS	(6) PRESENT VALUE OF NON-ANNUALLY RECURRING O&M COSTS	(7) PRESENT VALUE OF REPLACEMENT	(8) PRESENT VALUE OF SALVAGE VALUE
TOTAL	☒	☒	☒	☒	0	0	0

E. Calculating TLCC Without the Retrofit

(1) Present Value of Energy Costs $ 15,560,503

(2) Present Value of Investment Costs + $ 0

(3) Present Value of Annually Recurring (Nonfuel) O&M Costs + $ 190,620

(4) Present Value of Nonannually Recurring (Nonfuel) O&M Costs + $ 0

(5) Present Value of Replacement Costs + $ 0

(6) Present Value of Salvage - $ 0

(7) TLCC Without the Retrofit = $ 15,751,123

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[1] For example, if nonannually recurring (nonfuel) O&M costs, replacement costs, or salvage value occur in 1990 and you are using 1980 as the base year, base-year dollars means stating the 1990 costs in 1980 dollars, i.e., without future inflation.

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Case I Centralized Peak Load System - Natural Gas Fueled - Prototype Unit

Retrofit LCC WORKSHEETS (Continued)

Parts F through J Calculate TLCC with the Retrofit

F. Calculating the Present Value of Fuel Costs With the Retrofit

TYPE	(1) ANNUAL UNITS OF ENERGY PURCHASED	(2) BASE-YEAR ENERGY PRICE PER UNIT	(3) BASE-YEAR ENERGY COSTS	(4) UPW FACTOR	(5) PRESENT VALUE OF ENERGY COSTS
ELECTRICITY	________	________	$ ________ BASE CHARGE $ ________ DEMAND CHARGE $ ________ TIME OF DAY CHARGE $ ________ CONTRACT CAPACITY CHARGE $ ________ OTHER CHARGE COMPONENT	____	$ ________ $ ________ $ ________ $ ________ $ ________
OIL					
GAS	1.432×10^{11} Btu	$7.00/MMBTU	$1,002,400	11.95	$11,978,680
OTHER LPG	1.962×10^{9} Btu	$8.80/MMBTU	$ 17268	14.37	$ 248,149
TOTAL					$12,226,829

G. Calculating Investment Costs with the Retrofit

(1) Estimated Actual Investment Costs for the Retrofit Project $5,126,000

(2) Investment Cost Adjustment Factor x 1.0

(3) Adjusted Investment Costs for the Retrofit Project = $5,126,000

(4) Base-Year Renovation Costs for the Existing System if the Retrofit Project is Implemented + 0

(5) Total Adjusted Present Value Investment Costs Attributable to the Retrofit Project = $5,126,000

RETROFIT LCC WORKSHEETS (Continued) CASE I

H. Calculating Annually Recurring (Nonfuel) Operation and Maintenance (O&M) Costs With the Retrofit

(1) Amount of Annually Recurring Costs in Base Year	(2) UPW Factor	(3) Present Value of Annually Recurring Costs
$ 79,672	10.59	$ 843,733

I. Calculating Nonannually Recurring (Nonfuel) O&M Costs, Replacement Costs, and Salvage Value With the Retrofit

(1) YEAR IN WHICH EXPENDITURE IS EXPECTED TO OCCUR	(2) AMOUNT OF NON-ANNUALLY RECURRING O&M COSTS (IN BASE-YEAR $)[1]	(3) AMOUNT OF REPLACEMENT COSTS (IN BASE-YEAR $)[1]	(4) AMOUNT OF SALVAGE VALUE (IN BASE-YEAR $)[1]	(5) SPW FACTORS	(6) PRESENT VALUE OF NON-ANNUALLY RECURRING O&M COSTS	(7) PRESENT VALUE OF REPLACEMENT	(8) PRESENT VALUE OF SALVAGE VALUE
3	0	$250,000	0	0.82	0	$205,000	0
5	0	$975,000	0	0.71	0	$692,250	0
6	0	$250,000	0	0.67	0	$167,500	0
9	0	$250,000	0	0.54	0	$135,000	0
10	0	$975,000	0	0.51	0	$497,250	0
12	0	$250,000	0	0.44	0	$110,000	0
15	0	$1,225,000	0	0.36	0	$441,000	0
18	0	$250,000	0	0.30	0	$75,000	0
TOTAL					0	$2,323,000	0

J. Calculating TLCC With the Retrofit Project

(1) Present Value of Energy Costs		$ 12,226,829
(2) Present Value of Adjusted Investment Costs	+	$ 5,126,000
(3) Present Value of Annually Recurring (Nonfuel) O&M Costs	+	$ 843,733
(4) Present Value of Nonannually Recurring (Nonfuel) O&M Costs	+	$ 0
(5) Present Value of Replacement Costs	+	$ 2,323,000
(6) Present Value of Salvage	-	$ 0
(7) TLCC With the Retrofit Project	=	$ 20,519,562

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[1] See footnote on page 57 for explanation.

[2] Worksheet format is expanded to allow for comparison of the two choices.

RETROFIT LCC WORKSHEETS (Continued) Case I

K. Net Savings or Excess Cost of the Retrofit Project		
(1) TLCC without the Retrofit		$15,751,123
(2) TLCC with the Retrofit	-	$20,519,562
(3) Net Savings (+) or net losses (-)	=	$-4,768,439
L. SIR Calculation		
(1) SIR Numerator		
(a) Energy Cost Savings from the Retrofit		$3,333,674
(b) Change in Nonfuel O&M Costs	-	$-653,113
(c) SIR Numerator	=	$2,680,561
(2) SIR Denominator		
(a) Adjusted Differential Investment Cost		$5,126,000
(b) Change in Replacement Costs	+	$2,323,000
(c) Change in Salvage Value	-	$0
(d) SIR Denominator	=	$7,449,000
(3) SIR for Ranking the Retrofit Project		0.360

Case II Centralized Peak Load System - Methanol Fueled Prototype Unit

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Retrofit LCC WORKSHEETS (Continued)

Parts F through J Calculate TLCC with the Retrofit

F. Calculating the Present Value of Fuel Costs with the Retrofit

TYPE	(1) ANNUAL UNITS OF ENERGY PURCHASED	(2) BASE-YEAR ENERGY PRICE PER UNIT	(3) BASE-YEAR ENERGY COSTS	(4) UPW FACTOR	(5) PRESENT VALUE OF ENERGY COSTS
ELECTRICITY	______	______	$ ______ BASE CHARGE $ ______ DEMAND CHARGE $ ______ TIME OF DAY CHARGE $ ______ CONTRACT CAPACITY CHARGE $ ______ OTHER CHARGE COMPONENT	______	$ ______ $ ______ $ ______ $ ______ $ ______
OIL					
Methanol	1.358 × 10¹¹ Btu	$14.50/MMBTU	$1,969,192	11.95	$23,531,843
OTHER LPG	1.962 × 10⁹ Btu	$8.80/MMBTU	$17,268	14.37	$248,149
TOTAL					$23,779,992

G. Calculating Investment Costs with the Retrofit

(1) Estimated Actual Investment Costs for the Retrofit Project $4,788,000

(2) Investment Cost Adjustment Factor × 1.0

(3) Adjusted Investment Costs for the Retrofit Project = $4,788,000

(4) Base-Year Renovation Costs for the Existing System if the Retrofit Project is Implemented − 0

(5) Total Adjusted Present Value Investment Costs Attributable to the Retrofit Project = $4,788,000

