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**ENERGY CONVERSION ALTERNATIVES STUDY
- ECAS -
WESTINGHOUSE PHASE I FINAL REPORT
Volume XII - FUEL CELLS**

by

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WESTINGHOUSE ELECTRIC CORPORATION RESEARCH LABORATORIES

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16. Abstract A parametric assessment of four fuel cell power systems -- based on phosphoric acid, potassium hydroxide, molten carbonate, and stabilized zirconia -- has shown that the most important parameters from electricity-cost reduction and/or efficiency improvement standpoints are fuel cell useful life and power density, use of a waste-heat recovery system, and fuel type. Typical capital costs, overall energy efficiencies (based on the heating value of the coal used to produce the power plant fuel), and electricity costs are: phosphoric acid \$350-450/kWe, 24-29%, and 11.7 to 13.9 mills/MJ (42 to 50 mills/kWh); alkaline \$450-700/kWe, 26-31%, and 12.8 to 16.9 mills/MJ (46 to 61 mills/kWh); molten carbonate \$480-650/kWe, 32-46%, and 10.6 to 19.4 mills/MJ (38 to 70 mills/kWh); stabilized zirconia \$420-950/kWe, 26-53%, and 9.7 to 16.9 mills/MJ (35 to 61 mills/kWh). Three types of fuel cell power plants -- solid electrolyte with steam bottoming, molten carbonate with steam bottoming, and solid electrolyte with an integrated coal gasifier -- are recommended for further study.		
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- L. E. Brecher, who provided information on fuel conditioning equipment performance and cost.
- J. T. Brown and E. S. Busselli, who developed and applied costing procedures.
- F. J. Sisk, who provided information on steam bottoming plants.
- E. J. Vidt, who developed information on oxygen plant costs.
- P. Wood, who provided information on the necessary power condition apparatus, and its performance and cost.
- C. T. McCreedy and S. M. Scherer of Chas. T. Main, Inc. of Boston, who prepared the balance of plant description and costing, site drawings, and provided consultation on plant island arrangements and plant constructability.

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SUMMARY

Four fuel cell power systems, differentiated by electrolyte type, have been investigated from cost and efficiency standpoints. For the phosphoric acid system, operating at 190°C, a power plant efficiency, an overall energy efficiency (based on the limiting value of the coal employed in the production of the power plant fuel), a capital cost and a cost of electricity were calculated for each of sixteen points in the parametric assessment. Similar calculations were performed for aqueous alkaline fuel cell power plants at 343°K [70°C (158°F)] (16 points), molten carbonate plants at 923°K [650°C (1202°F)] (17 points), and stabilized zirconia plants at 1273°K [1000°C (1832°F)] (20 points).

In parametric assessments, the following parameters were varied: useful life and rating of the fuel cell subsystem, fuel cell power density and electrolyte thickness, fuel and oxidant types, performance degradation over the useful life of the fuel cell subsystem, anode and cathode catalyst loadings in the acid and alkaline systems, and temperature of operation and use of waste-heat recovery systems in the molten carbonate and stabilized zirconia power systems. Four of these -- fuel cell useful life and power density, use of a waste-heat recovery system, and fuel type -- proved to be of particular importance in efficiency improvement and/or electricity-cost reduction.

Typical capital costs, overall energy efficiencies, and electricity costs of fuel cell power plants were found to be as follows:

- Phosphoric acid \$350-450/kWe, 24-29%, and 11.7 to 13.9 mills/MJ (42 to 50 mills/kWh)
- Alkaline \$450-700/kWe, 26-31%, and 12.8 to 16.9 mills/MJ (46 to 61 mills/kWh)
- Molten carbonate \$480-650/kWe, 32-46%, and 10.6 to 19.4 mills/MJ (38 to 70 mills/kWh)

- Stabilized zirconia \$420-950/kWe, 26-53%, and 9.7 to 16.9 mills/MJ (35 to 61 mills/kWh).

Projections as to the lowest possible cost of electricity (in mills/kWh) for the acid, alkaline, carbonate, and zirconia systems are mid to high 30's, low 40's, low 30's, and high 20's, respectively.

Three types of fuel cell power plants are recommended for further study in the conceptual design (Task II) and implementation assessment (Task III) phases of the Energy Conversion Alternatives Study. These are: (i) solid-electrolyte plant with steam bottoming; (ii) molten carbonate plant with steam bottoming; and (iii) solid electrolyte plant with an integrated coal-gasification reactor for waste-heat recovery (the Westinghouse Fuel Cell Power System).

13. FUEL CELLS

13.1 State of the Art

Fuel cell power-generation systems may be grouped in four distinct classes differentiated by electrolyte type. These are:

- Aqueous acid, which encompasses systems based on phosphoric and sulfuric acids, and the solid polymer electrolyte
- Aqueous alkali, based on potassium hydroxide and buffered carbonate-bicarbonate solutions
- Molten salt, specifically molten carbonates
- High-temperature solid electrolyte, specifically stabilized zirconia.

NASA has specified that the minimum power plant rating to be considered in this study should be 25 MWe. No fuel cell power system of even two orders of magnitude lower in rating has been constructed or operated, so there is no utility experience to guide this investigation. All that is available is the self-serving advocacy of their own systems by the corporations engaged in fuel cell research. While there are, of necessity, some constraints on the published estimates of system efficiencies, there appears to be no such restraint shown in the projections of the expected cost and performance of the fuel cell subsystems.

A fuel cell power plant will consist, in general, of fuel-processing and power-conditioning equipment in addition to the fuel cell subsystem. Most of the effort of this study was directed toward the fuel cell component because of the greater degree of uncertainty with respect to its cost and performance. Further, as no fuel cell power plant concept has been conclusively demonstrated as optimal from the standpoints

of technical feasibility or economic desirability, four base cases are considered, rather than the minimum of two mandated by the contract work statement, with one base case selected from each of the four electrolyte-type categories described above.

The scientific literature pertaining to fuel cells was reviewed in order to facilitate this selection procedure and, particularly, to assist in establishing a meaningful framework of parametric values. Within each of the four general areas the fuel cell power systems were considered from the standpoint of their suitability for use in central-station power generation. Among the more important considerations were the voltage efficiency, demonstrated cell and battery useful lifetimes, life-limiting processes, quantities of noble-metal catalysts required, problems posed by the use of coal-derived fuels, and the state of the art with respect to the engineering of the overall power systems. The results of the literature survey in each area are outlined in the following subsections.

13.1.1 Aqueous Acid Fuel Cells

Systems based on three distinct electrolyte types have been considered in this area:

- Phosphoric acid fuel cells, which operate in the temperature range 423 to 463°K (302 to 374°F)
- Solid polymer electrolyte fuel cells, operating at 348 or 423°K (167 or 302°F), depending on whether air or oxygen is employed as the oxidant
- Sulfuric acid fuel cells which operate at approximately 333°K (140°F).

The Power Systems Division of United Technologies Corporation (formerly Pratt and Whitney Aircraft Division of United Aircraft) is the undisputed leader in acid fuel cell systems. It has field-tested complete power systems of up to 40 kW capacity and is currently building 26 MW systems (FCG-1) for a number of utilities. The preferred system

employs immobilized phosphoric acid electrolyte at temperatures between 423 and 463°K (302 and 374°F). The higher temperatures in this range are desirable mainly because of air electrode (cathode) activity, but they also serve to accelerate life-limiting processes such as platinum electrocatalyst recrystallization and degeneration of the phosphoric acid matrix. One problem with this system is typical of all acid systems—fuel electrode deterioration due to carbon monoxide blockage of active sites on the anode electrocatalyst, thus denying them to the electrochemically more active hydrogen. This problem is a good deal less severe here than in other acid systems because the fuel electrode functions adequately, provided the carbon monoxide concentration in the fuel gas is maintained at 0.5% or below.

In a recent ERDA-directed assessment study of devices for the generation of electricity from stored hydrogen, conducted at Argonne National Laboratory (ANL), an efficiency of 38% (based on the higher heating value of the fuel gas) was quoted by United Technologies personnel for a fuel consisting of a typical reformer effluent gas, shifted (and perhaps methanated) to meet the carbon monoxide concentration specifications quoted above (Reference 13.1). The selling price (FOB factory) of the fuel cell power system, including the inverter, was given as \$225/kW of installed capacity. If an allowance for the inverter system of \$40/kW (Reference 13.2) is subtracted from this total, the final cell subsystem cost is \$185/kW. The targeted (but as yet unachieved) useful life of the power system is 144 Ms (40,000 hr) of operation with a 5% loss in efficiency in that period.

Over the past 630 Ms (20 yr), the General Electric Company has developed a fuel cell based on an electrolyte consisting of a porous film of a fluorocarbon with chemically-bound sulfonic acid groups (Reference 13.3). The solid polymer electrolyte (SPE), developed and marketed by DuPont under the tradename NAFION, is reportedly extremely stable in the fuel cell environment. With near ambient-pressure air as oxidant, the optimum temperature of operation is approximately 348°K (167°F). Because the air (and fuel) streams must be presaturated to

ensure membrane stability, higher temperatures cause excessive dilution of the oxygen in the air stream by water vapor. When oxygen is employed, operational temperatures up to 423°K (302°F) are possible.

State-of-the-art performances are 180 mW/cm² (167 W/ft²) of active area at 0.66 V per cell (44% HHV) with air at 348°K (167°F) and 470 mW/cm² (437 W/ft²) at 0.75 V per cell (51% HHV) with oxygen at 423°K (302°F) (Reference 13.4). This work has also shown that at 348°K (167°F), with a fuel gas containing 0.3% carbon monoxide, the current noble metal catalyst will not tolerate this carbon monoxide level without a prohibitive performance loss, even at noble metal loadings in the anode of approximately 4 mg/cm² (5.69×10^{-5} lb/in²) (Reference 13.4). This phenomenon, discussed above, is obviously an even severer limitation on this power system. The state-of-the-art fuel cell manufacturing cost (not selling price) is estimated to be approximately \$400/kW for the system operating on hydrogen and air. Of this total, approximately \$260/kW is attributable to the noble-metal loadings and approximately \$120/kW to the membrane. The economic viability of this system is seen, therefore, to depend on significant breakthroughs in the areas of noble-metal loading reductions and membrane substitution or cost reduction, as well as the development of methods for the lowering of the carbon monoxide content substantially below concentrations of 0.3%. A carbon monoxide concentration of 10 ppm is considered to be tolerable (Reference 13.3).

Sulfuric acid-based systems, in which the anode electrocatalyst is tungsten carbide, do not suffer from the carbon monoxide poisoning problem described above for the phosphoric acid and SPE systems (Reference 13.5). As is usual in sulfuric acid systems which operate typically at approximately 333°K (140°F), however, the major problem is the low activity of the cathode even when it is operated on oxygen. A further problem arising with long-time operation is the need to reject the water formed in the cathode reaction.

The phosphoric acid-based system was chosen as the best representative of the acid systems because it is the fuel cell which best

addresses the problems of tolerance towards carbon monoxide and minimization of noble-metal loadings while maintaining high levels of anodic and cathodic activities. The confidence of the electrical utilities, as displayed by their support of the United Technologies fuel cell program, tends to support this judgement.

13.1.2 Alkaline Fuel Cells

The systems considered in this electrolyte-type group are those based on 30 wt % aqueous potassium hydroxide solutions at approximately 343°K (158°F), 75 wt % aqueous potassium hydroxide solutions at approximately 473°K (392°F), and saturated carbonate-bicarbonate aqueous solutions at temperatures of approximately 333°K (140°F). Although alkaline hydrogen-oxygen (H_2-O_2) fuel cells have been investigated by many companies, principally Union Carbide, Shell (U.K.), Allis Chalmers, United Technologies, Exxon, Alsthom (France), Varta (G.F.R.), and Siemens (G.F.R.), only a small fraction of the overall effort has been devoted to the use of carbonaceous fuel gases. This neglect is principally due to problems associated with carbonation of the electrolyte by carbon dioxide in air or the fuel gas, or from the oxidation of carbon monoxide from the dissolution of the carbon dioxide formed in the alkaline electrolyte.

Various techniques (References 13.6) have been employed to overcome this problem in bench-scale systems. The electrolyte is circulated in the Allis-Chalmers methanol-oxygen system and is regenerated externally. The Shell (U.K.) methanol-air system employs a silver-palladium tube to separate the hydrogen from the gas mixture after external reformation, so that carbon-containing gases do not have access to the electrolyte. A similar scheme has been employed by United Technologies to provide hydrogen, formed by steam reformation of carbonaceous fuels, to the fuel electrodes of Bacon cells. This technique is impractical from technical and economic considerations in fuel cell power plants of the type considered in this study.

Commercial acid gas-scrubbing systems—for example, Lurgi Rectisol and Benfield—are available, which will permit reduction of the

carbon dioxide content of feed streams to less than 10 ppm. There is, however, a serious question as to whether this level is sufficiently low to prevent electrolyte carbonation and, worse still, solid potassium carbonate formation in the pores of the gas diffusion electrodes, thereby leading to relatively rapid performance degradation. The problem of electrolyte carbonation may be dealt with by electrolyte circulation, followed ultimately by replacement with fresh caustic solution, or by cyclic decarbonation of the electrolyte in an external electrolytic cell, as proposed for the Exxon-Alstom methanol-air fuel cell power system (Reference 13.7).

Scrubbing carbon dioxide from the air feed to the alkaline fuel cells may also be accomplished by the Lurgi and Benfield processes. This would probably be as expensive as the fuel gas scrubbing described above, despite the much lower carbon dioxide content of air, as the capital costs of the processes are a strong function of the total number of moles of throughput gas. Another approach involves simple caustic scrubbing of the incoming air. This could cost as little as \$5/kW (Reference 13.8). Rather than dispose of the approximately one ton of potassium carbonate formed each day in a 25 MW fuel cell power plant, it is probably more desirable to regenerate the caustic by the Exxon-Alstom technique alluded to above.

Despite the problems associated with scrubbing carbon dioxide from the fuel gas and air feeds, alkaline fuel cell subsystems are very attractive for large-scale power generation from the standpoints of cost and useful life. For example, long-lived air cathodes in alkaline solution operate with significantly better polarization characteristics than they do in acidic solutions. This is all the more remarkable since these cathodes do not require noble metals. Silver may be employed as the perhydroxide ion elimination catalyst with a concomitant substantial lowering of catalyst cost. A further advantage of alkaline fuel cell power systems is that nickel at approximately \$4.41/kg (\$2/lb) is stable in the cell environment. In contrast, acid systems are restricted to

graphite and other carbon products because of the prohibitive expense of tantalum at \$117/kg (\$53/lb) and niobium at \$33/kg (\$15/lb).

Estimated overall system efficiencies for alkaline fuel cell power systems, using carbonaceous gas as a fuel, are similar to those quoted for acid systems, about 35 to 40%. The efficiency advantage of the alkaline fuel cell, resulting from lower cathodic polarization, is almost totally negated by the efficiency penalty associated with elimination of carbon dioxide from the air and fuel gas streams (Reference 13.9).

An alkaline fuel cell, operating in 30 wt % potassium hydroxide solution at approximately 343°K (158°F) has been selected to represent this class of fuel cells in the rest of this study. It is preferred over the Bacon fuel cell, which operates typically in 75 wt % potassium hydroxide solutions at temperatures in excess of 473°K (392°F), because of the well-known severe corrosion problems of this latter system. In contrast, the Exxon-Alstom carbonate-bicarbonate fuel cell is disqualified from further consideration because of problems relating to low cathodic activity due mainly to excessive concentration polarization (Reference 13.8).

13.1.3 Molten Carbonate Fuel Cells

No large molten carbonate fuel cell batteries have been produced. Systems studies have been made for 15 to 22 kW units by Texas Instruments (Reference 13.10) and IGT (References 13.11 and 13.12); and United Technologies has a development program on which few details have been published (References 13.13 and 13.22). Broers (References 13.14 and 13.15) did much of the earlier research on the devices upon which current technology is based but did little work on multiple-cell devices.

Texas Instruments designed a 1 kW test unit which was delivered to the U. S. Army Mobility Equipment R&D Center at Fort Belvoir, Virginia. IGT has operated fuel cell batteries in excess of 2 kW, but the results of their recent work are not published. These units use a separate reformer to supply a suitable mixture of hydrogen and carbon monoxide to the cells.