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RETROFIT LCC WORKSHEETS (Continued) Case II

H. Calculating Annually Recurring (Nonfuel) Operation and Maintenance (O&M) Costs With the Retrofit

(1) Amount of Annually Recurring Costs in Base Year	(2) UPW Factor	(3) Present Value of Annually Recurring Costs
$ 79,672	10.59	$ 843,733

I. Calculating Nonannually Recurring (Nonfuel) O&M Costs, Replacement Costs, and Salvage Value With the Retrofit

(1) YEAR IN WHICH EXPENDITURE IS EXPECTED TO OCCUR	(2) AMOUNT OF NON-ANNUALLY RECURRING O&M COSTS (IN BASE-YEAR $)[1]	(3) AMOUNT OF REPLACEMENT COSTS (IN BASE-YEAR $)[1]	(4) AMOUNT OF SALVAGE VALUE (IN BASE-YEAR $)[1]	(5) SPW FACTORS	(6) PRESENT VALUE OF NON-ANNUALLY RECURRING O&M COSTS	(7) PRESENT VALUE OF REPLACEMENT	(8) PRESENT VALUE OF SALVAGE VALUE
3	0	$250,000	0	0.82	0	$205,000	0
5	0	$975,000	0	0.71	0	$692,250	0
6	0	$250,000	0	0.67	0	$167,500	0
9	0	$250,000	0	0.54	0	$135,000	0
10	0	$975,000	0	0.51	0	$497,250	0
12	0	$250,000	0	0.44	0	$110,000	0
15	0	$1,225,000	0	0.36	0	$441,000	0
18	0	$250,000	0	0.30	0	$75,000	0
TOTAL					0	$2,323,000	0

J. Calculating TLCC With the Retrofit Project

(1) Present Value of Energy Costs		$ 23,779,992
(2) Present Value of Adjusted Investment Costs	+	$ 4,788,000
(3) Present Value of Annually Recurring (Nonfuel) O&M Costs	+	$ 843,733
(4) Present Value of Nonannually Recurring (Nonfuel) O&M Costs	+	$ 0
(5) Present Value of Replacement Costs	+	$ 2,323,000
(6) Present Value of Salvage	-	$ 0
(7) TLCC With the Retrofit Project	=	$ 31,734,725

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[1] See footnote on page 57 for explanation.

[2] Worksheet format is expanded to allow for comparison of the two choices.

RETROFIT LCC WORKSHEETS (Continued) CASE II

K. Net Savings or Excess Cost of the Retrofit Project

(1) TLCC without the Retrofit		$15,751,123
(2) TLCC with the Retrofit	-	$31,734,725
(3) Net Savings (+) or net losses (-)	=	$-15,983,602

L. SIR Calculation

(1) SIR Numerator		
(a) Energy Cost Savings from the Retrofit		$-8,219,489
(b) Change in Nonfuel O&M Costs	-	$-653,113
(c) SIR Numerator	=	$-8,872,602
(2) SIR Denominator		
(a) Adjusted Differential Investment Cost		$4,788,000
(b) Change in Replacement Costs	+	$2,323,000
(c) Change in Salvage Value	-	$0
(d) SIR Denominator	=	$7,111,000
(3) SIR for Ranking the Retrofit Project		-1.248

Case III Centralized Peak Load System – Natural Gas Fueled

Commercially Available

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Retrofit LCC WORKSHEETS (Continued)

Parts F through J Calculate TLCC with the Retrofit

F. Calculating the Present Value of Fuel Costs With the Retrofit

TYPE	(1) ANNUAL UNITS OF ENERGY PURCHASED	(2) BASE-YEAR ENERGY PRICE PER UNIT	(3) BASE-YEAR ENERGY COSTS	(4) UPW* FACTOR	(5) PRESENT VALUE OF ENERGY COSTS
ELECTRICITY	______	______	$ ______ BASE CHARGE $ ______ DEMAND CHARGE $ ______ TIME OF DAY CHARGE $ ______ CONTRACT CAPACITY CHARGE $ ______ OTHER CHARGE COMPONENT	______	$ ______ $ ______ $ ______ $ ______ $ ______
OIL					
GAS	1.432×10^{11} Btu	$7.00/MMBTU	$1,002,400	11.95	$11,978,680
OTHER ____	1.962×10^{9} Btu	$8.80/MMBTU	$17,268	14.37	$248,149
TOTAL					$12,226,829

G. Calculating Investment Costs with the Retrofit

(1) Estimated Actual Investment Costs for the Retrofit Project $1,589,000

(2) Investment Cost Adjustment Factor × 1.0

(3) Adjusted Investment Costs for the Retrofit Project = $1,589,000

(4) Base-Year Renovation Costs for the Existing System if the Retrofit Project is Implemented – 0

(5) Total Adjusted Present Value Investment Costs Attributable to the Retrofit Project = $1,589,000

RETROFIT LCC WORKSHEETS (Continued) Case III

H. Calculating Annually Recurring (Nonfuel) Operation and Maintenance (O&M) Costs With the Retrofit

(1) Amount of Annually Recurring Costs in Base Year	(2) UPW Factor	(3) Present Value of Annually Recurring Costs
$ 79,672	10.59	$ 843,733

I. Calculating Nonannually Recurring (Nonfuel) O&M Costs, Replacement Costs, and Salvage Value With the Retrofit

(1) YEAR IN WHICH EXPENDITURE IS EXPECTED TO OCCUR	(2) AMOUNT OF NON-ANNUALLY RECURRING O&M COSTS (IN BASE-YEAR $)[1]	(3) AMOUNT OF REPLACEMENT COSTS (IN BASE-YEAR $)[1]	(4) AMOUNT OF SALVAGE VALUE (IN BASE-YEAR $)[1]	(5) SPW FACTORS	(6) PRESENT VALUE OF NON-ANNUALLY RECURRING O&M COSTS	(7) PRESENT VALUE OF REPLACEMENT	(8) PRESENT VALUE OF SALVAGE VALUE
3	0	$250,000	0	0.82	0	$205,000	
5	0	$302,250	0	0.71	0	$214,598	
6	0	$250,000	0	0.67	0	$167,500	
9	0	$250,000	0	0.54	0	$135,000	
10	0	$302,250	0	0.51	0	$154,148	
12	0	$250,000	0	0.44	0	$110,000	
15	0	$552,250	0	0.36	0	$198,810	
18	0	$250,000	0	0.30	0	$75,000	
TOTAL					0	$1,260,056	0

J. Calculating TLCC With the Retrofit Project

(1) Present Value of Energy Costs		$ 12,226,829
(2) Present Value of Adjusted Investment Costs	+	$ 1,589,000
(3) Present Value of Annually Recurring (Nonfuel) O&M Costs	+	$ 843,733
(4) Present Value of Nonannually Recurring (Nonfuel) O&M Costs	+	$ 0
(5) Present Value of Replacement Costs	+	$ 1,260,056
(6) Present Value of Salvage	-	$ 0
(7) TLCC With the Retrofit Project	=	$ 15,919,618

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[1] See footnote on page 57 for explanation.

[2] Worksheet format is expanded to allow (comparison of the two choices.