An economic assessment of molten carbonate fuel cells for large-scale power production with propane feed to an integrated reformer-molten carbonate fuel cell system was published by Hart and Womack (Reference 13.16) in 1967. The calculations they presented were based in part on the unpublished work at the Central Electricity Generating Board Research Laboratories and Marchwood Engineering Laboratories. They used a fuel cell model representing their projection of the best fuel cell performance to be expected in the near future plus an estimation of the cost of individual parts of the model cell. The fuel cell operating at maximum power was assumed to use its waste heat to generate steam. This steam supplied a steam turbine generator for additional power generation. For this system a plant efficiency of 46.5% was calculated based on propane feed as the fuel and assuming that a propane reformer was integrated with the fuel cell system. They concluded that initial capital costs for the system would be at least 25% higher than for a coal-fired steam turbine generator system of similar life. Since the fuel cell life was expected to be short [optimistically 157.7 Ms (5 yr) compared to perhaps 315.4 to 630.7 Ms (10 to 20 yr) for a gas turbine and 630.7 to 946.1 Ms (20 to 30 yr) for a steam turbine], they concluded that such a fuel cell plant was not economically attractive. The economics of a fuel cell for domestic power are different, and von Fredersdorff (IGT) published an analysis of the molten carbonate fuel cell for this use in 1963 (Reference 13.17), concluding that the system could be economical if the fuel cell investment cost was no more than \$300 over a 315.4 Ms (10 yr) operation.

All costs—fuel costs in particular—are much higher now than they were at the time of the earlier studies. The molten carbonate fuel cell power capability has been improved significantly; and there are stricter regulations on emissions from power plants, substantially increasing capital cost and reducing efficiency.

The practical efficiencies which may be attained for large fuel cell batteries of this type, based on small battery performance, are still

in dispute, but 45% efficiency based on the higher heating value (HHV) of natural gas is probably feasible (Reference 13.18).

The cells with highest performance reported recently use a paste electrolyte of alkali aluminates suggested by Broers (Reference 13.15), a nickel anode and lithiated nickel oxide cathode. The cells give useful power densities at 923°K (1202°F) of up to about 161 mW/cm² (150 W/ft²) using air as oxidant, depending on the fuel gas. Oxygen allows greater power densities. Cell temperatures of 873 to 1023°K (1112 to 1382°F) have been used, and corrosion problem increases at the higher temperatures.

In large systems it is anticipated that increased efficiency can be attained by recovering waste heat and using it to generate steam for use in a steam turbine or as process steam, as suggested by Hart and Womack (Reference 13.16). The highest efficiency plants would be large ones to minimize the turbogenerator plant costs.

If methanated (high-Btu) gas or methanol is used as a fuel, a reforming step is necessary. A higher efficiency is attainable if this can be accomplished on the fuel cell electrodes or nearby surfaces within the cell, since the reforming process absorbs heat which would be supplied in situ by the heat produced at the electrodes of the molten carbonate cell. Although a commercial reformer operates at a higher temperature than does the molten carbonate cell, the large electrode area within the cell and the probable slower throughput per unit area makes internal reforming a reasonable possibility. Experiments on internal reforming or reforming on a catalyst at the same temperature as the fuel cell have been successful (References 13.19 and 13.20), but the feasibility of internal reforming at high power densities in multicell systems is still to be demonstrated. The fuel cell plant would have a clean exhaust, since most of the sulfur is removed from the fuel gases initially, and oxides of nitrogen are not formed in significant quantities at these low temperatures. Unused fuel gases can be burned with excess oxygen in the cathode exit gas stream.

In addition to the fuel cell plant cost analysis reported by Hart and Womack for large-scale power production by a molten carbonate system, Bockris and Srinivasan (Reference 13.21) present a list, from different sources, of recent (1969) and future fuel cell costs. A projected cost for the molten carbonate cell system attributed to Broers is \$600/kW for large-scale production using natural gas and a reformer. More recently, United Technologies reports (References 13.13 and 13.22) an estimate of ~ \$225/kW capital cost (EPRI RP114 Program) for the molten carbonate system with a lower heating value (LHV) efficiency of 47%. They also report multicell life tests, in conjunction with the Electric Power Research Institute, of 36 Ms (10,000 hr) between overhaul, are good performance stability during subscale tests, some of which lasted more than 18 Ms (5000 hr).

Broers and IGT have also reported up to 36 Ms (10,000 hr) fuel cell life, but there is no consensus on the eventual maximum which may be possible.

Variations of the Broers type of molten carbonate fuel cell have been suggested, such as the use of a liquid lead electrode catalyst for oxidation of solid forms of carbon or coal (Reference 13.23), but none has received an extended study effort or been developed as yet.

In summary, on the basis of available information, the eventual useful fuel cell life, cost, type of reforming necessary, and efficiency are still very questionable.

13.1.4 Stabilized Zirconia Fuel Cells

High-temperature solid electrolyte fuel cells have many advantages over other types of fuel cell. As Markin describes (Reference 13.24), these are:

- There are no liquids involved, so problems associated with pore flooding and maintenance of a stable three-phase interface are totally avoided

- The electrolyte composition is invariant and does not depend on the composition of the fuel and oxidant streams
- Activation polarization losses are negligible.

Further, a power system, based on the use of stabilized zirconia at operating temperatures of approximately 1273°K (1832°F), has a unique advantage when coal is employed as a fuel in that the waste heat, generated because of the thermodynamic and electrochemical inefficiencies of the fuel cells, may be used directly in the gasification of coal, thus providing fuel gas for the fuel cells. This thermal coupling leads to high overall efficiencies for the power system, and practical efficiencies of greater than 60% are considered possible (Reference 13.25). This efficiency is all the more remarkable because this type of power system employs coal, rather than natural gas, methanol, or naphtha, as a fuel.

The disadvantages of this fuel cell are related to:

- The relatively high electrolyte resistivity
- The need for an effective, low-cost method of interconnection of cells to form a battery
- Problems involving battery component interactions and adequate sealing techniques.

Many companies and other research organizations [e.g., Westinghouse, General Electric, C. G. E. (France), Brown Boveri (G. F. R.), Battelle (Geneva, Switzerland), and AERE (Harwell, England)] have explored possible solutions to these problems in their efforts at component development and device fabrication. The largest device demonstrated to date was based on a bell and spigot geometry. It was constructed and operated by the Westinghouse Electric Corporation and delivered 100 W of electrical power (References 13.25 and 13.26).

The Westinghouse thin-film concept (Reference 13.27) provides an economical and effective method for the series-connection of individual cells in a solid-electrolyte battery. The largest of these devices

demonstrated to date delivered 8 W and lived for approximately 360 ks (100 hr)—the first 108 ks (30 hr) saw a 5% voltage degradation at a given current output (Reference 13.28). The life-limiting problem lay in the interconnection-electrolyte junction where inadequate sealing led to fuel gas leakage, resulting in air electrode reduction and, ultimately, performance degradation. This design has been selected for this study.

13.2 Description of Parametric Points

To facilitate a comprehensive comparison of the four fuel cell power-generation systems, the parameter values listed in Table 13.1 below were fixed for the base cases of all systems.

Table 13.1 - Base Case Values Common to All Systems

Parameter	Value
Size of Bower Plant	25 MW dc
Type of Fuel	High-Btu gas
Type of Oxidizer	Air
Fuel Cell Useful Life	10,000 hr (5% efficiency degradation)

The smallest system to be to be considered in this study, 25 MW, was chosen for the base cases, as no economies of scale are expected for the fuel cell subsystem. Further, no fuel cell power system of even two orders of magnitude lower in rating has been constructed or operated, so that even in this small a power plant the system problems can only be addressed in the most general fashion, as will be evident from the schematics provided in the following subsections.

A 25 MW fuel cell power plant is too small to justify the expense of a dedicated coal gasification reactor. Thus, it will be fueled with hydrogen, high-Btu gas, or methanol, all of which may be derived from

coal. Because of the flexibility afforded by the existing extensive network of natural gas lines, high-Btu gas was selected as the fuel for the base cases.

A value for the useful lives of all of the fuel cell subsystems was arbitrarily fixed for the base cases at 36 Ms (10,000 hr) at a constant power output, with a 5% degradation in terminal voltage. The degradation specification is important in the determination of the power conditioning costs. The useful life value specified is considered a reasonable estimate of the state of the art for molten carbonate fuel cells, an overestimate by a factor of approximately one hundred for the solid electrolyte fuel cell, and an underestimate by a factor of approximately two for aqueous acid and alkaline fuel cells, as described in the previous section.

In general, the parametric assessment involved the variation of one parameter, with the others retaining the value used in the base case. In certain cases, however, a change in one parameter caused the variation of other parametric values. The change from air to oxygen as the oxidant, for example, was thought to cause an increase in the current density in the acid, alkaline, and molten carbonate fuel cells; and an increase in the cell voltage in the solid electrolyte fuel cells.

Point 1 of the parametric point list for each fuel cell type is the base case. The abbreviations AC, AL, MC, and SE were chosen for the acid, alkaline, molten carbonate, and solid electrolyte fuel cell power systems, respectively. Thus, AC1 represents the base case in the phosphoric acid system. The first eight parametric value changes common to all four systems are shown in Table 13.2. Points 2 through 4 explored the economics of scale realized by increasing the power plant rating. Because of their modular nature, no economy of scale was assumed for the fuel cell subsystems. Any economic benefits realized, therefore, come from the other subsystems. The effect of the replacement of air with oxygen from a dedicated oxygen plant was tested for all systems in Point 5. Three more parametric points per system (6 to 8)

Table 13.2 - Parametric Changes Common to all Four Systems

Point No.	Parameter	Value	Base Case
2	Size of power plant	100 MW	25 MW
3	Size of power plant	250 MW	25 MW
4	Size of power plant	900 MW	25 MW
	Fuel type	Medium-Btu gas	High-Btu gas
5	Oxidant type	Oxygen	Air
6	Fuel cell useful life	30,000 hr (5% Voltage degradation)	10,000 hr
7	Fuel cell useful life	50,000 hr (5%)	10,000 hr
8	Fuel cell useful life	100,000 hr (5%)	10,000 hr
9	Fuel cell useful life	100,000 hr (15%)	10,000 hr

were expended to investigate the impact of fuel cell subsystem useful life on the electricity costs. The effect of a terminal voltage decrease of 15% (at constant power output) after 360 Ms (100,000 hr) of plant operation was explored in Point 9.

A cross-comparison of the efficiencies and electricity costs is possible for all systems in Points 4 through 9, and for the acid, alkaline, and molten carbonate systems in Points 2 and 3. The solid electrolyte power system is not available for comparison purposes in Points 2 and 3 because medium-Btu gas was employed as a fuel instead of the high-Btu gas used in the other three systems. Further details of Points 2 through 8, and a description of the parametric points specific to each system, are provided in the following subsections.

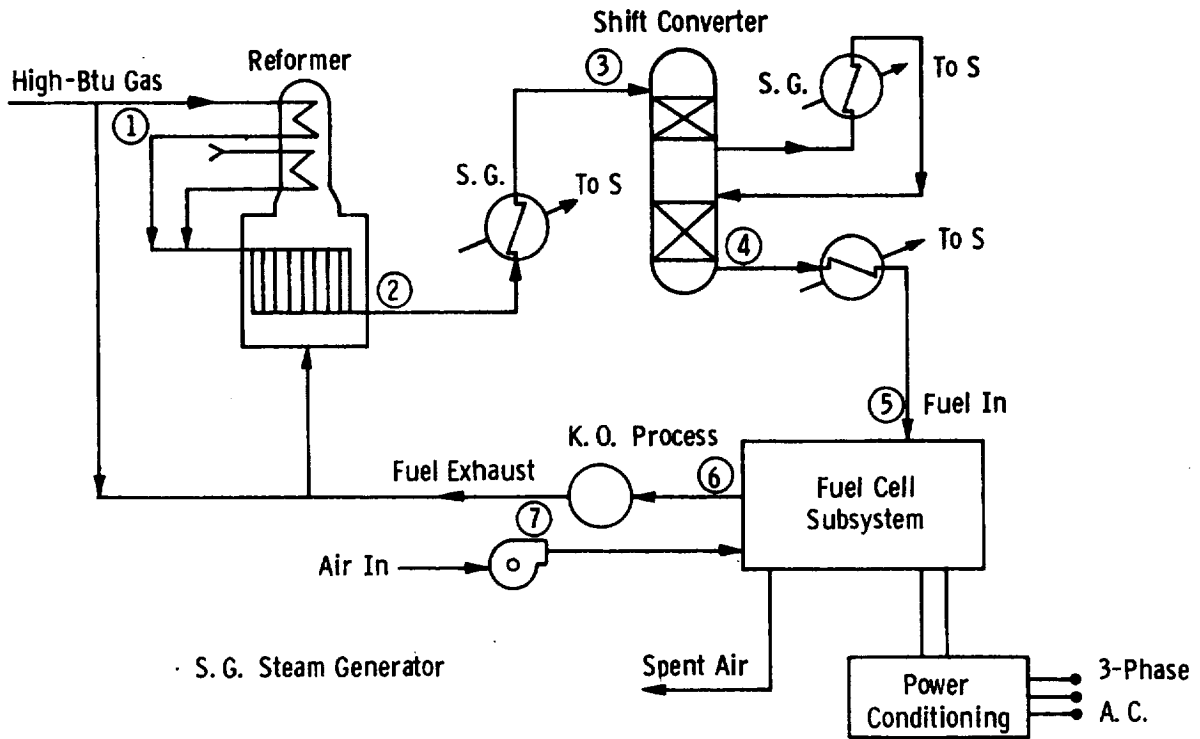
13.2.1 Phosphoric Acid Fuel Cell Power System

As indicated in the previous subsection, the base case in this power system, AC1, involves a fuel cell subsystem that has a 25 MW dc rating, a useful life of 36 Ms (10,000 hr) of operation, and a 5%

voltage degradation at constant power output. The fuel is a high-Btu gas; air is the oxidant. In the fuel cell subsystem, the fuel cell is considered to have a bipolar design similar to that described by Baker et al. (Reference 13.29). The anode and cathode, each fabricated from carbon as described by Kordesch and Scarr (Reference 13.30), are catalyzed by a platinum addition to each electrode of 1 mg/cm^2 (0.002 lb/ft^2). The addition of other electrocatalysts for the reduction of performance sensitivity to carbon monoxide in the fuel stream (Reference 13.31) are considered desirable but were not included in the economic analysis. The electrolyte is 85 wt % aqueous phosphoric acid immobilized in a zirconium pyrophosphate matrix with an effective resistivity of $2 \Omega\text{-cm}$. The electrolyte thickness is 0.5 mm (0.020 in). The fuel cell subsystem operates at 463°K (375°F). Based on a conservative estimate of anticipated advances in the state of the art, beyond that reported by Schiller and Meyer (Reference 13.32) in 1971, values of 0.7 V and 200 mA/cm^2 (186 A/ft^2) were selected for the cell voltage and current density, respectively.

A schematic of the complete power system is shown in Figure 13.1. High-Btu gas from a 0.689 MPa (100 psi) abs line is assumed to be available, and is fed after preheating to a steam-methane reformer operating at a pressure of 0.689 MPa (100 psi) abs and a temperature of 1144°K (1600°F). The reformer effluent, consisting mainly of carbon monoxide, hydrogen, carbon dioxide and steam, is cooled and fed to a shift converter operating at 0.483 MPa (70 psi) abs and 700°K (800°F). The shift converter is operated at as low a temperature as possible in order to minimize the carbon monoxide concentration in the exit gas. The hydrogen-rich fuel gas is further cooled to approximately 422°K (375°F) and is fed to the ten fuel cell modules. Air is supplied to the modules by means of blowers, as shown in Figure 13.1.

Steam, required for the steam reformation of methane (the principal constituent of high-Btu gas) is raised in the cooling of the fuel gas between the reformer and the shift converter, in the shift converter, and between the shift converter and fuel cell subsystem. The water required for the steam generators is reclaimed from the fuel-gas



Point No	Temperature, °F	Pressure, psia	Point No	Temperature, °F	Pressure, psia
1	60	100	5	300	25
2	1600	85	6	375	20
3	800	70	7	120	18
4	800	40			

Fig. 13.1 - Schematic of a phosphoric acid fuel cell power system which employs high-Btu gas as fuel

exhaust by the knock-out process shown in Figure 13.1. The water-vapor depleted exhaust gases, containing approximately 10% of the hydrogen fed to the fuel cell modules, and the unused carbon monoxide, is mixed with a portion of the incoming high-Btu gas, and the mixture is burned to supply the heat required to cover the endothermic processes in the reformer. Further details of the fuel processing subsystem are provided in Appendix A 13.1.

The fuel cell subsystem, rated at 25 MW dc, consists of ten 2.5 MW modules. Each module is wired into a dedicated power conditioning unit consisting of a force-commutated dc to ac inverter, a transformer, and filters. A more detailed description of the power conditioning subsystem is provided in Appendix A 13.2.

As indicated in the previous subsection, the changes from the base case, characterized by Points 2 through 9 and shown in Table 13.2, are common to all systems. AC2 and AC3 differ from AC1 only in that the fuel cell subsystems are rated at 100 MW and 250 MW dc, respectively. In AC4, the fuel cell subsystem is rated at 900 MW dc. The fuel used in the 900 MW plant is medium-Btu gas, instead of the high-Btu gas employed in AC1, AC2, and AC3. Thus, the steam-methane reformer, shown in Figure 13.1, is unnecessary in this power plant.

AC5 involves the use of oxygen as the fuel cell oxidant instead of air. Because of the reduction of concentration polarization at the fuel cell cathode, the apparent current density was considered to have doubled, i.e., from 200 mA/cm^2 (186 A/ft^2) to 400 mA/cm^2 (372 A/ft^2), despite the concomitant increase in the fuel electrode polarization and the cell ohmic polarization losses.

In parametric points AC6, AC7, and AC8, the fuel cell subsystem useful life in a 25 MW dc plant is increased to 108 Ms (30,000 hr), 180 Ms (50,000 hr), and 360 Ms (100,000 hr), respectively. In all three points, as in the base case, AC1, a 5% efficiency or voltage degradation, by comparison with the initial performance, is assumed at the end of useful life. Thus, the initial cell voltage, 0.7 V, will have fallen to 0.665 V at end of life. In AC9, a voltage degradation of 15% (0.105 V) is assumed at the end of a 360 Ms (100,000 hr) useful life.

The changes from the base case, represented in Points AC10 through AC16, are specific to the phosphoric acid fuel cell power system. These are shown in Table 13.3.

Table 13.3 - Parametric Points - Acid Fuel Cell

Point No.	Parameter	Value/Type	Base Case Value
10	Current density	300 mA/cm ²	200 mA/cm ²
11	Current density	400 mA/cm ²	200 mA/cm ²
12	Catalyst loading	0.3 mg Pt/cm ²	1 mg Pt/cm ²
13	Catalyst loading	0.1 mg Pt/cm ²	1 mg Pt/cm ²
14	Type of fuel	Methanol	High-Btu gas
15	Electrolyte thickness	0.25 mm	0.5 mm
16	Oxidant type	Oxygen	Air
	Size of power plant	250 MW	25 MW

AC10 and 11 were included to explore the effect of fuel cell subsystems cost reductions, resulting from increases in the apparent current density of the base case, 200 mA/cm² (186 A/ft²), to 300 mA/cm² (279 A/ft²) and 400 mA/cm² (372 A/ft²). Similarly, AC12 and 13 represent advances in the state of the art which result in the lowering of platinum loadings in both cathode and anode from 1 mg/cm² (0.002 lb/ft²) in the base case to 0.3 mg/cm² (6 x 10⁻⁴ lb/ft²) and 0.1 mg/cm² (2 x 10⁻⁴ lb/ft²) respectively.

High-Btu gas is replaced by methanol as the fuel in AC14. This will involve storage of methanol in tanks at the power plant. The electrolyte thickness of 0.5 mm (0.020 in) in the base case is reduced to 0.25 mm (0.010 in) in AC15. Because of the resultant reduction in ohmic losses in the electrolyte, the cell voltage at 200 mA/cm² (180 A/ft²) is considered to have increased from 0.70 V to 0.71 V.

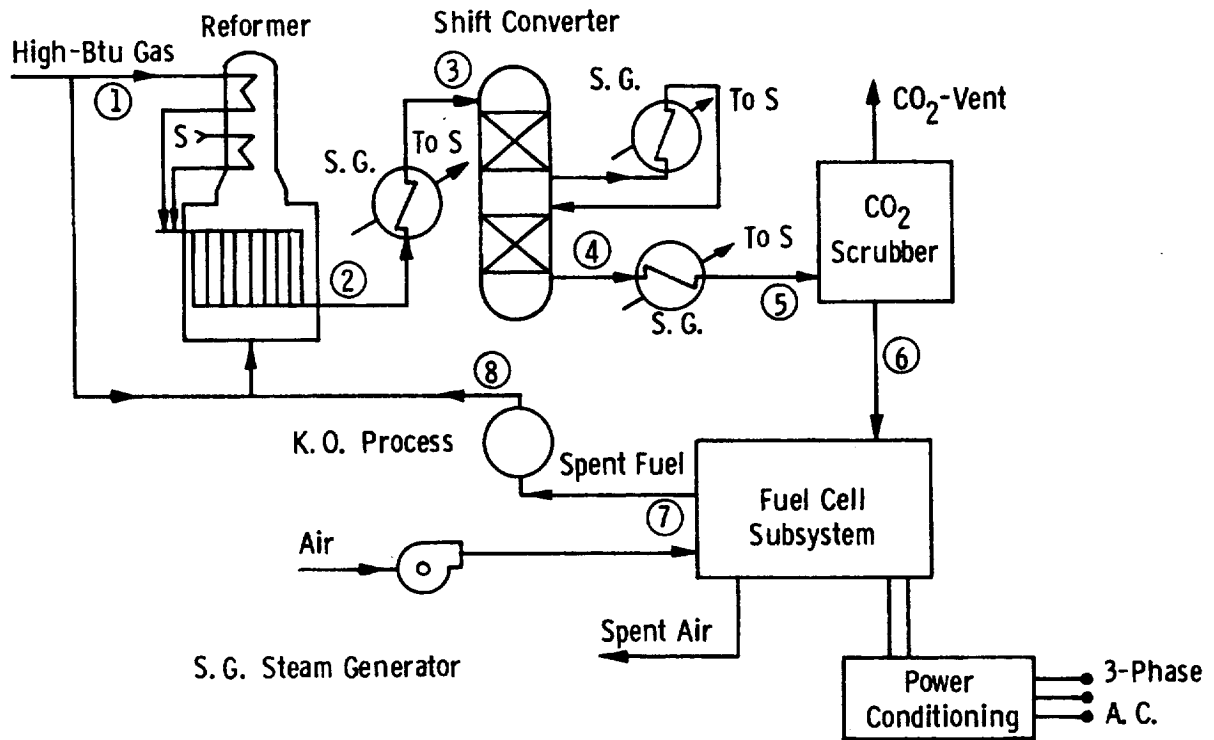
The effects of replacement of air by oxygen in a 250 MW dc power plant are explored in AC16. Once again, as in AC5, the assumed apparent current density of the cell was doubled to 400 mA/cm^2 (372 A/ft^2). This point was included to demonstrate, by comparison with AC5, the effect of scale in lowering the oxygen cost.

13.2.2 Alkaline Fuel Cell Power System

A schematic of the alkaline fuel cell power system base case, All, is shown in Figure 13.2. The temperatures and pressures at various system locations are also presented in this figure. For the purposes of this study, it has been assumed that the alkaline power system is similar to its acid counterpart, described in the previous subsection, but with the following differences:

- Carbon dioxide scrubbers - To prevent carbonation of the electrolytes the Lurgi Rectisol process is employed to scrub both the air and the fuel gas inlet streams, as shown in Figure 13.2. Further details of this process, which employs refrigerated methanol to scrub acid gases, are found in Appendix A 13.1.

- Fuel cell subsystem - A bipolar configuration, identical with that described for the acid system, is assumed also for the alkaline fuel cell battery. However, the apparent current density in the base case for this system, 100 mA/cm^2 (93 A/ft^2), is only half that in the acid fuel cell. The cell voltage, 0.8 V, is higher because of the generally lower cathodic polarizations in the alkaline fuel cell. These values represent a conservative estimate of advances in the state of the art since 1969 (Reference 13.33) for this fuel cell. The anode and cathode are catalyzed with 1 mg of platinum/ cm^2 (0.002 lb/ft^2) and 5 mg of silver/ cm^2 (0.1 lb/ft^2), respectively. The electrolyte thickness is 0.5 mm (0.020 in) with an assumed effective resistivity of $2 \Omega\text{-cm}$ (5 ohm-in). The fuel cell operates at a temperature of 343°K (158°F), not only for the base case but also for all parametric points described below.



Point No	Temperature, °F	Pressure, psia	Point No	Temperature, °F	Pressure, psia
1	60	100	5	400	25
2	1600	85	6	100	25
3	800	70	7	160	20
4	800	40	8	100	18

Fig. 13.2— Schematic of an alkaline fuel cell power system which employs high-Btu gas as fuel

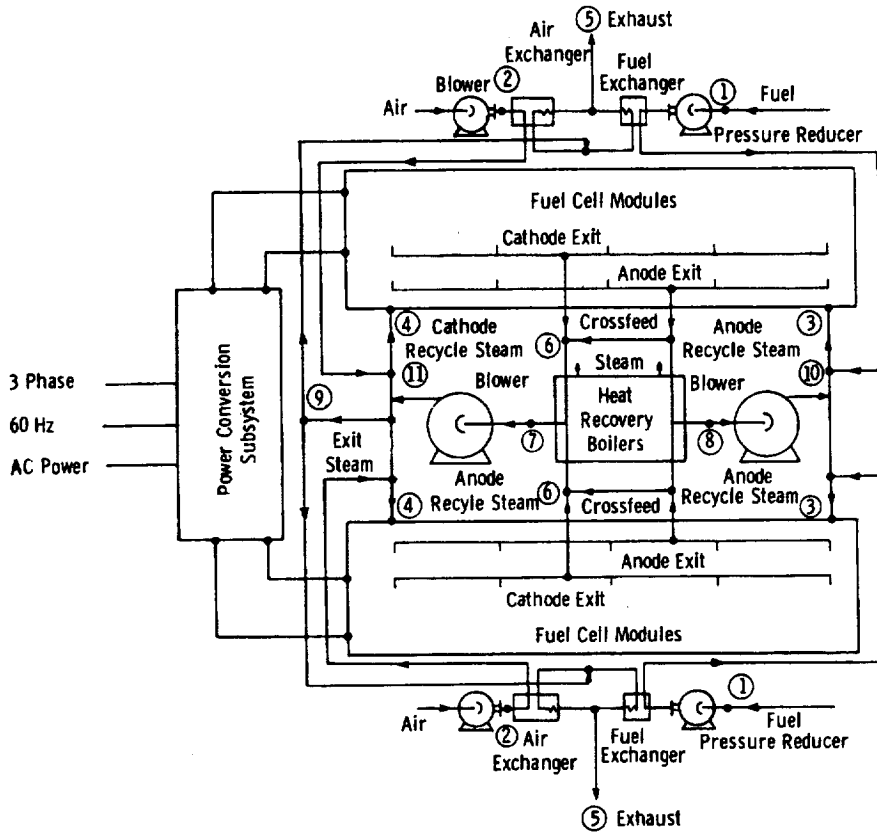
Points AL2 through AL9 involve the same changes in parameter values as those for AC1 through AC9 discussed in the previous subsection. Once again, replacement of air as oxidant by oxygen from a dedicated oxygen plant (AL5) is considered to result in a doubling of the base case apparent current density, i.e., from 100 mA/cm² (93 A/ft²) to 200 mA/cm² (186 A/ft²).

The changes from the base case, represented by parametric points AL10 through AL16, are specific to the alkaline fuel cell system, and are shown in Table 13.4 below.

Table 13.4 - Parametric Points - Alkaline Fuel Cell

Point No.	Parameter	Value/Type	Base Case Value
AL10	Current density	175 mA/cm ²	100 mA/cm ²
AL11	Current density	250 mA/cm ²	100 mA/cm ²
AL12	Anode catalyst loading	0.1 mg Pt/cm ²	1 mg/cm ²
AL13	Anode catalyst loading	0.01 mg Pt/cm ²	1 mg/cm ²
AL14	Anode and cathode materials	Raney Nickel	Pt/C
AL15	Cathode catalyst loading	1 mg Ag/cm ²	5 mg/cm ²
AL16	Electrolyte thickness	0.25 mm (0.010 in)	0.5 mm (0.020 in)

The effects of increases in the fuel cell apparent current density from 100 mA/cm² (93 A/ft²) to 175 mA/cm² (163 A/ft²), brought about by improvements in the state of the art, are tested in parametric points AL10 and AL11, respectively. Points AL12 and AL13 explore the effects of reductions of platinum loadings in the anode from 1 mg/cm² (2.10⁻³ lb/ft²). Replacement of the carbon gas-diffusion electrodes by Raney nickel electrodes is examined in parametric Point AL14. Point AL15



Point No.	Temperature, °F	Pressure, psia	Point No.	Temperature, °F	Pressure, psia
1	ambient	100	7	752	15
2	ambient	35	8	752	15
3	752	30	9	752	30
4	752	30	10	752	30
5	302	15	11	752	30
6	1202	25			

Fig. 13.3—Schematic of molten carbonate fuel cell power systems (MC 1-3, 6-14 and 16)

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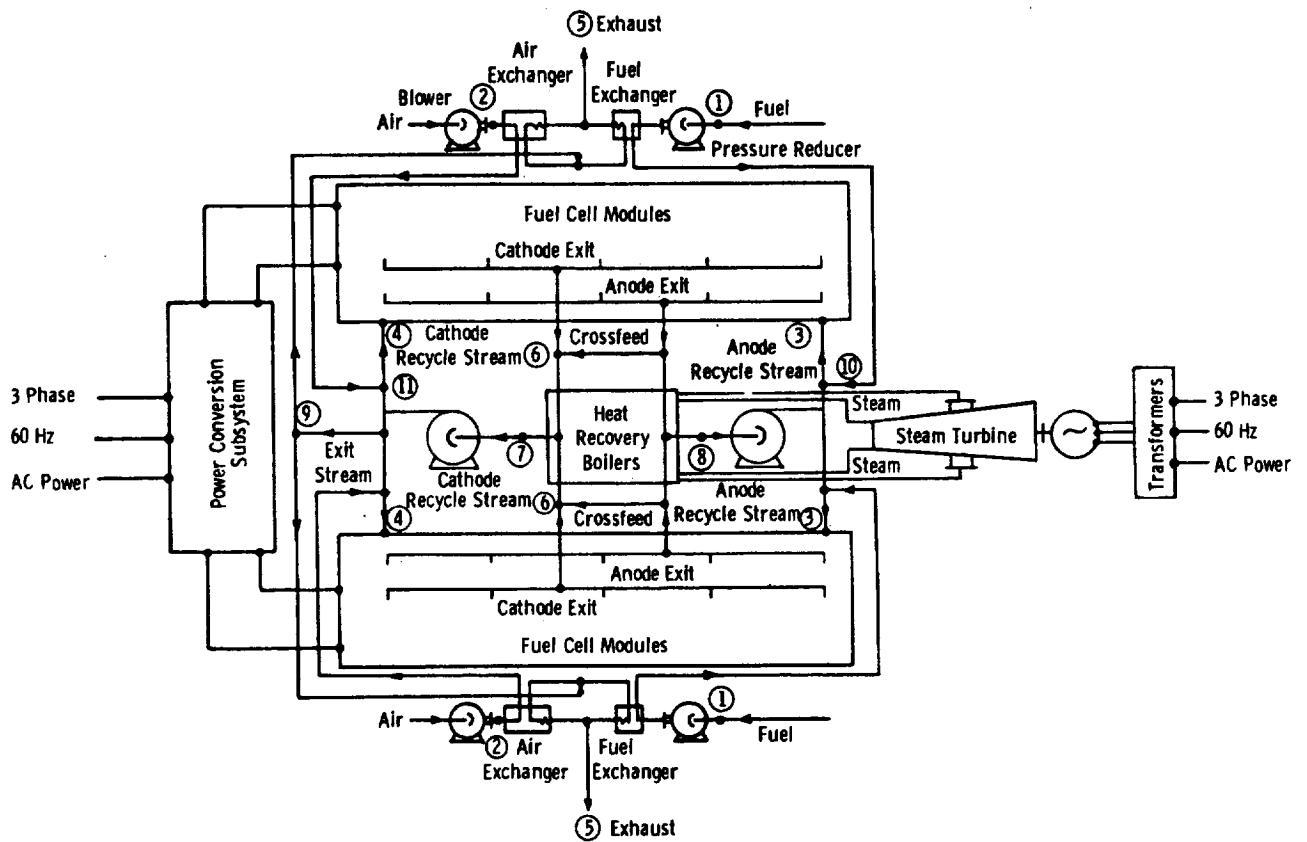
tests the effect of the reduction of the silver loading in the cathode from 5 mg/cm^2 (0.01 lb/ft^2) to 1 mg/cm^2 (0.002 lb/ft^2). In Point AL16, the electrolyte thickness, or electrode separation, is 0.025 cm (0.01 in), i.e., half of that in the base case. This results in a cell voltage increase of 5 mV , so that the cell voltage at 100 mA/cm^2 (93 A/ft^2) is 0.805 V in AL16.

13.2.3 Molten Carbonate Fuel Cells

The effect of ten plant and operating variables on the cost of electricity was investigated. These are the effect of power plant size, fuel cell life, cell output degradation, current density at fuel cell electrodes, electrolyte thickness, temperature of the fuel cell, replacement of air by oxygen, oxygen plant size, fuel type, and recovery of waste heat from the fuel cell by a steam turbine generator system.