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RETROFIT LCC WORKSHEETS (Continued) Case III

K. Net Savings or Excess Cost of the Retrofit Project		
(1) TLCC without the Retrofit		$ 15,751,123
(2) TLCC with the Retrofit	-	$ 15,919,618
(3) Net Savings (+) or net losses (-)	=	$ -168,495
L. SIR Calculation		
(1) SIR Numerator		$ 3,333,674
(a) Energy Cost Savings from the Retrofit		
(b) Change in Nonfuel OM Costs	-	$ -653,113
(c) SIR Numerator	=	$ 2,680,561
(2) SIR Denominator		
(a) Adjusted Differential Investment Cost		$ 1,589,000
(b) Change in Replacement Costs	+	$ 1,260,056
(c) Change in Salvage Value	-	$ 0
(d) SIR Denominator	=	$ 2,849,056
(3) SIR for Ranking the Retrofit Project		0.941

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Case IV. Centralized Peak Load System - Methanol Fueled

Commercially Available

Retrofit LCC WORKSHEETS (Continued)

Parts F through J Calculate TLCC with the Retrofit

F. Calculating the Present Value of Fuel Costs With the Retrofit

TYPE	(1) ANNUAL UNITS OF ENERGY PURCHASED	(2) BASE-YEAR ENERGY PRICE PER UNIT	(3) BASE-YEAR ENERGY COSTS	(4) UPW* FACTOR	(5) PRESENT VALUE OF ENERGY COSTS
ELECTRICITY	______	______	$ ______ BASE CHARGE	______	$ ______
			$ ______ DEMAND CHARGE		$ ______
			$ ______ TIME OF DAY CHARGE		$ ______
			$ ______ CONTRACT CAPACITY CHARGE		$ ______
			$ ______ OTHER CHARGE COMPONENT		$ ______
OIL					
Methanol	1358.x10^9 Btu	$14.50/MMBTU	$1,969,192	11.95	$23,531,843
OTHER LPG	1962 x 10^9 Btu	$8.80/MMBtu	$17,268	14.37	$248,149
TOTAL					$23,779,992

G. Calculating Investment Costs with the Retrofit

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(1) Base-Year Actual Investment Costs for the Retrofit Project $1,484,280

(2) Investment Cost Adjustment Factor x 10

(3) Adjusted Investment Costs for the Retrofit Project = $1,484,280

(4) Base-Year Renovation Costs for the Existing System if the Retrofit Project is Implemented + 0

(5) Total Adjusted Present Value Investment Costs Attributable to the Retrofit Project = $1,484,280

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RETROFIT LCC WORKSHEETS (Continued) Case IV

H. Calculating Annually Recurring (Nonfuel) Operation and Maintenance (O&M) Costs With the Retrofit

(1) Amount of Annually Recurring Costs in Base Year	(2) UPW Factor	(3) Present Value of Annually Recurring Costs
$ 79,672	10.59	$ 843,733

I. Calculating Nonannually Recurring (Nonfuel) O&M Costs, Replacement Costs, and Salvage Value With the Retrofit

(1) YEAR IN WHICH EXPENDITURE IS EXPECTED TO OCCUR	(2) AMOUNT OF NON-ANNUALLY RECURRING O&M COSTS (IN BASE-YEAR $)[1]	(3) AMOUNT OF REPLACEMENT COSTS (IN BASE-YEAR $)[1]	(4) AMOUNT OF SALVAGE VALUE (IN BASE-YEAR $)[1]	(5) SPW FACTORS	(6) PRESENT VALUE OF NON-ANNUALLY RECURRING O&M COSTS	(7) PRESENT VALUE OF REPLACEMENT	(8) PRESENT VALUE OF SALVAGE VALUE
3	0	$250,000	0	0.82	0	$205,000	0
5	0	$302,250	0	0.71	0	$214,598	0
6	0	$250,000	0	0.67	0	$167,500	0
9	0	$250,000	0	0.54	0	$135,000	0
10	0	$302,250	0	0.51	0	$154,148	0
12	0	$250,000	0	0.44	0	$110,000	0
15	0	$552,250	0	0.36	0	$198,810	0
18	0	$250,000	0	0.30	0	$75,000	0
TOTAL					0	$1,260,056	0

J. Calculating TLCC With the Retrofit Project

(1) Present Value of Energy Costs		$ 23,779,992
(2) Present Value of Adjusted Investment Costs	+	$ 1,484,280
(3) Present Value of Annually Recurring (Nonfuel) O&M Costs	+	$ 843,733
(4) Present Value of Nonannually Recurring (Nonfuel) O&M Costs	+	$ 0
(5) Present Value of Replacement Costs	+	$ 1,260,056
(6) Present Value of Salvage	-	$ 0
(7) TLCC With the Retrofit Project	=	$ 27,368,061

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[1] See footnote on page 57 for explanation.

[2] Worksheet format is expanded to allow for comparison of the two choices.

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RETROFIT LCC WORKSHEETS (Continued) Case IV

K. Net Savings or Excess Cost of the Retrofit Project

(1) TLCC without the Retrofit		$ 15,751,123
(2) TLCC with the Retrofit	-	$ 27,368,061
(3) Net Savings (+) or net losses (-)	=	$ -11,616,938

L. SIR Calculation

(1) SIR Numerator		
(a) Energy Cost Savings from the Retrofit		$ -8,219,489
(b) Change in Nonfuel O&M Costs	-	$ -653,113
(c) SIR Numerator	=	$ -8,872,602
(2) SIR Denominator		
(a) Adjusted Differential Investment Cost		$ 1,489,280
(b) Change in Replacement Costs	+	$ 1,260,056
(c) Change in Salvage Value	-	$ 0
(d) SIR Denominator	=	$ 2,744,336
(3) SIR for Ranking the Retrofit Project		-3,233

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Case V. Dispersed Peak Load System - Natural Gas Fueled
Prototype Units

Retrofit LCC WORKSHEETS (Continued)

Parts F through J Calculate TLCC with the Retrofit

F. Calculating the Present Value of Fuel Costs With the Retrofit

TYPE	(1) ANNUAL UNITS OF ENERGY PURCHASED	(2) BASE-YEAR ENERGY PRICE PER UNIT	(3) BASE-YEAR ENERGY COSTS	(4) UPW* FACTOR	(5) PRESENT VALUE OF ENERGY COSTS
ELECTRICITY	______	______	$ ______ BASE CHARGE	______	$ ______
			$ ______ DEMAND CHARGE		$ ______
			$ ______ TIME OF DAY CHARGE		$ ______
			$ ______ CONTRACT CAPACITY CHARGE		$ ______
			$ ______ OTHER CHARGE COMPONENT		$ ______
OIL					
GAS	1.932 x 10^{11} Btu	$7.00/MM Btu	$1,002,400	11.95	$11,978,680
OTHER ___	1.962 x 10^{9} Btu	$8.80/MM Btu	$17,268	14.37	$248,149
TOTAL					$12,226,826

G. Calculating Investment Costs with the Retrofit

(1) Estimated Actual Investment Costs for the Retrofit Project $5,323,154

(2) Investment Cost Adjustment Factor = 1.0

(3) Adjusted Investment Costs for the Retrofit Project = $5,323,154

(4) Base-Year Renovation Costs for the Existing System if the Retrofit Project is Implemented = 0

(5) Total Adjusted Present Value Investment Costs Attributable to the Retrofit Project = $5,323,154

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RETROFIT LCC WORKSHEETS (Continued) CASE V

H. Calculating Annually Recurring (Nonfuel) Operation and Maintenance (O&M) Costs With the Retrofit

(1) Amount of Annually Recurring Costs in Base Year	(2) UPW Factor	(3) Present Value of Annually Recurring Costs
$ 80,123	10.59	$ 848,499

I. Calculating Nonannually Recurring (Nonfuel) O&M Costs, Replacement Costs, and Salvage Value With the Retrofit

(1) YEAR IN WHICH EXPENDITURE IS EXPECTED TO OCCUR	(2) AMOUNT OF NON-ANNUALLY RECURRING O&M COSTS (IN BASE-YEAR $)[1]	(3) AMOUNT OF REPLACEMENT COSTS (IN BASE-YEAR $)[1]	(4) AMOUNT OF SALVAGE VALUE (IN BASE-YEAR $)[1]	(5) SPW FACTORS	(6) PRESENT VALUE OF NON-ANNUALLY RECURRING O&M COSTS	(7) PRESENT VALUE OF REPLACEMENT	(8) PRESENT VALUE OF SALVAGE VALUE
3	0	$250,000	0	0.82	0	$205,000	0
5	0	$1,012,500	0	0.71	0	$718,875	0
6	0	$250,000	0	0.67	0	$167,500	0
9	0	$250,000	0	0.54	0	$135,000	0
10	0	$1,012,500	0	0.51	0	$516,375	0
12	0	$250,000	0	0.44	0	$110,000	0
15	0	$1,262,500	0	0.36	0	$454,500	0
18	0	$250,000	0	0.30	0	$75,000	0
TOTAL					0	$2,382,250	0

J. Calculating TLCC With the Retrofit Project

(1) Present Value of Energy Costs		$ 12,226,826
(2) Present Value of Adjusted Investment Costs	+	$ 5,323,154
(3) Present Value of Annually Recurring (Nonfuel) O&M Costs	+	$ 848,499
(4) Present Value of Nonannually Recurring (Nonfuel) O&M Costs	+	$ 0
(5) Present Value of Replacement Costs	+	$ 2,382,250
(6) Present Value of Salvage	-	$ 0
(7) TLCC With the Retrofit Project	=	$ 20,780,729

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[1] See footnote on page 57 for explanation.

[2] Worksheet format is expanded to allow for comparison of the two choices.

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RETROFIT LCC WORKSHEETS (Continued) CASE V

G. Net Savings or Excess Cost of the Retrofit Project

(1) TLCC without the Retrofit		$ 15,751,123
(2) TLCC with the Retrofit	=	$ 20,780,729
(3) Net Savings (+) or net losses (-)	=	$ -5,029,606

H. SIR Calculation

(1) SIR Numerator		
(a) Energy Cost Savings from the Retrofit		$ 3,333,677
(b) Change in Nonfuel O&M Costs	-	$ -657,879
(c) SIR Numerator	=	$ 2,675,798
(2) SIR Denominator		
(a) Adjusted Differential Investment Cost		$ 5,323,154
(b) Change in Replacement Costs	+	$ 2,382,250
(c) Change in Salvage Value	-	$ 0
(d) SIR Denominator	=	$ 7,705,404
(3) SIR for Ranking the Retrofit Project		0.347

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Case VI. Dispersed Peak Load System - Natural Gas Fueled

Commercially Available Units

Retrofit LCC WORKSHEETS (Continued)

Parts F through J Calculate TLCC with the Retrofit

F. Calculating the Present Value of Fuel Costs With the Retrofit

TYPE	(1) ANNUAL UNITS OF ENERGY PURCHASED	(2) BASE-YEAR ENERGY PRICE PER UNIT	(3) BASE-YEAR ENERGY COSTS	(4) UPW* FACTOR	(5) PRESENT VALUE OF ENERGY COSTS
ELECTRICITY	______	______	$ ______ BASE CHARGE $ ______ DEMAND CHARGE $ ______ TIME OF DAY CHARGE $ ______ CONTRACT CAPACITY CHARGE $ ______ OTHER CHARGE COMPONENT	______	$ ______ $ ______ $ ______ $ ______ $ ______
OIL					
GAS	1432 x 10^{11} Btu	$ 7.00/MMBTU	$1,002,400	11.95	$11,978,680
OTHER LPG	1.962 x 10^{9} Btu	$ [illegible]/MMBTU	$ 17,268	14.37	$ 248,149
TOTAL					$12,226,829

G. Calculating Investment Costs with the Retrofit

(1) Estimated Actual Investment Costs for the Retrofit Project	$1,650,178
(2) Investment Cost Adjustment Factor	x 1.0
(3) Adjusted Investment Costs for the Retrofit Project	= $1,650,178
(4) Base-Year Renovation Costs for the Existing System if the Retrofit Project is Implemented	-. 0
(5) Total Adjusted Present Value Investment Costs Attributable to the Retrofit Project	= $1,650,178

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RETROFIT LCC WORKSHEETS (Continued) Case VI

H. Calculating Annually Recurring (Nonfuel) Operation and Maintenance (O&M) Costs With the Retrofit

(1) Amount of Annually Recurring Costs in Base Year	(2) UPW Factor	(3) Present Value of Annually Recurring Costs
$ 80,123	10.59	$ 848,499

I. Calculating Nonannually Recurring (Nonfuel) O&M Costs, Replacement Costs, and Salvage Value With the Retrofit

(1) YEAR IN WHICH EXPENDITURE IS EXPECTED TO OCCUR	(2) AMOUNT OF NON-ANNUALLY RECURRING O&M COSTS (IN BASE-YEAR $)[1]	(3) AMOUNT OF REPLACEMENT COSTS (IN BASE-YEAR $)[1]	(4) AMOUNT OF SALVAGE VALUE (IN BASE-YEAR $)[1]	(5) SPW FACTORS	(6) PRESENT VALUE OF NON-ANNUALLY RECURRING O&M COSTS	(7) PRESENT VALUE OF REPLACEMENT	(8) PRESENT VALUE OF SALVAGE VALUE
3	0	$250,000	0	0.82	0	$205,000	0
5	0	$313,875	0	0.71	0	$222,851	0
6	0	$250,000	0	0.67	0	$167,500	0
9	0	$250,000	0	0.54	0	$135,000	0
10	0	$313,875	0	0.51	0	$160,076	0
12	0	$250,000	0	0.44	0	$110,000	0
15	0	$563,875	0	0.36	0	$202,995	0
18	0	$250,000	0	0.30	0	$75,000	0
TOTAL					0	$1,278,422	0

J. Calculating TLCC With the Retrofit Project

(1) Present Value of Energy Costs		$ 12,226,829
(2) Present Value of Adjusted Investment Costs	+	$ 1,650,178
(3) Present Value of Annually Recurring (Nonfuel) O&M Costs	+	$ 848,499
(4) Present Value of Nonannually Recurring (Nonfuel) O&M Costs	+	$ 0
(5) Present Value of Replacement Costs	+	$ 1,278,422
(6) Present Value of Salvage	-	$ 0
(7) TLCC With the Retrofit Project	=	$ 16,003,928

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[1] See footnote on page 57 for explanation.

[2] Worksheet format is expanded to allow for comparison of the two choices.