MC1, the reference or base case, has a 25 MW dc rating with a filter press design similar to that described by IGT (Reference 13.22), with a porous nickel anode and lithiated nickel oxide cathode. The electrolyte is the Broers type (Reference 13.15), consisting of a semisolid paste of alkali aluminate powder and molten alkali carbonates with the ternary eutectic composition (Li_2CO_3 -43.5 mole %, Na_2CO_3 -31.5 mole %, K_2CO_3 -25.0 mole %, m.p. 670°K (747°F)). Electrolyte resistance (Reference 13.23) is assumed to be 1.5 times that of the free electrolyte of the same thickness. This value would actually vary with the paste structure and chemical composition. High-Btu gas is used as fuel, and it is assumed that it can be reformed on internal cell surfaces. A schematic of the plant configuration for MC1 is shown in Figure 13.3. It consists of a split series of fuel cell modules consisting of ten separate banks, each with its own dc-to-ac inverter, transformer, and filters. It operates at 923°K (1202°F), as do all other plants except MC12 and MC13, which operate at 973 and 1023°K (1292 and 1382°F), respectively.

Fuel and oxidant streams enter the fuel cell at (3) and (4), (Figure 13.3) and react electrochemically at the anode and cathode, respectively, at the fuel cell temperature of 923°K (1202°F). After



Point No.	Temperature, °F	Pressure, psia	Point No.	Temperature, °F	Pressure, psia
1	ambient	100	7	752	15
2	ambient	35	8	752	15
3	752	30	9	752	30
4	752	30	10	752	30
5	302	15	11	752	30
6	1202	25			

Fig. 13.4—Schematic of a molten carbonate fuel cell power plant with heat recovery by a steam turbine generator (MC 4)

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leaving the fuel cell, part of the anode (fuel) gas stream is added to the cathode gas stream at (6) to replace carbon dioxide lost in the cathode electrochemical reactions. The two gas streams, thus modified, enter a heat recovery steam generator and exit at a temperature between 673 and 823°K (752 and 1022°F), depending on the recycle flow rate necessary to minimize concentration polarization at the fuel cell electrodes and to remove excess heat from the fuel cell.

Part of the cathode gas stream is then diverted at (9) through a counterflow heat exchanger, leaving as stack exhaust at (5) at 423°K (302°F), and in the process preheating incoming air and desulfurized fuel from ambient temperature to 573 to 723°K. The exhaust gas stream contains carbon dioxide, water vapor, nitrogen, a small amount of oxygen, but no combustible gases.

Preheated air and desulfurized fuel, entering at (2) and (1) are then combined with the cathode and anode gas streams at (11) and (10) and enter the fuel cell again at (4) and (3). This completes the cycle.

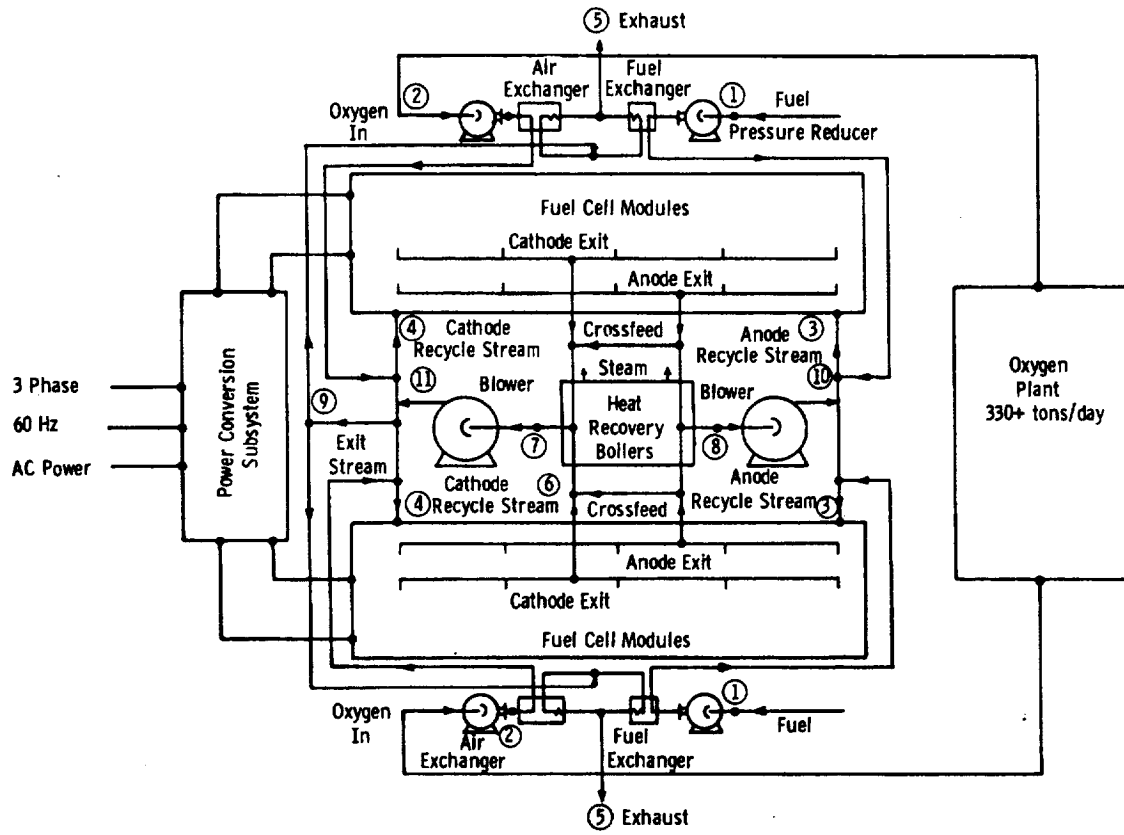
The steam produced can be sold as process steam, used to operate a steam turbine to produce more electricity as in MC4, or operate a turbine drive for an oxygen plant as in Points MC5 and MC17.

MC1, MC2, and MC3 differ only in plant size, being 25 MW dc, 100 MW dc, and 250 MW dc, respectively.

MC4 differs in size (900 MW dc) and also in utilizing a steam turbine generator to produce additional electricity from waste heat, as shown in Figure 13.4. The plant configuration differs from MC1 not only in size but also in allowing space for the steam turbine generator system with its ac transformer.

MC5 is similar to MC1 except for inclusion of an oxygen plant capable of supplying about 3.461 to 3.623 kg/s (330 to 345 tons/day) of oxygen to be used in place of air. The schematic is shown in Figure 13.5.

MC6, 7, and 8 differ from MC1 only in fuel cell lifetimes of 108, 180, and 360 Ms (30,000, 50,000, and 100,000 hr), respectively; and



Point No.	Temperature, °F	Pressure, psia	Point No.	Temperature, °F	Pressure, psia
1	ambient	100	7	752	15
2	ambient	35	8	752	15
3	752	30	9	752	30
4	752	30	10	752	30
5	302	15	11	752	30
6	1202	25			

Fig. 13.5—Schematic of a molten carbonate fuel cell system using oxygen (MC 5 and 17)

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MC9 is similar to MC8 except for the assumption of 15 rather than 5% voltage degradation.

MC10 and 11 assume a current density of 150 and 250 mA/cm² (139 A/ft² and 232 A/ft²) as compared to 200 mA/cm² (186 A/ft²) for MC1.

MC12 and 13 assume an operating temperature of 973 and 1023°K (1292 and 1382°F), respectively, as compared to 923°K (1202°F) for MC1. Electrolyte resistance (Reference 13.23) is a little lower at the higher temperatures, but corrosion is worse.

MC14 uses medium-Btu gas as fuel rather than the high-Btu gas used by MC1.

MC15 uses methanol as fuel, which requires a storage tank as shown in Figure 13.6. Otherwise, it is similar to MC1.

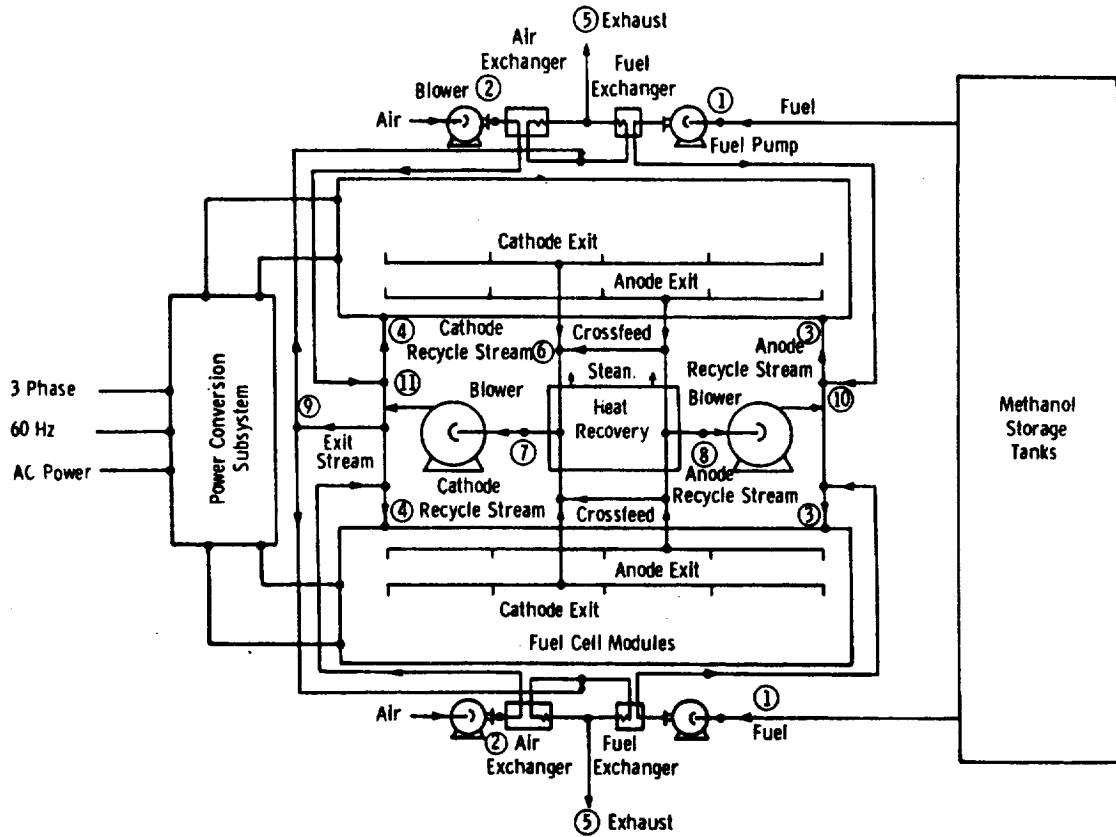
MC16 differs from MC1 in using an electrolyte thickness of 0.5 mm (0.020 in) rather than 1 mm (0.040 in) used for MC1.

MC17 is similar to MC3 (250 MW dc) but uses oxygen rather than air as oxidant. This is included as a comparison to MC5 (25 MW dc) to show the effect of the lower oxygen cost with larger plant size.

13.2.4 Solid Electrolyte Fuel Cell Power System

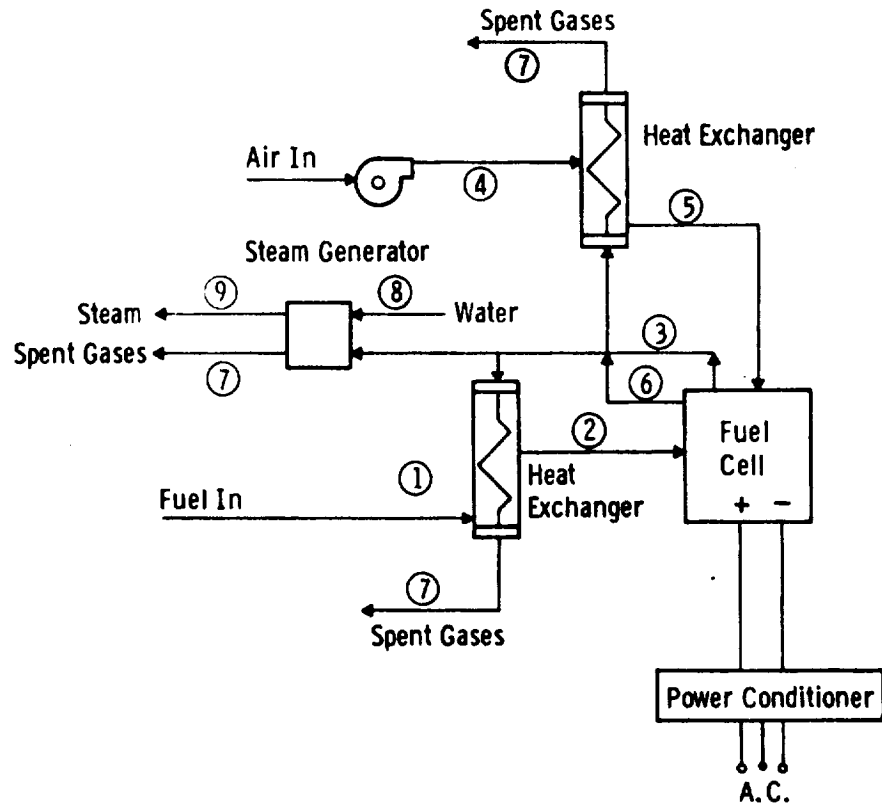
A schematic of the power plant corresponding to the base case, SE1, is shown in Figure 13.7. The basic plant layout is relatively simple in that it consists of the fuel cell generator, the power conditioning subsystem, recuperative heat exchangers (to allow the fuel cell exit gases to heat up the incoming fuel and air streams), and an air blower. The temperatures and pressures at various locations in the power system are also tabulated in Figure 13.7.

All parametric points, including SE1, employ the Westinghouse thin-film solid-electrolyte fuel cell battery in the fuel cell subsystem. This device, built on porous tubes of stabilized zirconia, is described in greater detail in Reference 13.25. The electrolyte film of yttria-stabilized zirconia (ZrO₂-10% Y₂O₃) is gas-impervious and 40 μm (1.6 mils)



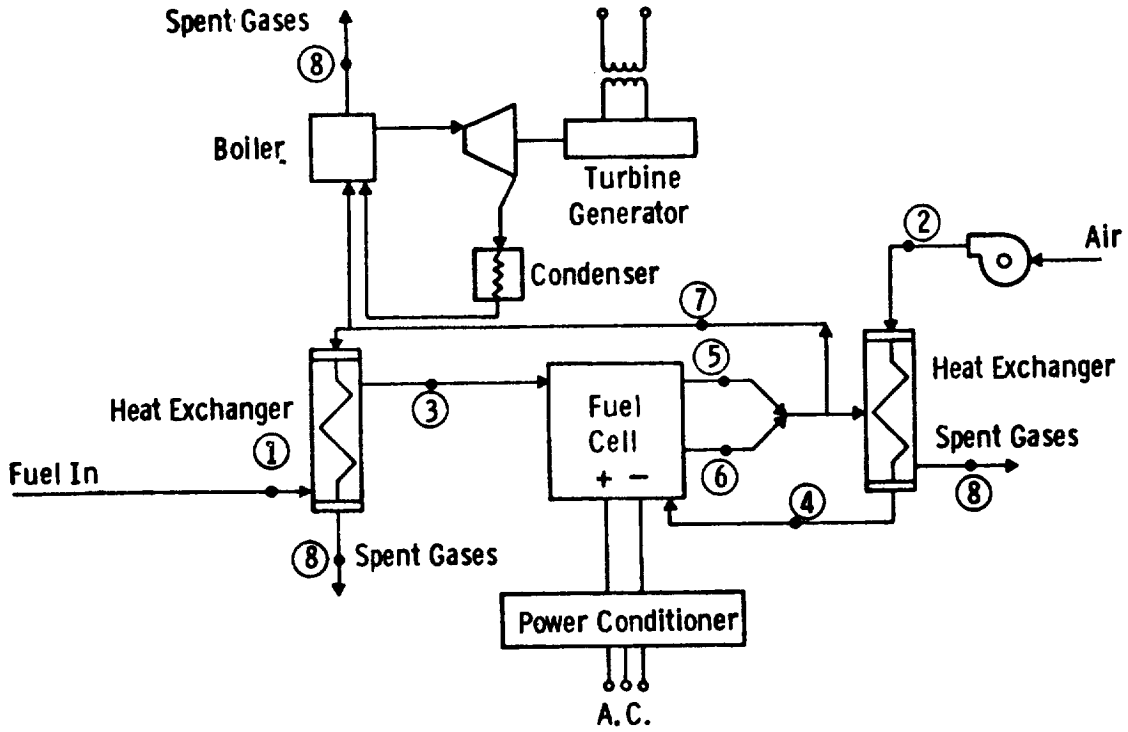
Point No.	Temperature, °F	Pressure, psia	Point No.	Temperature, °F	Pressure, psia
1	ambient	15	7	752	15
2	ambient	35	8	752	15
3	752	30	9	752	30
4	752	30	10	752	30
5	302	15	11	752	30
6	1202	25			

Fig. 13.6—Schematic of a molten carbonate fuel cell system using methanol as fuel (MC 15)



Point No.	Temperature (°F)	Pressure (psia)	Point No.	Temperature (°F)	Pressure (psia)
1	80	60	6	1300	30
2	1100	45	7	200	15
3	1300	30	8	70	650
4	80	60	9	1000	600
5	1100	45			

Fig. 13.7—Schematic of a solid-electrolyte fuel cell power plant employing high-or medium - Btu gas as a fuel. (SE 1-3, SE 6-17, SE 20)



Point No.	Temperature (°F)	Pressure (psia)	Point No.	Temperature (°F)	Pressure (psia)
1	80	60	5	1300	30
2	80	60	6	1300	30
3	1100	45	7	1300	30
4	1100	45	8	200	15

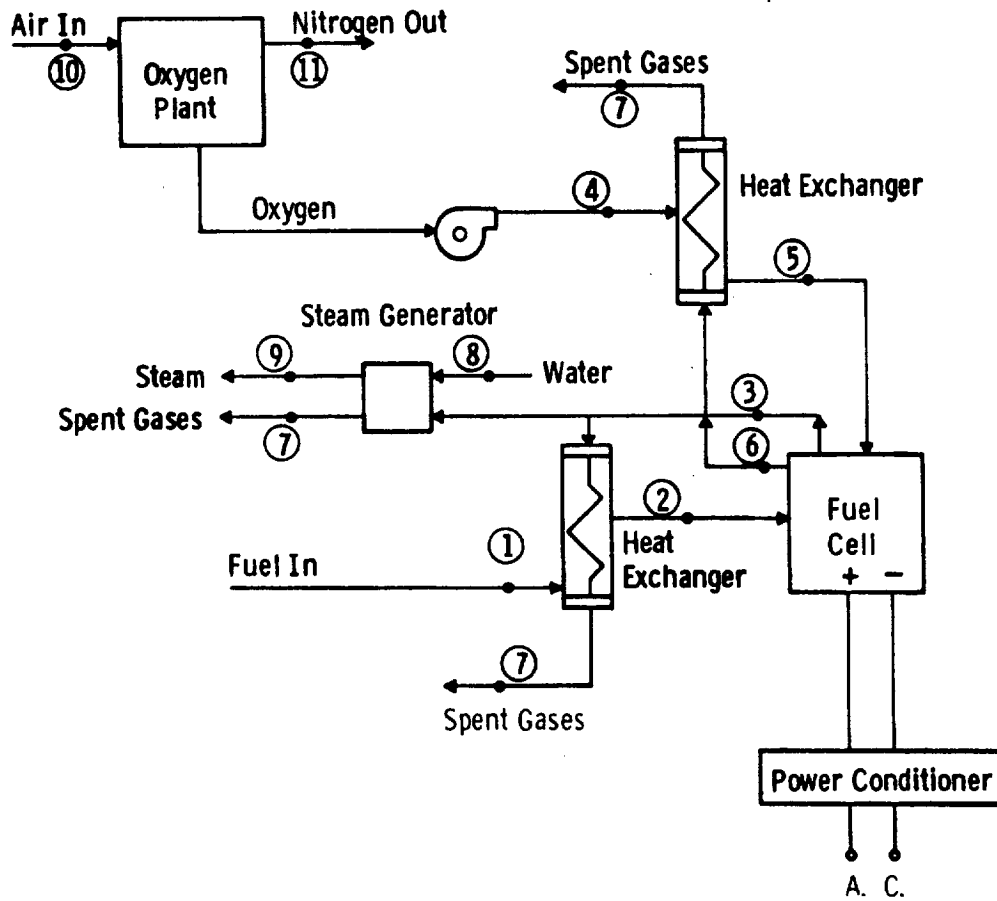
Fig. 13.8—Schematic of a solid-electrolyte fuel cell power plant which employs a steam bottoming plant for waste heat recovery. (SE4)

in thickness. The air electrode is a porous layer of tin oxide-doped indium sesquioxide activated with praesodymium cobaltite. A porous layer of a nickel-stabilized zirconia cermet serves as the fuel electrode. The interconnection layer, which serves to series-connect adjacent cells, consists of a gas-impervious layer of chromium sesquioxide, 20 μm (0.8 mils) in thickness. The device is assumed to operate at 1273°K (1832°F) and with high-Btu gas as fuel. The current density in the electrolyte region of the unit cell is taken as 400 mA/cm² (372 A/ft²).

Parametric Points SE2, SE3, and SE4, explore the economies of scale achievable in power plants. The fuel cell subsystems are rated at 100 MW dc, 250 MW dc and 900 MW dc, respectively. These three power plants are fueled with medium-Btu gas, as are all plants, except in the base case (high-Btu) and in parametric Point SE19, in which low-Btu gas from an integrated gasifier system is employed. A steam bottoming plant, is included in the 900 MW power plant, SE4, similar to that described for MC4 in the previous subsection. With medium-Btu gas, the average cell voltage at 400 mA/cm² (372 A/ft²) is 0.66 V, corresponding to a voltage efficiency of 80%. The plant layout corresponding to SE4 is shown schematically in Figure 13.8.

Point SE5 explores the effect of the replacement of air as the oxidant by oxygen from a dedicated oxygen plant (see Figure 13.9). The average cell voltage increases to 0.76 V, as most of the concentration polarization in the fuel cell battery is associated with the cathode. Points SE6 through SE8 explore the effects of increases in the useful life of the fuel cell batteries. The effect of increased power conditioning costs associated with an increase in the permissible voltage degradation and described in the previous subsections is investigated in Point SE9.

The changes represented by Points SE10 through SE20 are specific to the solid electrolyte power system and are presented in Table 13.5.



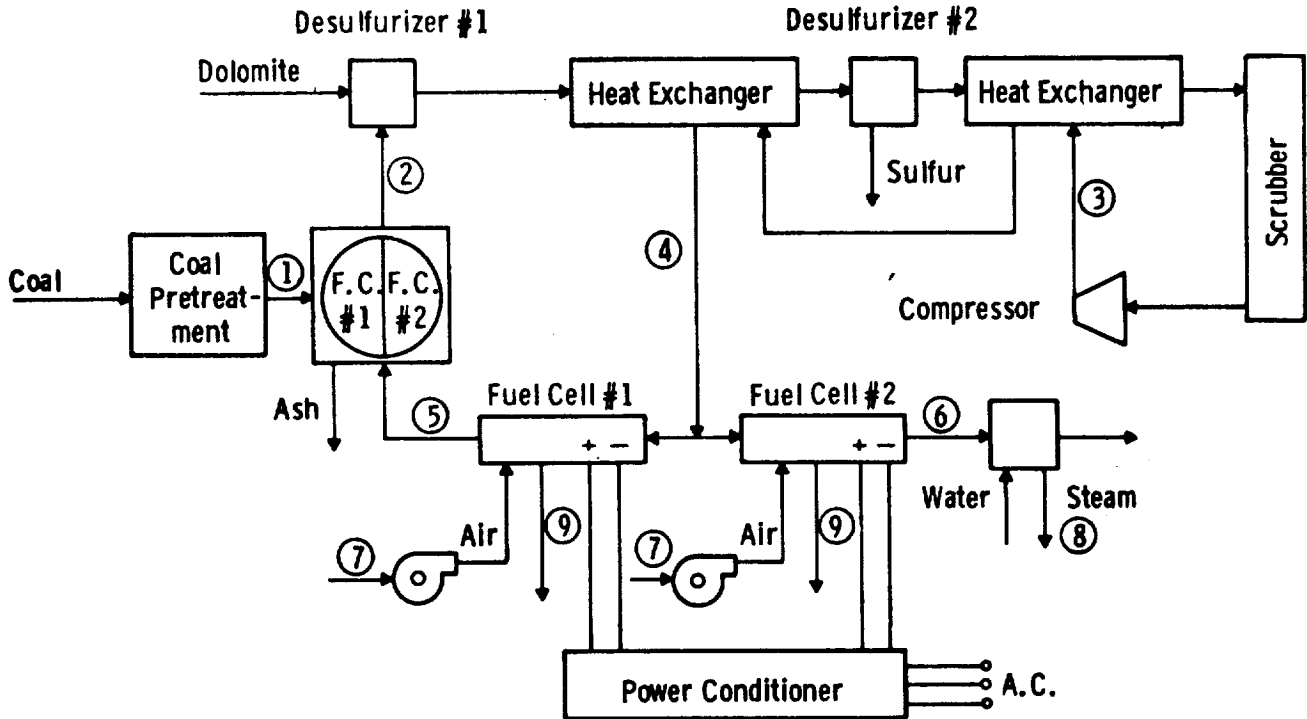
Point No.	Temperature (°F)	Pressure (psia)	Point No.	Temperature (°F)	Pressure (psia)
1	80	60	7	200	15
2	1100	45	8	70	650
3	1300	30	9	1000	600
4	80	60	10	70	15
5	1100	45	11	70	15
6	1300	30			

Fig. 13.9—Schematic of a solid-electrolyte fuel cell power plant employing oxygen as an oxidant. (SE5)

Table 13.5 - Parametric Points - Solid-Electrolyte Fuel Cell

Point No.	Parameter	Value/Type
10	Current density	600 mA/cm ² (557 A/ft ²)
11	Current density	800 mA/cm ² (743 A/ft ²)
12	Electrolyte thickness	20 μm (0.8 mils)
13	Interconnection material	Mn-doped CoCr ₂ O ₄
14	Electrolyte material	Calcium-stabilized zirconia
15	Air electrode material	Sb-doped SnO ₂
16	Temperature	1173°K (1652°F)
17	Temperature	1373°K (2012°F)
18	Type of fuel	Coal
	Current density	800 mA/cm ² (743 A/ft ²)
	Power plant size	250 MW dc
	Electrolyte thickness	20 μm (0.8 mils)
19	Type of fuel	Low-Btu gas
	Power plant size	900 MW dc
20	Current density	800 mA/cm ² (743 A/ft ²)
	Electrolyte thickness	20 μm (0.8 mils)
	Type of fuel	High-Btu gas

Points SE10 and SE11 investigate the effect of battery operation at higher power densities. At current densities of 600 and 800 mA/cm² (557 and 743 A/ft²), the cell voltages were 0.59 and 0.51 V, respectively, because of increased ohmic and concentration polarization losses. A reduction of electrolyte thickness in SE12 from 40 to 20 μm (1.6 to 0.8 mils) leads to a cell voltage increase from 0.66 to 0.67 V. The effect of the replacements of chromium sesquioxide by manganese-doped cobalt chromite, and of tin-doped indium oxide by antimony-doped tin oxide, are explored in SE12 and SE14. Substitution of yttria-stabilized



Point No.	Temperature (°F)	Pressure (psia)	Point No.	Temperature (°F)	Pressure (psia)
1	70	—	6	1870	23.5
2	1750	19	7	70	15
3	200	35	8	1000	600
4	1550	35	9	200	17
5	1650	35			

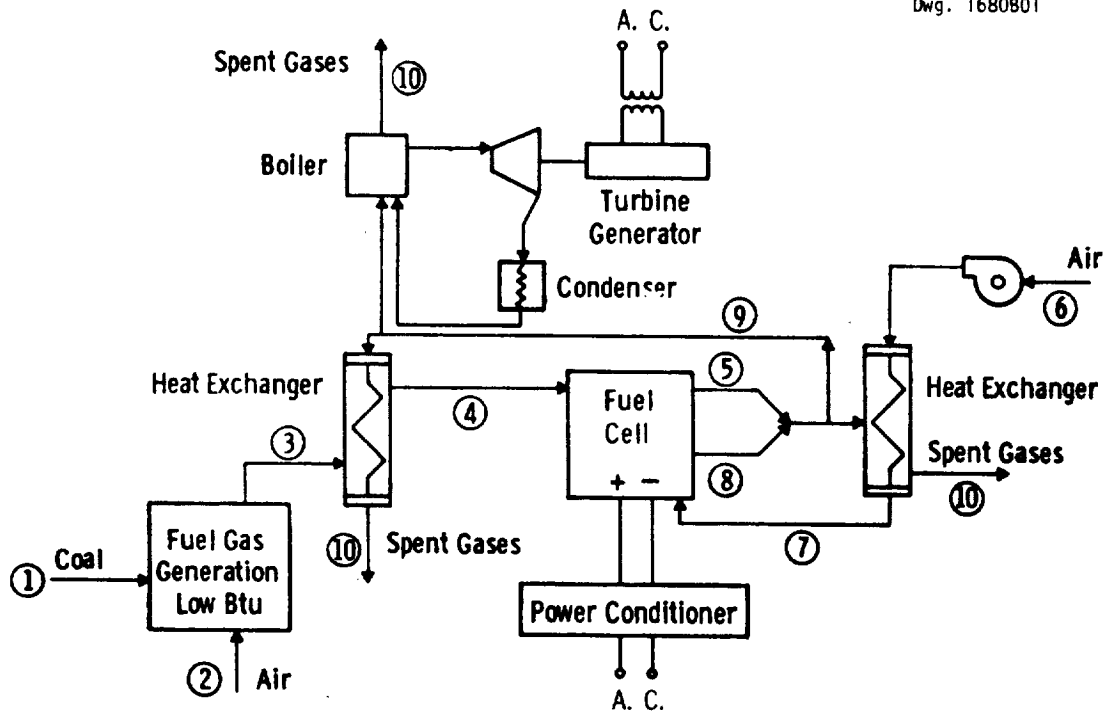
Fig. 13. 10—Schematic of the Westinghouse fuel cell power system involving thermal coupling of the fuel cell subsystem and the coal gasifier (SE 18)

zirconia by calcia-stabilized zirconia in SE13 leads to a reduction of the cell voltage from 0.66 to 0.64 V because of the higher resistivity of the better electrolyte. If the fuel cell subsystem operates at 1173°K (1652°F), resistive losses in the electrolyte and interconnection regions lead to a reduction in cell voltage from 0.66 to 0.55 V. Conversely, an increase in the temperature of operation to 1373°K (2012°F), as in SE17, results in an increase in cell voltage to 0.77 V.

The Westinghouse-OCR fuel cell power generation system is represented by Point SE18. In a 250 MW dc plant, shown schematically in Figure 13.10, a coal gasification reactor and the fuel cell subsystem are thermally coupled. The heat released because of thermodynamic and electrochemical inefficiencies in the fuel cell batteries is employed to gasify coal. The fuel gas thus generated is then fed to the fuel electrodes of the fuel cell subsystem. All the oxygen reaching the gasifier as carbon dioxide and water vapor enters through the fuel cell electrolyte. An electrolyte thickness of 20 μm (0.8 mils), an electrolyte current density of 800 mA/cm^2 (743 A/ft^2), and an average cell voltage of 0.68 V are assumed. Point SE19 explores the advantages and disadvantages of coupling a 900 MW dc fuel cell subsystem with a low-Btu coal gasifier, as shown schematically in Figure 13.11. Because of the nitrogen diluent in the fuel gas, the fuel electrode concentration polarization is relatively higher. Consequently, the cell voltage is lowered to 0.56 V. Point SE20 investigates the effect of a higher current density, 800 mA/cm^2 (743 A/ft^2), and a reduced electrolyte thickness, 20 μm (0.8 mils). Even with high-Btu gas as fuel, the average cell voltage is 0.69 V compared with 0.84 V in the base case.

13.3 Approach to Efficiency Calculations

Efficiency calculations without the detailed mass and energy balances provided by a comprehensive conceptual design study are, of necessity, approximate. Thus, for Task I, these calculations have been performed with a number of assumptions, some with less justification than



Point No.	Temperature (°F)	Pressure (psia)	Point No.	Temperature (°F)	Pressure (psia)
1	70	—	6	70	15
2	70	75	7	1100	60
3	100	60	8	1300	30
4	1100	45	9	1300	45
5	1300	30	10	250	15

Fig. 13. 11—Schematic of a solid-electrolyte fuel cell power system employing low - Btu gas as a fuel. A steam bottoming plant is employed for waste heat recovery. (SE19)

desirable. These assumptions, presumably, will undergo further exploration and clarification as part of Task II.

The approaches for the aqueous acid and alkaline power systems were similar; those for the high-temperature systems were specific to the system. Where possible, the approach was consistent, e.g., in the efficiency assumptions for the power conditioning and fuel processing subsystem.

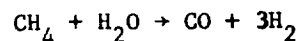
13.3.1 Phosphoric Acid Fuel Cell Power System

The power plant efficiency is defined as the quotient of the power output at grid voltages and the higher heating value (HHV) of the fuel (e.g., high-Btu gas etc.) fed to the power system. Similarly, the overall energy efficiency is the quotient of the power output and the HHV of the coal required to produce the fuel.