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RETROFIT LCC WORKSHEETS (Continued) CASE VI

K. Net Savings or Excess Cost of the Retrofit Project

(1) TLCC without the Retrofit		$ 15,751,123
(2) TLCC with the Retrofit	-	$ 16,003,928
(3) Net Savings (+) or net losses (-)	=	$ -252,805

L. SIR Calculation

(1) SIR Numerator		
(a) Energy Cost Savings from the Retrofit		$ 3,333,674
(b) Change in Nonfuel O&M Costs	-	$ -657,879
(c) SIR Numerator	=	$ 2,675,795
(2) SIR Denominator		
(a) Adjusted Differential Investment Cost		$ 1,650,178
(b) Change in Replacement Costs	+	$ 1,278,422
(c) Change in Salvage Value	-	$ 0
(d) SIR Denominator	=	$ 2,928,600
(3) SIR for Ranking the Retrofit Project		0.914

Case VII Dispersed Peak Load System - Methanol Fueled Prototype Units

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Retrofit LCC WORKSHEETS (Continued)

Parts F through J Calculate TLCC with the Retrofit

F. Calculating the Present Value of Fuel Costs with the Retrofit

TYPE	(1) ANNUAL UNITS OF ENERGY PURCHASED	(2) BASE-YEAR ENERGY PRICE PER UNIT	(3) BASE-YEAR ENERGY COSTS	(4) UPW* FACTOR	(5) PRESENT VALUE OF ENERGY COSTS
ELECTRICITY	______	______	$ ______ BASE CHARGE	______	$ ______
			$ ______ DEMAND CHARGE		$ ______
			$ ______ TIME OF DAY CHARGE		$ ______
			$ ______ CONTRACT CAPACITY CHARGE		$ ______
			$ ______ OTHER CHARGE COMPONENT		$ ______
OIL					
Methanol	1.358×10^{11} BTU	$14.50/MMBTU	$1,969,192	11.95	$23,531,843
OTHER —	1.962×10^{9} BTU	$8.80/MMBTU	$17,268	14.37	$248,149
TOTAL					$23,779,992

G. Calculating Investment Costs with the Retrofit

(1) Estimated Actual Investment Costs for the Retrofit Project	$4,972,154
(2) Investment Cost Adjustment Factor	× 1.0
(3) Adjusted Investment Costs for the Retrofit Project	= $4,972,154
(4) Base-Year Renovation Costs for the Existing System if the Retrofit Project is Implemented	+ 0
(5) Total Adjusted Present Value Investment Costs Attributable to the Retrofit Project	= $4,972,154

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RETROFIT LCC WORKSHEETS (Continued) Case VII

H. Calculating Annually Recurring (Nonfuel) Operation and Maintenance (O&M) Costs With the Retrofit

(1) Amount of Annually Recurring Costs in Base Year	(2) UPW Factor	(3) Present Value of Annually Recurring Costs
$ 80,123	10.59	$ 848,499

I. Calculating Nonannually Recurring (Nonfuel) O&M Costs, Replacement Costs, and Salvage Value With the Retrofit

(1) YEAR IN WHICH EXPENDITURE IS EXPECTED TO OCCUR	(2) AMOUNT OF NON-ANNUALLY RECURRING O&M COSTS (IN BASE-YEAR $)[1]	(3) AMOUNT OF REPLACEMENT COSTS (IN BASE-YEAR $)[1]	(4) AMOUNT OF SALVAGE VALUE (IN BASE-YEAR $)[1]	(5) SPW FACTORS	(6) PRESENT VALUE OF NON-ANNUALLY RECURRING O&M COSTS	(7) PRESENT VALUE OF REPLACEMENT	(8) PRESENT VALUE OF SALVAGE VALUE
3	0	$250,000	0	0.82	0	$205,000	0
5	0	$1,012,500	0	0.71	0	$718,875	0
6	0	$250,000	0	0.67	0	$167,500	0
9	0	$250,000	0	0.54	0	$135,000	0
10	0	$1,012,500	0	0.51	0	$516,375	0
12	0	$250,000	0	0.44	0	$110,000	0
15	0	$1,262,500	0	0.36	0	$454,500	0
18	0	$250,000	0	0.30	0	$75,000	0
TOTAL					0	$2,382,250	0

J. Calculating TLCC With the Retrofit Project

(1) Present Value of Energy Costs $ 23,779,992
(2) Present Value of Adjusted Investment Costs + $ 4,972,154
(3) Present Value of Annually Recurring (Nonfuel) O&M Costs + $ 848,499
(4) Present Value of Nonannually Recurring (Nonfuel) O&M Costs + $ 0
(5) Present Value of Replacement Costs + $ 2,382,250
(6) Present Value of Salvage − $ 0
(7) TLCC With the Retrofit Project = $ 31,982,895

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[1] See footnote on page 57 for explanation.

[2] Worksheet format is expanded to allow for comparison of the two choices.

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RETROFIT LCC WORKSHEETS (Continued) Case VII

K. Net Savings or Excess Cost of the Retrofit Project

(1) TLCC without the Retrofit		$ 15,751,123
(2) TLCC with the Retrofit	•	$ -31,982,895
(3) Net Savings (+) or net losses (-)	•	$ -16,231,772

L. SIR Calculation

(1) SIR Numerator		
(a) Energy Cost Savings from the Retrofit		$ -8,219,489
(b) Change in Nonfuel OM Costs	•	$ -657,879
(c) SIR Numerator	•	$ -8,877,368
(2) SIR Denominator		
(a) Adjusted Differential Investment Cost		$ 4,972,154
(b) Change in Replacement Costs	•	$ 2,382,250
(c) Change in Salvage Value	•	$ 0
(d) SIR Denominator	•	$ 7,359,409
(3) SIR for Ranking the Retrofit Project		-1,207

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Case VIII. Dispersed Peak Load System - Methanol Fueled

Retrofit LCC WORKSHEETS (Continued)

Commercially Available Units

Parts F through J Calculate TLCC with the Retrofit

F. Calculating the Present Value of Fuel Costs With the Retrofit

TYPE	(1) ANNUAL UNITS OF ENERGY PURCHASED	(2) BASE-YEAR ENERGY PRICE PER UNIT	(3) BASE-YEAR ENERGY COSTS	(4) UPW* FACTOR	(5) PRESENT VALUE OF ENERGY COSTS
ELECTRICITY	________	________	$ ________ BASE CHARGE $ ________ DEMAND CHARGE $ ________ TIME OF DAY CHARGE $ ________ CONTRACT CAPACITY CHARGE $ ________ OTHER CHARGE COMPONENT	________	$ ________ $ ________ $ ________ $ ________ $ ________
OIL					
Methanol	1.358×10^{11} Btu	$14.50/MMBTU	$1,969,192	11.95	$23,531,843
OTHER LPG	1962×10^{9} Btu	$8.80/MMBTU	$17,268	14.37	$248,149
TOTAL					$23,779,992

G. Calculating Investment Costs with the Retrofit

(1) Estimated Actual Investment Costs for the Retrofit Project $1,541,368

(2) Investment Cost Adjustment Factor × 1.0

(3) Adjusted Investment Costs for the Retrofit Project = $1,541,368

(4) Base-Year Renovation Costs for the Existing System if the Retrofit Project is Implemented + 0

(5) Total Adjusted Present Value Investment Costs Attributable to the Retrofit Project = $1,541,368

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RETROFIT LCC WORKSHEETS (Continued) Case VIII

H. Calculating Annually Recurring (Nonfuel) Operation and Maintenance (O&M) Costs With the Retrofit

(1) Amount of Annually Recurring Costs in Base Year	(2) UPW Factor	(3) Present Value of Annually Recurring Costs
$ 80,123	10.59	$ 848,499