High-Btu gas, medium-Btu gas and methanol were fuels employed in the acid fuel cell power plants. These are discussed in the first subsection. In the subsequent subsection, the electrical losses associated with the various parasitic plant subsystems are detailed. The methods of calculating the power plant and overall energy efficiencies are discussed in the final subsection.

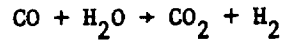
13.3.1.1 Quantities of Fuel Required

All parametric points involve the use of high-Btu gas, with the exceptions of AC4 (medium-Btu gas) and AC14 (methanol). High-Btu gas must be reformed with steam to accomplish the following reaction:



This reaction is endothermic, having an enthalpy of 226 kJ/g-mole (97,380 Btu/lb-mole) at 1144°K (1600°F). The exit stream from the reformer cannot be fed directly to the fuel cell subsystem because the fuel electrode may be deactivated by the presence of carbon monoxide, as discussed in Section 13.1. Consequently, it is passed through a shift

converter in which the mildly exothermic reaction



is caused to occur at approximately 700°K (800°F). The carbon dioxide/carbon monoxide ratio in the shift converter exit stream is considered to be approximately 50 to 1. The fuel processing subsystem is discussed further in Appendix A 13.1.

The molar compositions of the high-Btu gas fed to the reformer (Reference 13.34) and to the fuel cell subsystem (after reformation and shift conversion) are given in Table 13.6.

Table 13.6 - Fuel Gas Molar Compositions

Component	Initial Composition	After Shift Conversion
H ₂	2.49	377.55
CH ₄	94.23	-
CO ₂	0.39	92.84
CO	0.08	1.86
N ₂	2.81	2.81
S	-	-
HHV	959.2 Btu/scf	-
LHV	864 Btu/scf	-
Molecular Weight	16.15	-

Thus, 1 std m³ of high-Btu gas yields 3.776 std m³ of hydrogen. Hydrogen comprises 79.5% of the exit stream from the shift converter, and the concentration of carbon monoxide is 0.39%.

The hydrogen requirements of the fuel cell subsystems are a function of the dc ratings of the subsystems, the average cell voltage, and the hydrogen utilizations. The four ratings are 25 MW dc (all parametric points with the exception of AC2, AC3, AC4, and AC16), 100 MW dc (AC2), 250 MW dc (AC3 and AC16), and 900 MW dc (AC4). To deliver one kilowatt dc from a fuel cell operating at a terminal voltage of 0.7 V requires a cell current of 1429 A. Thus, 1429/96,489 equivalents or 14.9 mg/s (3.29×10^{-5} lb/s) of hydrogen are required if the utilization of hydrogen in the fuel gas is 100%. Published data (Reference 13.35) for phosphoric acid fuel cells, however, indicated that hydrogen utilization of 75% represented the state of the art in 1966-67. For simplicity, this treatment assumes a hydrogen utilization of 90% based on the assumptions of a better shift-conversion capability so that the minimum hydrogen-carbon monoxide ratio constraint (i.e., at near-exhaust compositions) can be met at this utilization, and an advance in fuel electrodes technology leading to more porous, thinner electrodes with lower concentration polarizations. Thus, the actual hydrogen requirement is 59.60 g/kWh (0.1314 lb/kWh), corresponding to a high-Btu gas requirement of 51.64 std cm³/kJ (6.56 scf/kWh).

As shown in Figure 13.1, the high-Btu gas stream, fed to the steam-methane reformer, is split in two parts. One part is converted to hydrogen, as described above, while the other is mixed with air and used to provide the heat required for the endothermic reformation process. It can be shown that, under ideal conditions, the minimum volume of the high-Btu gas necessary for firing comprises 22% of the total feedstock. Practical systems in 1963 (Reference 13.36) used 41% of the feedstock in this manner. In this study, a value of 30% has been assumed on the basis of

- Firing the hydrogen not utilized in the fuel cell subsystem to provide some of the reformer heat requirement
- Advances in reformer technology.

Thus, high-Btu gas is fed to the fuel processing subsystem at the rate of 73.49 std cm³/kJ (9.34 scf/kWh) of dc output from the fuel cell subsystem.

The medium-Btu gas, used as a feedstock in the 900 MW dc power plant, does not require reformation, but it must be shift-converted to yield a fuel gas suitable for use in the fuel cell subsystem. The compositions before (Reference 13.37) and after shift conversion are shown in Table 13.7.

Table 13.7 - Medium-Btu Fuel Gas Molar Compositions

Component	Initial Composition	After Shift Conversion
H ₂	0.3276	0.8618
CO ₂	0.0573	0.5915
CO	0.5460	0.0118
N ₂	0.0043	0.0043
H ₂ O	0.0647	--
HHV	281.5 Btu/scf	--
LHV	265.0 Btu/scf	--
Molecular Weight	19.76	--

Because of the higher concentration of carbon monoxide in the exit stream of the shift converter, a hydrogen utilization of 80% has been assumed in this case. Thus, the overall medium-Btu gas requirement is 253.6 std cm³/kJ (32.24 scf/kWh) of dc output from the fuel cell subsystem.

The methanol required in parametric point AC14 may be shown to be 0.1183 g/kJ [426 g/kWh (0.939 lb/kWh)] with the assumptions employed in the above discussion of high-Btu gas reformation and shift conversion.

13.3.1.2 AC Power Outputs

An efficiency of 95.5% was assumed for the force-commutated inverter power-conditioning subsystem of all 25 MW dc power systems. The rationale for the selection of this type of power-conditioning equipment is outlined in Appendix A 13.2. In addition, in the acid power system a further 2% of the ac output of the power-conditioning equipment was allocated for air blowers and sundry control systems. Thus, the net ac power output from the acid power system, with a dc rating of 25 MW, is $(25 \text{ MW})(0.955)(0.98)$, or 23.4 MWe.

In the cases of the 100 MW (AC2), 250 MW (AC3 and AC16), and 900 MW dc (AC4) power plants, a line-commutated inverter was employed in the power-conditioning subsystem (see Appendix A 13.2). An efficiency of 95% was assumed. Double transformation was considered necessary in the 100 MW and 250 MW dc systems, since the contract statement of work specified that all power plants of greater than 50 MWe should deliver ac power to the grid at 500 kV. It was assumed that the added transformer would operate at an efficiency of 99.5%. The 900 MW dc system was considered not to need this double transformation. As in the case of the 25 MW system, 2% of the ac output was assigned to the air blowers, etc. Thus, the net ac outputs were 92.6 MWe, 231.5 MWe and 838 MWe, respectively.

When a dedicated oxygen plant is employed, as in parametric points AC5 (25 MW dc) and AC16 (250 MW dc), the ac output to the grid must be reduced further. The power required for oxygen plant operation may be calculated from information given in Appendix A 13.3. The power plant in AC5 requires approximately 3.15 kg/s (300 tons/d) of oxygen. The energy required is 0.0975 kJ/Mg (318.4 kWh/ton) or 95.52 MWh corresponding to a power usage of 3.98 MW. With a 4% allowance for oxygen vented to the atmosphere, the power required is 4.14 MW. The net ac power output of a 25 W dc power plant employing oxygen as the oxidant (AC5) is 19.3 MW. Similarly, the net ac power output from the power plant of parametric point AC16 is 190.3 MW.

13.3.1.3 Calculation of Efficiencies

The heat rates of the power plants may be calculated as follows:

$$\text{Heat Rate} = \frac{(\text{Fuel Rate})(\text{HHV of Fuel})}{(\text{Net ac power output})/(\text{Nominal dc rating})}$$

where the required fuel rate and the fuel HHV are given in units of scf/kWh and Btu/scf, or lb/kWh and Btu/lb, respectively. For example, for AC1:

$$\text{Heat Rate} = \frac{(9.34 \text{ scf/kWh})(959.2 \text{ Btu/scf})}{(23.4 \text{ MW})/(25 \text{ MW})} = 9570 \text{ Btu/kWh}$$

13.3.2 Alkaline Fuel Cell Power System

The approach to the calculation of heat rates in the alkaline power system is very similar to that employed for the acid system. The higher cell voltages in alkaline cells (0.8 V vs 0.7 V for acid fuel cells) result in a lower hydrogen requirement, 0.0145 g/kJ [52.2 g/kWh (0.1151 lb/kWh)], and a lower rate at which fuel is fed to the power system. The assumptions of the efficiencies of the power-conditioning subsystems are identical to those given in the previous section and in Appendix A 13.2.

The Lurgi Rectisol process for the removal of carbon dioxide from both fuel and air streams before they are fed to the fuel cell subsystems has been arbitrarily assumed to consume an additional 6% of the ac power output from the power-conditioning subsystem when high-Btu is used as feedstock. When added to the 2% consumed by the air blowers and control systems, the inefficiencies amount to 8% of the total ac output. In parametric point AL4, the use of the medium-Btu feedstock results in a doubling of the carbon dioxide partial pressure in the shift converter exit stream. The inefficiencies are now assumed to total 10% of the ac output of the power-conditioning subsystem.

The oxygen requirements in the power plants corresponding to parametric point AL5, in which oxygen from a dedicated oxygen plant is employed as the fuel cell oxidant, are lower than in the similar acid plant (AC5) because of the higher cell voltages. The power requirement is 3.6 MW, calculable from the value given for AC5 by multiplication by the reciprocal of the ratio of the cell voltages, 0.875.

Thus, all the heat rates for the parametric points of the alkaline fuel cell power system are calculable from the corresponding values of the acid system. For example, in the base case, AL1,

$$\begin{aligned} \text{Heat Rate} &= \frac{(9570 \text{ Btu/kWh})(0.7 \text{ V}/0.8 \text{ V})}{(92\%/98\%)} \\ &= 8920 \text{ Btu/kWh} \end{aligned}$$

13.3.3 Molten Carbonate

The plant efficiency is calculated on the basis of the higher heating value of the fuel and of steady-state operation at the voltages and current densities described in Section 13.2.3.

All cells produce an excess of heat during operation at rated power over that necessary to heat the fuel and oxidant streams to the fuel cell temperature. This excess heat is partially converted to steam but does not contribute to the power plant efficiency for MC1-3, and 6-16. For MC4, this excess heat is used to operate a steam turbine generator to produce additional ac electricity and thereby increase the overall plant electrical efficiency. MC5 and 17 also use the excess heat, but to operate a steam turbine drive for the compressor in the oxygen plant. The steam turbine drive was assumed to convert the available energy to shaft work with an efficiency of 40%. The efficiency of plants using pure oxygen was not reduced by 2% since the electrical output is not derated to provide power for the electrical motor drive compressor that is commonly used for an oxygen plant. The nominal dc output of the fuel cell is further derated by the dc-to-ac power conversion subsystem as described in Appendix A 13.2. It is assumed that 90% of the fuel is

consumed by the fuel cell. The remaining 10% is burned in the exit gases by excess oxygen from the cathode exit gas and appears as waste heat.

It is further assumed that various electrical auxiliaries (such as recirculating fans) consume electrical energy equivalent to 3% of the HHV of the fuel. This may be high.

For simplicity, reformed high-Btu gas or methanol are assumed to have only carbon monoxide and hydrogen as fuel species after reforming. Some CH_4 and other fuel species, however, will, in practice, be present.

Equation 13.1 is used to calculate all molten carbonate fuel cell efficiencies except for MC4.

$$\text{Efficiency} = \frac{0.9 [(\Sigma N)(n)(F)(\epsilon)] \text{PCE} - 0.03 \text{HHV}}{\text{HHV}} = \frac{(A)(\text{PCE}) - B}{\text{HHV}} \quad (13.1)$$

where 0.9 is the fraction of the fuel utilized by the fuel cell,
 ΣN is the sum mole fraction of carbon monoxide and hydrogen present after reforming 1 mole of the original fuel gas,
 n is 2 (the number of electrons involved in electrochemical oxidation of 1 molecule of carbon monoxide or hydrogen with 1/2 molecule oxygen),
 PCE is the power-conditioning efficiency (see Appendix A 13.2),
 HHV is the higher heating value per mole of original fuel gas,
 F is the Faraday 23.062 kcal/eV,
 ϵ is the cell voltage.

For MC4 we utilize the waste heat to increase the plant efficiency. It is assumed that we exchange heat between exhaust gas and incoming fuel and air, with the exhaust leaving the stack at 423°K (302°F); and that other heat loss to the surroundings is 2% of the HHV per mole of fuel gas. It is further assumed that a steam turbine generator with 40% efficiency is used to convert the waste heat to ac electricity. We calculate the overall efficiency of MC4 by Equation 13.2.

Overall Efficiency MC4 =

$$\frac{(A)(PCE) - B + 0.4(\Delta H_{650^\circ}^C - A - 0.02 \text{ HHV} - [\Delta H_{650^\circ} - \Delta H_{25^\circ}]_{i.g.} + [\Delta H_{650^\circ} - \Delta H_{150^\circ}]_{e.g.})}{\text{HHV}}$$

(13.2)

where 0.4 is the turbine-generator efficiency,
 $\Delta H_{650^\circ}^C$ is the heat of combustion of the fuel gases at 650°C,
 $[\Delta H_{650^\circ} - \Delta H_{25^\circ}]_{i.g.}$ is the sensible heat difference in the given
 °C temperature interval for the input gases to the fuel cell,
 $[\Delta H_{650^\circ} - \Delta H_{150^\circ}]_{e.g.}$ is the sensible heat difference in the given
 °C temperature interval for the exhaust gases from the plant.

Other symbols are as in Equation 13.1. ΔH values were taken from the JANAF Thermochemical Tables (Reference 13.38).

13.3.4 High-Temperature Solid Electrolyte Fuel Cell

13.3.4.1 Thermodynamic Efficiency

The quotient of average cell voltage and the voltage corresponding to the higher heating value of the fuel is defined as the thermodynamic efficiency η_{th} .

$$\eta_{th} = \frac{E_{cell}}{E_{HHV}}$$

E_{cell} was calculated in the cases SE2-SE18 on the basis of actual measurements on cells and batteries.

$$E_{cell} = E_O - E_R - E_P$$

where E_O is the open-cell voltage, average over the total range of fuel/combustion product ratios,
 E_R is the voltage losses due to ohmic resistance in battery components,
 E_P is the polarization loss due to diffusion problems of fuel and combustion products.

The voltage which corresponds to the higher heating value of the fuel, E_{HHV} is calculated from the following relation:

$$E_{\text{HHV}} = \frac{\text{HHV}}{n \cdot F}$$

where HHV is the higher heating value of fuel in kcal/mole,
 n is the number of electrons transferred per mole of fuel gas reacted,
 F is the Faraday constant 23.06 kcal/eV.

For SE1 and SE20 an average open-cell voltage was calculated from the free energy change of the oxidation of methane at 1300°K (1880°F). In this instance, a high average open-cell voltage of 1.04 V is achieved. The reason for this is the thermodynamic instability of methane at 1300°K (1880°F). Kinetically, however, it is possible to burn methane electrochemically at this temperature without carbon deposition if, except for the fuel electrode, the cell contains no metal surfaces. An in-situ reformation and oxidation of the methane takes place at the anode. The question, however, remains whether this concept is practical, as experimental data are limited.

13.3.4.2 Power Output of the Plant

Since the ac power output is included in the heat rate calculations, we must explain how we derived these figures. The power output of the fuel cell subsystem is reduced by the inefficiencies of the power-conditioning subsystem, as described in Appendix A 13.2. As in the case of molten carbonate fuel cells, it is assumed that an amount of energy equivalent to 3% of the HHV of the fuel is consumed to provide plant auxiliary power. The power output P_{net} was calculated according to the following equation

$$P_{\text{net}} = (\text{MW})(\text{PCE}) - (0.03) (\text{MW}/\eta_{\text{th}}) + P_{\text{turb}}$$

where MW is the nominal power output,
PCE is the power conditioning efficiency,
 P_{turb} is the ac power output from steam turbine generator
 η_{th} is the thermodynamic efficiency.

13.3.4.3 Heat Rate

Heat rate (HR) calculations for all parametric points except SE18 were performed in the following sequence:

1. Establish dc power rating in kilowatts
2. Calculate thermal equivalent from (1) in Btu/hr
3. Calculate fuel rate from (2) by considering thermodynamic inefficiencies in Btu/hr
4. Calculate heat rate in Btu/kWh in dividing (3) by the actual ac power output of the power plant.

For example, for a Point SE1

$$\text{HR} = \frac{\text{MW}}{\eta_{\text{th}} P_{\text{net}}} = \frac{(25 \times 10^6)(3413)}{(0.76)(22.9 \times 10^6)} = 4900 \text{ Btu/kWh}$$

The heat rate for SE18 was established earlier by Westinghouse under the sponsorship of the Office of Coal Research. Based on a detailed mass and energy balance for a 100 kW power plant (Reference 13.25) a heat balance of 6370 Btu/kWh was calculated. This heat rate was used in the present calculations taking no credit for possible improvements as the plant size was increased from 100 kW to 250 MW.

13.3.4.4 Power Plant Efficiency

The power plant efficiency, η_{pp} , is the quotient of the theoretical and calculated heat rates for all parametric points:

$$\eta_{\text{pp}} = \frac{3413}{\text{HR}}$$

In the cases of SE4 and SE19, additional power is produced by utilizing the waste heat of the fuel cell generator. This, of course, is only

possible in large installations and was not considered for smaller plants. The waste heat was used via a steam turbine generator with an assumed 40% efficiency. This is why the power plant efficiency is higher than the thermodynamic efficiency of the electrochemical generator. This is not a contradiction, because the fuel cell for which the thermodynamic efficiency is calculated delivers only a part of the electrical energy of these plants which could be considered combined-cycle plants. The heat rates of SE4 and SE19 must be viewed this way also.

13.4 Capital, Site-Labor, and Operation and Maintenance Costs

This section is devoted mainly to a description of the approaches taken in calculating the capital costs of the fuel cell subsystems. The power-conditioning costs for the power plants corresponding to all 69 parametric points are given in Appendix A 13.2. Similarly, oxygen plant costs for Point 5 in all fuel cell power systems and for AC16 and MC17 are discussed in Appendix A 13.3. The approach to balance of plant costing has already been described in Section 2. The fuel-processing subsystem, required for the low-temperature aqueous acid and alkaline power systems, is costed as described in Appendix A 13.1. This appendix also includes a brief description of costing the air blowers for the low-temperature fuel cell power systems.

It must be emphasized here, as in Subsection 13.3, that without a conceptual design of the fuel cell modules the costing techniques are at best approximate. The assumptions employed as to design and materials of construction are stated, even if extensive justification is not provided. Refinement of these assumptions was anticipated as part of Task II of this study.

The site labor costs for installing the four types of fuel cell module were assumed arbitrarily to be \$5/kW for the acid and alkaline systems, \$8/kW for the molten carbonate system, and \$10/kW for the solid electrolyte system. These relatively low site labor charges reflect the modularity of the assumed fuel cell subsystem, which leads to relatively

straightforward and simple installation procedures at the power plant location. The higher charges for the high-temperature system reflect the slightly more complex procedures of subsystems interconnection, because of increased insulation requirements.

The operation and maintenance (O&M) costs for the fuel cell power plants have, in general, three component charges related to: (a) plant labor; (b) material and component replacement in subsystems other than the fuel cell subsystem, and (c) full replacement of the fuel cell subsystem. These costs are described in greater detail below.

(a) An hourly labor cost is calculable, as described already in Section 2.6.3.4, by the formula:

$$\text{O\&M (labor)} = Z \left(\frac{\text{ac output}}{0.6 + 0.004 (\text{ac output})} \right) \left(\frac{\$15,000}{8740 \text{ hr (capacity factor)}} \right)$$

where Z, a factor related to the complexity of O&M procedures, is taken as 1.0 for steam plants. For all plants with a dc rating of 25 MW, it has been assumed that, because of their probable substation locations, remote operation will be possible with control exercised from a central station. Also, because of this probable remote control, the value of Z has been arbitrarily assumed to be 0.2 for these plants. For all power plants of 250 MW or larger, however, a factor of 0.4 has been chosen, reflecting the need for full-time personnel at the plant location. This Z value is substantially less than that for steam plants and may be justified on the basis of the relative cleanliness of the fuels employed in the fuel cell plants, and the ease of maintenance of the power-conditioning subsystem.

(b) To allow for material and component replacement in the fuel-processing and power-conditioning subsystems, and other major components, an O&M charge, amounting to 5% of the total capital and site-costs of the major components (with the exception of the fuel cell subsystem), is allotted for each power plant.

(c) The fuel cell subsystem useful life is only 36 Ms (10,000 hours) (exceptions: Points 6 through 9). This is much less than the 946 Ms (30 yr) life assumed for the other plant components. Accordingly, a special charge, calculated by dividing the replacement cost of the fuel cell subsystem by the product of the ac output and the life, is included in the O&M costs for every parametric point. Because of uncertainties, this calculation omits price escalation, learning, and sinking fund factors.

13.4.1 Phosphoric Acid Fuel Cell Power System

A bipolar design has been assumed for the acid fuel cell battery stack. The basic element is a corrugated or embossed graphite plate which serves as the anode current collector of one cell and the cathode current collector of an adjacent cell. The anode and cathode are assumed to be thin carbon electrodes, similar to those described by Kordesch and Scarr (Reference 13.30). These contain active carbon catalyzed by platinum. The electrolyte consists of a matrix-immobilized 85 wt % phosphoric acid. The fuel gas stream flows in the corrugations between the thin carbon anode and the graphite piece. Air flows similarly between the cathode and the corrugated graphite plate.

The cost of the corrugated graphite plate, which may be fabricated by extrusion, is estimated (Reference 13.39) to be approximately $\$10.76/\text{m}^2$ ($\$1/\text{ft}^2$). If battery stack end requirements are ignored only one of these plates is required per cell. The active carbon electrodes, without catalyst, are similarly estimated (Reference 13.39) to cost approximately $\$10.76/\text{m}^2$ ($\$1/\text{ft}^2$) each. For an inter-electrode separation (i.e., electrolyte thickness) of 0.5 mm (19.7 mils), the quantity of electrolyte required, assuming that it forms 80% by volume of the matrix, is approximately 100 g (0.22 lb). At $\$0.349/\text{kg}$ ($\$15.85/100$ lb) (Reference 13.40), the electrolyte cost is only $\$0.377/\text{m}^2$ ($\$0.035/\text{ft}^2$). The electrolyte matrix component has been assumed to be cost-determining, so that the cost of the immobilized phosphoric acid is taken as $\$5.38/\text{m}^2$ ($\$0.50/\text{ft}^2$). Thus, the cost of the bipolar battery stack materials, unassembled and without catalyst, is approximately

$\$37.67/\text{m}^2$ ($\$3.50/\text{ft}^2$). Doubling of this materials cost results in a manufactured cost for the battery stack of $\$75.35/\text{m}^2$ ($\$7/\text{ft}^2$). A further allowance of $\$5.38/\text{m}^2$ ($\$0.50/\text{ft}^2$) for gasketing and insulating the battery stack yields a manufactured cost of $\$80.73/\text{m}^2$ ($\$7.50/\text{ft}^2$) for the fuel cell module. Adding a profit of 20% results in a selling price of $\$96.88/\text{m}^2$ ($\$9/\text{ft}^2$) for a module with uncatalyzed electrodes.

Most parametric points (except for AC12 and AC13) involve platinum loadings in the anode and cathode of $1 \text{ mg}/\text{cm}^2$ ($0.00205 \text{ lb}/\text{ft}^2$). If chloroplatinic acid (H_2PtCl_6) is employed in the catalyst-addition process, the cost of platinum per unit cell area is $\$117.33/\text{m}^2$ ($\$10.90/\text{ft}^2$), for a chloroplatinic acid cost of $\$5.8675/\text{g}$ of contained platinum ($\$182.48/\text{oz t}$ platinum) (Reference 13.40). Thus, for most parametric points, the fuel cell module cost is $\$214.20/\text{m}^2$ ($\$19.90/\text{ft}^2$). In Point AC12, where the platinum loadings in the anode and cathode are decreased to $0.3 \text{ mg}/\text{cm}^2$ ($6.144 \times 10^{-4} \text{ lb}/\text{ft}^2$), the catalyst cost is reduced to $\$35.20/\text{m}^2$ ($\$3.27/\text{ft}^2$), resulting in a fabricated module cost of $\$132.07/\text{m}^2$ ($\$12.27/\text{ft}^2$). The catalyst and module costs for AC13 [$0.1 \text{ mg Pt}/\text{cm}^2$ ($2.05 \times 10^{-4} \text{ lb}/\text{ft}^2$)] are calculated in the same manner.

The cost per kW of the fuel cell module is assumed to be inversely proportional to the power density. For AC1, the power density is $140 \text{ mW}/\text{cm}^2$ ($130.1 \text{ W}/\text{ft}^2$). This corresponds to a cell area requirement of $0.7143 \text{ m}^2/\text{kW}$ ($7.689 \text{ ft}^2/\text{kW}$), and a module cost of $\$153/\text{kW}$, based on the costs given above. This leads to a vendor selling price of $\$3.8$ million for the 25 MW dc fuel cell subsystem. The costs of all fuel cell subsystems are calculated in the same manner, based on the cell power densities, the dc ratings, and the catalyst loadings.

The replacement cost of the acid fuel cell subsystem is calculated under the following assumptions:

- The cost of recovering the platinum at the end of battery life and reprocessing it into chloroplatinic acid is $\$1.00/\text{g}$ ($\$31.10/\text{oz t}$).

- The rest of the fuel cell module has zero salvage value.
- The installation cost remains at \$5/kW.

Thus, the catalyst cost in the replacement modules is only \$20/m² (\$1.86/ft²), resulting in a fabricated module cost of \$116.88/m² (\$10.86/ft²). In parametric point AC1, the replacement cost of the 25 MW dc fuel cell subsystem is \$2,087,000 plus \$125,000 for installation. The corresponding O&M charge is:

$$\frac{\$2,087,000 + \$125,000}{(23,400 \text{ kW})(10,000 \text{ hr})} = 9.45 \text{ mills/kWh}$$

or a charge of \$221/operational hour.

13.4.2 Alkaline Fuel Cell Power System

A bipolar design, identical to that selected for the acid system, has been assumed also for the alkaline fuel cell modules. Thus, the selling price of \$96.88/m² (\$9/ft²) for a module with uncatalyzed electrodes given for the acid fuel cell system is applicable here also. The major differences between the acid and alkaline modules lie in the areas of cathodic catalyst type and power density.

The carbon anode is catalyzed by platinum with a loading of 1 mg/cm² (0.00205 lb/ft²), and the cathode is catalyzed by silver with a loading of 5 mg/cm² (0.0103 lb/ft²). Thus, the platinum loading is half of that in the acid system and adds \$58.68/m² (\$5.45/ft²) to the uncatalyzed cost quoted above. It is assumed that the cathodes are silver catalyzed with silver nitrate, which costs \$0.094/g of contained silver (\$2.69/oz av Ag (Reference 13.40)). This adds a further \$4.745/m² (\$0.4408/ft²) to the uncatalyzed module cost. The catalyzed cell cost is, therefore, \$160.30/m² (\$14.89/ft²).

The power density for most parametric points is 80 MW/cm² (74.32 W/ft²) leading to a cell area requirement of 1.25 m²/kW (13.46 ft²/kW). The module cost is, therefore, \$200/kW, or \$5 million for a 25 MW fuel cell subsystem. Subsystem costs for other power

density values are calculated on the basis of the inverse relationship assumed between the cost and the power density. The replacement cost is calculated, as in the acid system, with recovery and reprocessing costs for platinum and silver of \$1/g (\$31.10/oz t) and \$0.03/g (\$0.93/oz av), respectively.

13.4.3 Molten Carbonate Fuel Cell Power System

This section describes cost estimates for the fuel cell, gas recirculation, and heat recovery systems. Plant costs for siting, construction, and electrical controls are described in Appendix A 13.4. Oxygen plant costs are described in Appendix A 13.3, and power-conditioning costs are given in Appendix A 13.2.

The same cost for materials, fabrication, and assembly per unit area of electrode has been assumed for all molten carbonate fuel cells. Since there have been no cell assemblies of more than a few kilowatts, and constructional details of current test units (Reference 13.22) are not available, only very rough estimates can be made. As a first approach to costing the full cell, we assume a filter press design; estimate a cost based on the somewhat analogous filter press technology; and approximate cell material costs suggested by IGT (Reference 13.11), with 0.715 m^2 (7.7 ft^2) of electrode area per kilowatt. From Perry's Handbook (Reference 13.41), the lowest cost (1970) of a filter press with filters uninstalled was about $\$161.50/\text{m}^2$ ($\$15/\text{ft}^2$) for iron or wood materials. If we update to July 1974 prices by the factor 1.4, subtract a material cost of about $\$21.50/\text{m}^2$ ($\$2/\text{ft}^2$) for the wood or iron materials, add a cell material cost suggested by IGT in 1966 (Reference 13.11) ($\$20$ to $\$40/\text{kW}$ escalated to July 1974 prices of about $\$31$ to $\$62/\text{kW}$ or an average of $\$46/\text{kW}$), and add a further 20% for assembly and leakproofing, a cost of $(\$15 \text{ ft}^2) (7.7 \text{ ft}^2)(1.4) - (\$2/\text{ft}^2) 7.7 \text{ ft}^2 - \$15.4 + [31 \text{ or } 62]$, i.e., $\$177$ or $\$208/\text{kW}$, may be calculated. If we add 20% to allow for vendor profit, the cost would be $\$213$ to $250/\text{kW}$, uninstalled.

A second approach to costing the cell is to take the values estimated by Hart and Womack (Reference 13.16). The list price was $\pounds 27.5/\text{kW}$ dc. Adjusting this for an exchange rate of $\$2.4/\pounds$ (in 1966),

for the ratio of power densities assumed in that and the present study of 0.2 to 0.13 (in terms of kW/ft²), a commodity escalation price of \$1.58 and a profit factor of 1.2 for fabrication gives (27.5)(2.4)(0.2/0.13)(1.58)(1.2) = \$192.51/kW uninstalled.

For this study we have arbitrarily selected an uninstalled capital cost of \$190/kW as being about the lower limit of what we might reasonably expect. An installation cost of \$8/kW is assumed.

Without a conceptual design of the fuel cell system, costing of the waste heat recovery system is also arbitrary. Refinement of the estimates on the basis of a detailed design was anticipated as a part of Task II. For example, normal waste heat boilers have a once-through passage of the hot gases, but in the present system we recycle the fuel cell exit gases between the fuel cell and boiler temperatures. Such a system would have to be specifically designed for this application. The installed cost of the heat recovery steam generator is assumed to be \$60/kW. Installation cost is estimated as 33% of the total, which is about the same as for the simple steam turbine power plant boiler but much more than the ~ 5 to 10% for a simple once-through waste heat boiler. The cost of the associated turbine-generator combination in MC4 is taken from a Westinghouse price list (Reference 13.42) adjusted to July 1974 prices.

Heat exchangers are arbitrarily assumed to cost \$12.50/kW dc installed. Of this cost about 45% is for the heat exchangers, and 55% involves the installation labor, piping, etc.

Blower costs will depend on the rate of recirculation necessary both to remove heat and to prevent cell concentration polarization. We assume about \$1.15/kW plus 10% installation for this item for all plants except MC4; and \$2.55 kW for MC4, the 900 MW plant which may require higher temperature blowers.

The sum totals of the blower, heat exchanger, and heat recovery steam generator costs for MC1, the base case, represent only about 3.85% of the total plant capital costs. Substantial errors in

these estimates, therefore, will not be significant in view of the large uncertainty in estimating the ultimate lifetimes and cost of the fuel cell itself, which represents about 38.5% of the total capital costs.

Capital costs are much greater where an oxygen plant (see Appendix A 13.3) is included, even though the fuel cell is significantly reduced in size due to the higher power output per cell that has been assumed. Due to economics of size the oxygen plant represents about 33% of the total capital cost for MC5, a 25 MW plant, but only 20.5% for MC17, a 250 MW plant. This is more than the corresponding fuel cell capital costs of roughly 15 and 19.6% of the total, respectively. The high capital cost in addition to large power requirement appears to preclude the use of oxygen rather than air in the molten carbonate plants.

13.4.4 Solid Electrolyte Fuel Cell Power System

The costing of the fuel cell subsystem is based on a cost analysis for the Westinghouse thin-film battery, discussed in detail in Reference 13.43. For a power output of 0.5 W/cm (1.27 W/in) of a tubular battery having an outside diameter of 1.27 cm (0.5 in), the cost of the battery raw material was \$21/kW in mid-1970. The following set of assumptions have been employed in calculating the cost per kilowatt dc output of a fuel cell module in mid-1974:

- Active electrolyte and active interconnection region lengths are equal to each other and to 2 mm (0.079 in).
- Fuel electrode and air electrode gap lengths are equal 0.5 mm (20 mils).
- A fabrication-cost/materials-cost ratio of 7 to 3 is based on Westinghouse manufacturing experience in thick-film device technology, which is comparable to that required for thin-film battery fabrication.

- A factor of 1.34, corresponding to the increase in the Marshall and Stevens index from mid-1970 to mid-1974, is used to estimate the mid-1974 costs of the fabricated batteries.
- An additional \$18/kW is allowed for materials and assembly labor necessary to manifold and sheath the fuel cell batteries in the fabrication of the fuel cell module.
- Three percent (3%) of the cost of the fabricated fuel cell module is taken as the cost of insulation (Reference 13.44).
- A profit of 20% is assumed in the estimation of a vendor price.