I. Calculating Nonannually Recurring (Nonfuel) O&M Costs, Replacement Costs, and Salvage Value With the Retrofit

(1) YEAR IN WHICH EXPENDITURE IS EXPECTED TO OCCUR	(2) AMOUNT OF NON-ANNUALLY RECURRING O&M COSTS (IN BASE-YEAR $)[1]	(3) AMOUNT OF REPLACEMENT COSTS (IN BASE-YEAR $)[1]	(4) AMOUNT OF SALVAGE VALUE (IN BASE-YEAR $)[1]	(5) SPW FACTORS	(6) PRESENT VALUE OF NON-ANNUALLY RECURRING O&M COSTS	(7) PRESENT VALUE OF REPLACEMENT	(8) PRESENT VALUE OF SALVAGE VALUE
3	0	$250,000	0	0.82	0	$205,000	0
5	0	$313,875	0	0.71	0	$222,851	0
6	0	$250,000	0	0.67	0	$167,500	0
9	0	$250,000	0	0.54	0	$135,000	0
10	0	$313,875	0	0.51	0	$160,076	0
12	0	$250,000	0	0.44	0	$110,000	0
15	0	$563,875	0	0.36	0	$202,995	0
18	0	$250,000	0	0.30	0	$75,000	0
TOTAL					0	$1,278,422	0

J. Calculating TLCC With the Retrofit Project

(1) Present Value of Energy Costs	$ 23,779,992
(2) Present Value of Adjusted Investment Costs	+ $ 1,541,368
(3) Present Value of Annually Recurring (Nonfuel) O&M Costs	+ $ 848,499
(4) Present Value of Nonannually Recurring (Nonfuel) O&M Costs	+ $ 0
(5) Present Value of Replacement Costs	+ $ 1,278,422
(6) Present Value of Salvage	- $ 0
(7) TLCC With the Retrofit Project	= $ 27,448,281

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[1] See footnote on page 57 for explanation.

[2] Worksheet format is expanded to allow for comparison of the two choices.

RETROFIT LCC WORKSHEETS (Continued) Case VIII

K. Net Savings or Excess Cost of the Retrofit Project

(1) TLCC without the Retrofit		$ 15,751,123
(2) TLCC with the Retrofit	=	$ -27,448,281
(3) Net Savings (+) or net losses (-)	=	$ -11,697,158

L. SIR Calculation

(1) SIR Numerator		
(a) Energy Cost Savings from the Retrofit		$ -8,219,489
(b) Change in Nonfuel O&M Costs	=	$ -657,879
(c) SIR Numerator	=	$ 8,877,368
(2) SIR Denominator		
(a) Adjusted Differential Investment Cost		$ 1,541,368
(b) Change in Replacement Costs	=	$ 1,278,422
(c) Change in Salvage Value	=	$ 0
(d) SIR Denominator	=	$ 2,819,790
(3) SIR for Ranking the Retrofit Project		- 3.148

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Case IX Centralized Average Load System – Natural Gas Fueled – Commercially Available Unit

Retrofit LCC WORKSHEETS (Continued)

Parts F through J Calculate TLCC with the Retrofit

F. Calculating the Present Value of Fuel Costs with the Retrofit

TYPE	(1) ANNUAL UNITS OF ENERGY PURCHASED	(2) BASE-YEAR ENERGY PRICE PER UNIT	(3) BASE-YEAR ENERGY COSTS	(4) UPW* FACTOR	(5) PRESENT VALUE OF ENERGY COSTS
ELECTRICITY	2,168,540	$0.084/kwhr	$ 182,157 BASE CHARGE	10.33	$ 1,881,685
			$ ____ DEMAND CHARGE		$ ____
			$ ____ TIME OF DAY CHARGE		$ ____
			$ ____ CONTRACT CAPACITY CHARGE		$ ____
			$ ____ OTHER CHARGE COMPONENT		$ ____
OIL					
GAS	1.232 x 10^{11} BTU	$7.00/MMBTU	$862,835	11.95	$10,310,872
OTHER LPG	2.313 x 10^{9} BTU	$8.80/MMBTU	$20,350	14.37	$292,431
TOTAL					$12,484,988

G. Calculating Investment Costs with the Retrofit

(1) Estimated Actual Investment Costs for the Retrofit Project $1,131,151

(2) Investment Cost Adjustment Factor x 1.0

(3) Adjusted Investment Costs for the Retrofit Project = $1,131,151

(4) Base-Year Renovation Costs for the Existing System if the Retrofit Project is Implemented + 0

(5) Total Adjusted Present Value Investment Costs Attributable to the Retrofit Project = $1,131,151

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RETROFIT LCC WORKSHEETS (Continued) Case IX

H. Calculating Annually Recurring (Nonfuel) Operation and Maintenance (O&M) Costs With the Retrofit

(1) Amount of Annually Recurring Costs in Base Year	(2) UPW Factor	(3) Present Value of Annually Recurring Costs
$ 66,156	10.59	$ 700,594

I. Calculating Nonannually Recurring (Nonfuel) O&M Costs, Replacement Costs, and Salvage Value With the Retrofit

(1) YEAR IN WHICH EXPENDITURE IS EXPECTED TO OCCUR	(2) AMOUNT OF NON-ANNUALLY RECURRING O&M COSTS (IN BASE-YEAR $)[1]	(3) AMOUNT OF REPLACEMENT COSTS (IN BASE-YEAR $)[1]	(4) AMOUNT OF SALVAGE VALUE (IN BASE-YEAR $)[1]	(5) SPW FACTORS	(6) PRESENT VALUE OF NON-ANNUALLY RECURRING O&M COSTS	(7) PRESENT VALUE OF REPLACEMENT	(8) PRESENT VALUE OF SALVAGE VALUE
3	0	$165,000	0	0.82	0	$135,300	0
5	0	$197,625	0	0.71	0	$140,313	0
6	0	$165,000	0	0.67	0	$110,550	0
9	0	$165,000	0	0.54	0	$89,100	0
10	0	$197,625	0	0.51	0	$100,788	0
12	0	$165,000	0	0.44	0	$72,600	0
15	0	$362,625	0	0.36	0	$130,545	0
18	0	$165,000	0	0.30	0	$49,500	0
TOTAL					0	$828,696	0

J. Calculating TLCC With the Retrofit Project

(1) Present Value of Energy Costs $ 12,484,988
(2) Present Value of Adjusted Investment Costs + $ 1,131,151
(3) Present Value of Annually Recurring (Nonfuel) O&M Costs + $ 700,594
(4) Present Value of Nonannually Recurring (Nonfuel) O&M Costs + $ 0
(5) Present Value of Replacement Costs + $ 828,696
(6) Present Value of Salvage - $ 0
(7) TLCC With the Retrofit Project = $ 15,145,429

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[1] See footnote on page 57 for explanation.

[2] Worksheet format is expanded to allow for comparison of the two choices.

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RETROFIT LCC WORKSHEETS (Continued) Case IX

G. Net Savings or Excess Cost of the Retrofit Project

(1) TLCC without the Retrofit		$ 15,751,123
(2) TLCC with the Retrofit	-	$ 15,145,429
(3) Net Savings (+) or net losses (-)	=	+ $ 605,694

H. SIR Calculation

(1) SIR Numerator		
(a) Energy Cost Savings from the Retrofit		$ 3,075,515
(b) Change in Nonfuel O&M Costs	-	$ -509,974
(c) SIR Numerator	=	$ 2,565,541
(2) SIR Denominator		
(a) Adjusted Differential Investment Cost		$ 1,131,151
(b) Change in Replacement Costs	+	$ 828,696
(c) Change in Salvage Value	-	$ 0
(d) SIR Denominator	=	$ 1,959,847
(3) SIR for Ranking the Retrofit Project		+ 1.309

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Case X Centralized Average Load System - Natural Gas Fueled

Prototype Unit

Retrofit LCC WORKSHEETS (Continued)

Parts F through J Calculate TLCC with the Retrofit

F. Calculating the Present Value of Fuel Costs With the Retrofit

TYPE	(1) ANNUAL UNITS OF ENERGY PURCHASED	(2) BASE-YEAR ENERGY PRICE PER UNIT	(3) BASE-YEAR ENERGY COSTS	(4) UPW FACTOR	(5) PRESENT VALUE OF ENERGY COSTS
ELECTRICITY	2,168,540	$0.084/kWhr	$ 182,157 BASE CHARGE	10.33	$ 1,881,685
			$ ____ DEMAND CHARGE		$ ____
			$ ____ TIME OF DAY CHARGE		$ ____
			$ ____ CONTRACT CAPACITY CHARGE		$ ____
			$ ____ OTHER CHARGE COMPONENT		$ ____
OIL					
GAS	1.232 x10^{11} BTU	$7.00/MMBTU	$862,835	11.95	$ 10,310,872
OTHER LPG	2.313x10^{9} BTU	$8.80/MMBTU	$20,350	14.37	$ 292,431
TOTAL					$12,484,988

G. Calculating Investment Costs with the Retrofit

(1) Estimated Actual Investment Costs for the Retrofit Project	$3,648,875
(2) Investment Cost Adjustment Factor	x 1.0
(3) Adjusted Investment Costs for the Retrofit Project	= $3,648,875
(4) Base-Year Renovation Costs for the Existing System if the Retrofit Project is Implemented	+ 0
(5) Total Adjusted Present Value Investment Costs Attributable to the Retrofit Project	= $3,648,875

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RETROFIT LCC WORKSHEETS (Continued) Case X

H. Calculating Annually Recurring (Nonfuel) Operation and Maintenance (O&M) Costs With the Retrofit

(1) Amount of Annually Recurring Costs in Base Year	(2) UPW Factor	(3) Present Value of Annually Recurring Costs
$ 66,156	10.59	$ 700,594-

I. Calculating Nonannually Recurring (Nonfuel) O&M Costs, Replacement Costs, and Salvage Value With the Retrofit

(1) YEAR IN WHICH EXPENDITURE IS EXPECTED TO OCCUR	(2) AMOUNT OF NON-ANNUALLY RECURRING O&M COSTS (IN BASE-YEAR $)[1]	(3) AMOUNT OF REPLACEMENT COSTS (IN BASE-YEAR $)[1]	(4) AMOUNT OF SALVAGE VALUE (IN BASE-YEAR $)[1]	(5) SPW FACTORS	(6) PRESENT VALUE OF NON-ANNUALLY RECURRING O&M COSTS	(7) PRESENT VALUE OF REPLACEMENT	(8) PRESENT VALUE OF SALVAGE VALUE
3	0	$165,000	0	0.82	0	$135,300	0
5	0	$637,500	0	0.71	0	$452,625	0
6	0	$165,000	0	0.67	0	$110,550	0
9	0	$165,000	0	0.54	0	$89,100	0
10	0	$637,500	0	0.51	0	$325,125	0
12	0	$165,000	0	0.44	0	$72,600	0
15	0	$802,500	0	0.36	0	$288,900	0
18	0	$165,000	0	0.30	0	$49,500	0
TOTAL					0	$1,523,700	0

J. Calculating TLCC With the Retrofit Project

(1) Present Value of Energy Costs		$ 12,484,988
(2) Present Value of Adjusted Investment Costs	+	$ 3,648,875
(3) Present Value of Annually Recurring (Nonfuel) O&M Costs	+	$ 700,594
(4) Present Value of Nonannually Recurring (Nonfuel) O&M Costs	+	$ 0
(5) Present Value of Replacement Costs	+	$ 1,523,700
(6) Present Value of Salvage	-	$ 0
(7) TLCC With the Retrofit Project	=	$ 18,358,157

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[1] See footnote on page 57 for explanation.

[2] Worksheet format is expanded to allow for comparison of the two choices.

RETROFIT LCC WORKSHEETS (Continued) Case X

K. Net Savings or Excess Cost of the Retrofit Project

(1) TLCC without the Retrofit		$ 15,751,123
(2) TLCC with the Retrofit	=	$ 18,358,157
(3) Net Savings (+) or net losses (-)	=	-$ 2,607,034

L. SIR Calculation

(1) SIR Numerator		
(a) Energy Cost Savings from the Retrofit		$ 3,075,515
(b) Change in Nonfuel O&M Costs	=	$ -509,974
(c) SIR Numerator	=	$ 2,565,541
(2) SIR Denominator		
(a) Adjusted Differential Investment Cost		$ 3,648,875
(b) Change in Replacement Costs	=	$ 1,523,700
(c) Change in Salvage Value	=	$ 0
(d) SIR Denominator	=	$ 5,172,575
(3) SIR for Ranking the Retrofit Project		0.496

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Case XI Centralized Average Load System – Methanol Fueled

Retrofit LCC WORKSHEETS (Continued)

Parts F through J Calculate TLCC with the Retrofit

Commercially Available Unit

F. Calculating the Present Value of Fuel Costs With the Retrofit

TYPE	(1) ANNUAL UNITS OF ENERGY PURCHASED	(2) BASE-YEAR ENERGY PRICE PER UNIT	(3) BASE-YEAR ENERGY COSTS	(4) UPW* FACTOR	(5) PRESENT VALUE OF ENERGY COSTS
ELECTRICITY	2,168,540	$0.084/kWhr	$ 182,157 BASE CHARGE	10.33	$ 1,881,685
			$ ______ DEMAND CHARGE		$ ______
			$ ______ TIME OF DAY CHARGE		$ ______
			$ ______ CONTRACT CAPACITY CHARGE		$ ______
			$ ______ OTHER CHARGE COMPONENT		$ ______
OIL					
Methanol	1.169×10^{11} BTU	$14.50/MM BTU	$1,694,946	11.95	$20,254,612
OTHER LPG	2.313×10^{9} BTU	$8.80/MM BTU	$20,350	14.37	$292,431
TOTAL					$22,428,728

G. Calculating Investment Costs with the Retrofit

(1) Estimated Actual Investment Costs for the Retrofit Project $1,056,565

(2) Investment Cost Adjustment Factor × 1.0

(3) Adjusted Investment Costs for the Retrofit Project = $1,056,565

(4) Base-Year Renovation Costs for the Existing System if the Retrofit Project is Implemented − 0

(5) Total Adjusted Present Value Investment Costs Attributable to the Retrofit Project = $1,056,565

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RETROFIT LCC WORKSHEETS (Continued) Case XI

H. Calculating Annually Recurring (Nonfuel) Operation and Maintenance (O&M) Costs With the Retrofit

(1) Amount of Annually Recurring Costs in Base Year	(2) UPW Factor	(3) Present Value of Annually Recurring Costs
$ 66,156	10.59	$ 700,594

I. Calculating Nonannually Recurring (Nonfuel) O&M Costs, Replacement Costs, and Salvage Value With the Retrofit

(1) YEAR IN WHICH EXPENDITURE IS EXPECTED TO OCCUR	(2) AMOUNT OF NON-ANNUALLY RECURRING O&M COSTS (IN BASE-YEAR $)[1]	(3) AMOUNT OF REPLACEMENT COSTS (IN BASE-YEAR $)[1]	(4) AMOUNT OF SALVAGE VALUE (IN BASE-YEAR $)[1]	(5) SPW FACTORS	(6) PRESENT VALUE OF NON-ANNUALLY RECURRING O&M COSTS	(7) PRESENT VALUE OF REPLACEMENT	(8) PRESENT VALUE OF SALVAGE VALUE
3	0	$165,000	0	0.82	0	$135,300	0
5	0	$197,625	0	0.71	0	$140,313	0
6	0	$165,000	0	0.67	0	$110,550	0
9	0	$165,000	0	0.54	0	$89,100	0
10	0	$197,625	0	0.51	0	$100,788	0
12	0	$165,000	0	0.44	0	$72,600	0
15	0	$362,625	0	0.36	0	$130,545	0
18	0	$165,000	0	0.30	0	$49,500	0
TOTAL					0	$928,696	0

J. Calculating TLCC With the Retrofit Project

(1) Present Value of Energy Costs $ 22,428,728
(2) Present Value of Adjusted Investment Costs + $ 1,056,565
(3) Present Value of Annually Recurring (Nonfuel) O&M Costs + $ 700,594
(4) Present Value of Nonannually Recurring (Nonfuel) O&M Costs + $ 0
(5) Present Value of Replacement Costs + $ 828,696
(6) Present Value of Salvage - $ 0
(7) TLCC With the Retrofit Project = $ 25,014,583

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[1] See footnote on page 57 for explanation.

[2] Worksheet format is expanded to allow for comparison of the two choices.

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RETROFIT LCC WORKSHEETS (Continued) Case XI

H. Net Savings or Excess Cost of the Retrofit Project

(1) TLCC without the Retrofit $ 15,751,123

(2) TLCC with the Retrofit = $ 25,014,583

(3) Net Savings (+) or net losses (-) = $ -9,263,960

I. SIR Calculation

(1) SIR Numerator

(a) Energy Cost Savings from the Retrofit $ -6,868,225

(b) Change in Nonfuel OM Costs = $ -509,974

(c) SIR Numerator = $ -7,378,199

(2) SIR Denominator

(a) Adjusted Differential Investment Cost $ 1,056,565

(b) Change in Replacement Costs = $ 828,696

(c) Change in Salvage Value - $ 0

(d) SIR Denominator = $ 1,885,261

(3) SIR for Ranking the Retrofit Project -3.914

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Case XII Centralized Average Load System - Methanol Fueled

Retrofit LCC WORKSHEETS (Continued)

Parts F through J Calculate TLCC with the Retrofit

Prototype Unit

F. Calculating the Present Value of Fuel Costs With the Retrofit

TYPE	(1) ANNUAL UNITS OF ENERGY PURCHASED	(2) BASE-YEAR ENERGY PRICE PER UNIT	(3) BASE-YEAR ENERGY COSTS	(4) UPW* FACTOR	(5) PRESENT VALUE OF ENERGY COSTS
ELECTRICITY	2,168,540	$0.084/KWhr	$ 182,157 BASE / CHARGE	10.33	$ 1,881,685
			$ ____ DEMAND CHARGE		$ ____
			$ ____ TIME OF DAY CHARGE		$ ____
			$ ____ CONTRACT CAPACITY CHARGE		$ ____
			$ ____ OTHER CHARGE COMPONENT		$ ____
OIL					
Methanol	1.169x10^11 BTU	$14.50/MMBTU	$1,694,946	11.95	$20,254,612
OTHER LPG	2.313x10^9 BTU	$8.80/MMBTU	$20,350	14.37	$292,431
TOTAL					$22,428,728

G. Calculating Investment Costs with the Retrofit

(1) Estimated Actual Investment Costs for the Retrofit Project $3,408274

(2) Investment Cost Adjustment Factor × 1.0

(3) Adjusted Investment Costs for the Retrofit Project = $3,408,274

(4) Base-Year Renovation Costs for the Existing System if the Retrofit Project is Implemented + 0

(5) Total Adjusted Present Value Investment Costs Attributable to the Retrofit Project = $3,408,274

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RETROFIT LCC WORKSHEETS (Continued) Case XII

H. Calculating Annually Recurring (Nonfuel) Operation and Maintenance (O&M) Costs With the Retrofit

(1) Amount of Annually Recurring Costs in Base Year	(2) UPW Factor	(3) Present Value of Annually Recurring Costs
$ 66,156	10.59	$ 700,594

I. Calculating Nonannually Recurring (Nonfuel) O&M Costs, Replacement Costs, and Salvage Value With the Retrofit

(1) YEAR IN WHICH EXPENDITURE IS EXPECTED TO OCCUR	(2) AMOUNT OF NON-ANNUALLY RECURRING O&M COSTS (IN BASE-YEAR $)[1]	(3) AMOUNT OF REPLACEMENT COSTS (IN BASE-YEAR $)[1]	(4) AMOUNT OF SALVAGE VALUE (IN BASE-YEAR $)[1]	(5) SPW FACTORS	(6) PRESENT VALUE OF NON-ANNUALLY RECURRING O&M COSTS	(7) PRESENT VALUE OF REPLACEMENT	(8) PRESENT VALUE OF SALVAGE VALUE
3	0	$165,000	0	0.82	0	$135,300	0
5	0	$637,500	0	0.71	0	$452,625	0
6	0	$165,000	0	0.67	0	$110,550	0
9	0	$165,000	0	0.54	0	$89,100	0
10	0	$637,500	0	0.51	0	$325,125	0
12	0	$165,000	0	0.44	0	$72,600	0
15	0	$802,500	0	0.36	0	$288,900	0
18	0	$165,000	0	0.30	0	$49,500	0
TOTAL					0	$1,523,700	0

J. Calculating TLCC With the Retrofit Project

(1) Present Value of Energy Costs		$ 22,428,728
(2) Present Value of Adjusted Investment Costs	+	$ 3,408,274
(3) Present Value of Annually Recurring (Nonfuel) O&M Costs	+	$ 700,594
(4) Present Value of Nonannually Recurring (Nonfuel) O&M Costs	+	$ 0
(5) Present Value of Replacement Costs	+	$ 1,523,700
(6) Present Value of Salvage	-	$ 0
(7) TLCC With the Retrofit Project	=	$ 28,061,296

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[1] See footnote on page 57 for explanation.

[2] Worksheet format is expanded to allow for comparison of the two choices.

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RETROFIT LCC WORKSHEETS (Continued) Case XII

K. Net Savings or Excess Cost of the Retrofit Project

(1) TLCC without the Retrofit		$	15,751,123
(2) TLCC with the Retrofit	-	$	28,061,296
(3) Net Savings (+) or net losses (-)	=	$	-12,310,173

L. SIR Calculation

(1) SIR Numerator			
(a) Energy Cost Savings from the Retrofit		$	-6,868,225
(b) Change in Nonfuel O&M Costs	+	$	-509,974
(c) SIR Numerator	=	$	-7,378,199
(2) SIR Denominator			
(a) Adjusted Differential Investment Cost		$	3,408,274
(b) Change in Replacement Costs	+	$	1,523,700
(c) Change in Salvage Value	-	$	0
(d) SIR Denominator	=	$	4,931,974
(3) SIR for Ranking the Retrofit Project			-1.496