Thus, for SE2-4, SE6-9, SE13, and SE15, the selling price in \$ per kilowatt of a fuel cell module is calculated as follows:

$$\begin{aligned}
 P &= \left(\frac{\$21}{\text{kW}} \left(\frac{0.5 \text{ W/cm}}{0.4224 \text{ W/cm}} \right) \left(\frac{100}{30} \right) (1.34) + \frac{\$18}{\text{kW}} \right) (1.03)(1.2) \\
 &= \$160/\text{kW}
 \end{aligned}$$

For all other points the selling price of the fuel cell module were calculated using the above value and an assumed inverse relationship between cost and power density.

No known heat exchanger technology is available for application in high-temperature fuel cells operating up to 1373°K (2012°F). It was assumed, therefore, that heat exchange above 873°K (1112°F) must be accomplished in the fuel cell generator itself. At this temperature, heat shock problems are largely reduced. Based on this assumption, heat exchanger calculations were performed where nearly a 50% reduction of heat exchange surface is achieved and where the maximum metal temperature in the heat exchanger is reduced to 873°K (1112°F). The calculations took into account a 100°K (180°F) mean temperature difference and a base cost of \$215/m² (\$20/ft²) of heat exchange surface. The installed cost was assumed to be 230% of this base cost, or \$495/m² (\$46/ft²). These costs

were broken down into material and site labor costs, following Guthrie (Reference 13.44), who recommends values of 73% (materials) and 27% (site labor) of the total installed cost.

The cost of the integrated coal gasification reactor in SE18, the Westinghouse Fuel Cell Power System, was estimated on the basis of an economic evaluation of a 200 MW coal-burning fuel cell power plant, performed by IGT in early 1969 (Reference 13.45). The updated installed cost is broken down according to the above Guthrie recommendations for shell-and-tube heat exchangers - 73% for materials and 27% for site labor - because fully 75% of the installed cost of the reactor is attributable to the Incoloy 800 pipe required for encapsulation of the fuel cell modules and for adequate heat exchange between the fuel cell batteries and the coal undergoing gasification. The low-Btu gasifier cost for SE19 was taken from estimates provided in Section 4 of this report.

Estimates of the cost of gas compressors, venturi scrubbers, and waste heat boilers were based on information provided in Reference 13.44. The waste-heat recovery system (i.e., the steam bottoming plant) of SE4 was costed as described in Section 13.4.3.

13.5 Results of Parametric Assessment

The power plant efficiency is defined as the quotient of the power output at grid voltages and the higher heating value (HHV) of the fuel fed to the power system. As described in Section 13.3, a heat rate was calculated for the power plant corresponding to each of the 69 points in the parametric assessment. These values were input to a computer program (described in Section 2) in which they were converted to fractional efficiencies, after a minor allowance was made for power plant electrical requirements.

The program was also employed to calculate an overall energy efficiency, defined as the quotient of the power plant ac output and the higher heating value of the coal required in the production of the fuel.

The factors employed to convert the power plant efficiencies to overall energy efficiencies are described in Sections 2.3.1.2, 2.3.1.3, and 2.3.1.6 for high-Btu gas, medium-Btu gas, and methanol, respectively. All of these fuels were considered to have been derived from Illinois No. 6 coal. Point SE18, the Westinghouse Fuel Cell Power System, involves the use of a coal (Illinois No. 6) as fuel so the power plant and overall energy efficiencies for this system are identical. North Dakota Lignite was selected for the low-Btu gasification reactor of SE19 because of its higher gasification efficiency (Section 4).

The cost of electricity (COE) and its three component parts, ascribable to capital, fuel, and O&M charges, were also calculated by means of the computer program from the cost input described in Section 13.4. These costs were based on NASA-mandated values for labor rate, contingency charge, escalation rate, interest during construction, fixed charge rate, fuel cost and capacity factor. The values specified are listed in Table 13.8.

Table 13.8 - Values Selected for Variables in Plant Construction and Operation

Labor Rate	\$10.60/hr
Contingency Charge	4.5%
Escalation Rate	6.5%
Interest during Construction	10%
Fixed Charge Rate	18%
Capacity Factor	65%
Fuel Costs -	
High-Btu gas	\$2.60/10 ⁶ Btu
Medium-Btu gas	\$2.00/10 ⁶ Btu
Methanol	\$2.70/10 ⁶ Btu
Coal	\$0.85/10 ⁶ Btu

The effect of changes in all of these base case values on the COE were explored for every point in the parametric assessment.

In Sections 13.5.1 through 13.5.4, the efficiency and COE results are shown and discussed for each fuel cell power system. A comprehensive comparison of the four types of fuel cell power systems is given in Section 13.6 in justification of the conclusions and recommendations of this study.

13.5.1 Phosphoric Acid Fuel Cell Power Systems

The power plant and overall energy efficiencies; the capital cost and COE; and the estimated time of construction for each of the plants corresponding to the sixteen parametric points of the acid fuel cell system are shown in Table 13.9. All relevant information pertaining to the operation of each of the fuel cell subsystems in these plants is provided also in this table.

When air is employed as the oxidant, the plant efficiency lies in the range of 35 to 36%, for all points. The overall energy efficiency with high-Btu gas is approximately 24%. For AC4 the overall efficiency is better than 29%, reflecting the use of medium-Btu gas in this power plant. There is no comparable advantage to the use of methanol as a fuel, as shown by the value of 25% for the overall energy efficiency of the power plant corresponding to Point AC14.

Points AC5 and 16, corresponding to 25 MW dc and 250 MW dc power plants which use oxygen instead of air as the oxidant, display power plant and overall energy efficiencies of 30% and 20%, respectively. The efficiency reduction, amounting to a sixth of the total, is attributable to the power required by the dedicated oxygen plants.

The computer output for Base Case AC1 is shown in Tables A 13.5.1 through A 13.5.3. Listed are all of the cost input, as well as the capital cost and COE for five levels of labor cost, contingency charge, escalation rate, interest rate during construction, fixed charge rate, fuel cost, and capacity factor. The capital cost and the COE broken

TABLE 13.9—VALUES OF ALL RELEVANT PARAMETERS FOR THE PARAMETRIC POINTS OF THE AQUEOUS-ACID FUEL CELL POWER SYSTEM

Parametric Point, AC #	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
Power Output, MWe	23.4	92.6	231.5	838	19.3	23.4	23.4	23.4	23.4	23.4	23.4	23.4	23.4	23.4	23.4	180.3
Fuel Cell Rating, MW	25	100	250	900	25	25	25	25	25	25	25	25	25	25	25	250
Fuel																
High-Btu Gas	X	X	X		X	X	X	X	X	X	X	X	X		X	X
Medium-Btu Gas				X												
Methanol														X		
Oxidant																
Air	X	X	X	X		X	X	X	X	X	X	X	X	X	X	
Oxygen					X											X
Fuel Cell Life, 10 ³ hr	10	10	10	10	10	30	50	100	100	10	10	10	10	10	10	10
Voltage Degradation, %	5	5	5	5	5	5	5	5	15	5	5	5	5	5	5	5
Temperature, °C	190	190	190	190	190	190	190	190	190	190	190	190	190	190	190	190
Electrolyte																
Type: 85 w/o H ₃ PO ₄	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X
Thickness, cm	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05
Anode																
Type: Pt/C	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X
Catalyst Loading: mg Pt/cm ²	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	0.3	0.1	1.0	1.0	1.0
Cathode																
Type: Pt/C	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X
Catalyst Loading: mg Pt/cm ²	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	0.3	0.1	1.0	1.0	1.0
Current Density, mA/cm ²	200	200	200	200	400	200	200	200	200	300	400	200	200	200	200	400
Average Cell Voltage, Volts	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Thermodynamic Eff, %																
Powerplant Eff, %	35.5	35.1	35.1	34.8	29.6	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.8	36.0	29.9
Overall Eff, %	23.9	23.6	23.6	23.3	19.9	23.9	23.9	23.9	23.9	23.9	23.9	23.9	23.9	24.9	24.2	20.1
Total Capital Cost × 10 ⁻⁶ , \$	10.4	39.5	97.6	367	16.0	10.4	10.4	10.4	10.8	8.8	7.9	8.6	8.1	10.9	10.2	117
Capital Costs, \$/kWe	448	429	424	440	833	448	448	448	463	377	339	371	346	468	440	595
Cost of Elect, Mills/kWh																
Capital	14.2	13.6	13.4	13.9	26.3	14.2	14.2	14.2	14.6	11.9	10.7	11.7	10.9	14.8	13.9	18.8
Fuel (2)	25.0	25.3	25.3	19.6	29.9	25.0	25.0	25.0	25.0	25.0	25.0	25.0	25.0	25.8	24.7	29.7
Oper. & Maint.	10.9	10.9	10.8	10.5	9.9	4.6	3.3	2.3	2.4	8.0	6.5	9.9	9.6	11.0	10.8	9.2
Total	50.1	49.7	49.5	44.0	66.2	43.8	42.5	41.5	42.1	44.9	42.2	46.6	45.5	51.5	49.4	57.7
Est. Time of Construction, yr	1.5	2.0	2.5	4.0	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	2.5
Est. Availability Date	1985	1985	1990	1990	1990	1990	1990	1990	1990	1990	1990	1990	1990	1990	1990	1990

Notes:

- 1 Where Applicable
- 2 Use Base Delivered Fuel Cost

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into its three components -- capital, fuel, and O&M -- are shown in Table 13.9 for the values of the construction and operation variables of Table 13.8.

The COE for AC1, 13.9 mills/MJ (50.1 mills/kWh) has as its major component the fuel charge, which comprises 50% of the total cost. This is understandable in terms of the low efficiency of this power plant and the costliness of the fuel gas at \$2.60/10⁶ Btu. The importance of the cost of high-Btu gas and its effect on the cost of electricity are shown in Figure 13.12. At \$1.50/10⁶ Btu, the electricity cost is 11.0 mills/MJ (39.5 mills/kWh); and at \$4.00/10⁶ Btu, the cost is 17.7 mills/MJ (63.5 mills/kWh).

A breakdown of the total capitalization required for this plant is provided in Tables A 13.4.1 through A 13.4.3. A graphic display of this breakdown is shown as Figure 13.16 in Section 13.6. The costs of the installed fuel cell, power-conditioning, and fuel-processing subsystems are \$169, \$62, and \$38/kWe of ac output, respectively. The balance of plant cost is \$67/kWe, and the indirect costs amount to \$112/kWe. The total capitalization is \$448/kWe.

In the standard case, defined by the values of Table 13.8, that portion of the COE ascribable to capital is 3.94 mills/MJ (14.2 mills/kWh). Because of the short time (one and a half years) required for plant construction, changes in the escalation rate and the interest rate during construction have very little effect on the cost of electricity. Similarly, because of the modular nature of the fuel cell and power-conditioning subsystems, which minimizes the site labor required, even a doubling of the labor rate results in a relatively small increase (5%) in the electricity cost (Table A 13.4.2). Changes in the fixed charge rate from the base value of 18%, however, have a more pronounced effect -- e.g., a lowering of the rate to 10% results in a reduction in the COE of 1.75 mills/MJ (6.30 mills/kWh).

Of the 3.03 mills/MJ (10.9 mills/kWh) O&M charge calculated for the power plant corresponding to Point AC1, only 0.388 mills/MJ

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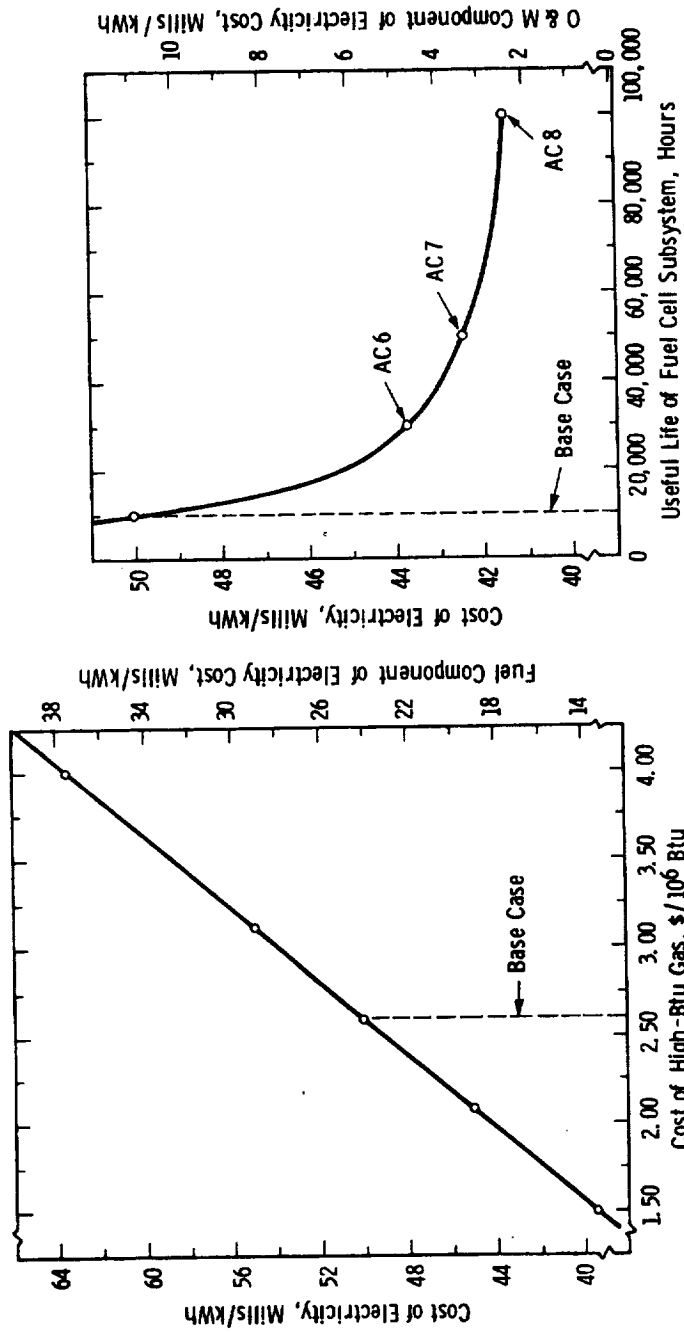


Fig. 13.12 - Dependence of the cost of electricity on the fuel cost and the useful life of a 25 MW dc phosphoric acid fuel cell power system

(1.4 mills/kWh) is attributable to labor and to material and component replacement in all parts of the plant, excluding the fuel cell subsystem. The balance, amounting to 87% of the total, is due to the need for fuel cell replacement after 36 Ms (10,000 hr) of operation. The hyperbolic relationship between the fuel cell useful life and the COE (and its O&M component) is evident from Figure 13.12, which is plotted from data generated for Points AC6, AC7, and AC8, as well as AC1. Lifetimes of 144 Ms (40,000 hr) and 360 Ms (100,000 hr) lead to reductions in the base case COE of 1.97 mills/MJ (7.1 mills/kWh) and 2.36 mills/MJ (8.5 mills/kWh), respectively.

The effect of reducing the platinum loadings in both cell electrodes from 1.0 mg/cm^2 (0.002 lb/ft^2) in the base case to 0.3 mg/cm^2 ($6.1 \times 10^{-4} \text{ lb/ft}^2$) and 0.1 mg/cm^2 ($2 \times 10^{-4} \text{ lb/ft}^2$) is explored in Points AC12 and AC13. These results are also included in Table 13.9, and are shown graphically in Figure 13.13. The order of magnitude reduction in electrode platinum loadings caused a reduction of only 1.28 mills/MJ (4.6 mills/kWh) in the cost of electricity. This result is surprising in view of the widely accepted tenet that it is desirable, from an economic standpoint, to avoid platinum as an electrocatalyst. Figure 13.13 indicates that, provided a 0.56 mills/MJ (2 mill/kWh) penalty can be absorbed and an adequate supply of platinum is available, there is little point to efforts to reduce the electrode platinum loadings much below 0.4 mg/cm^2 ($8.2 \times 10^{-4} \text{ lb/ft}^2$), corresponding to a platinum usage of approximately 6 g/kW (13.2 lb/MW). This conclusion is supported by Abens, Baker, DiPasquale, and Michalko (Reference 13.29), who stated in a recent paper that "much obfuscation of cell costs has been caused by belaboring the advances made in catalyst cost reductions."

The power density, the product of the electrode current density and the cell voltage, has a marked effect on the COE, as shown by the data for AC10 and AC11 in Table 13.9. These results are also presented in Figure 13.13. A doubling of the base-case power density results in a 2.2 mills/MJ (7.9 mills/kWh) reduction in the COE. This

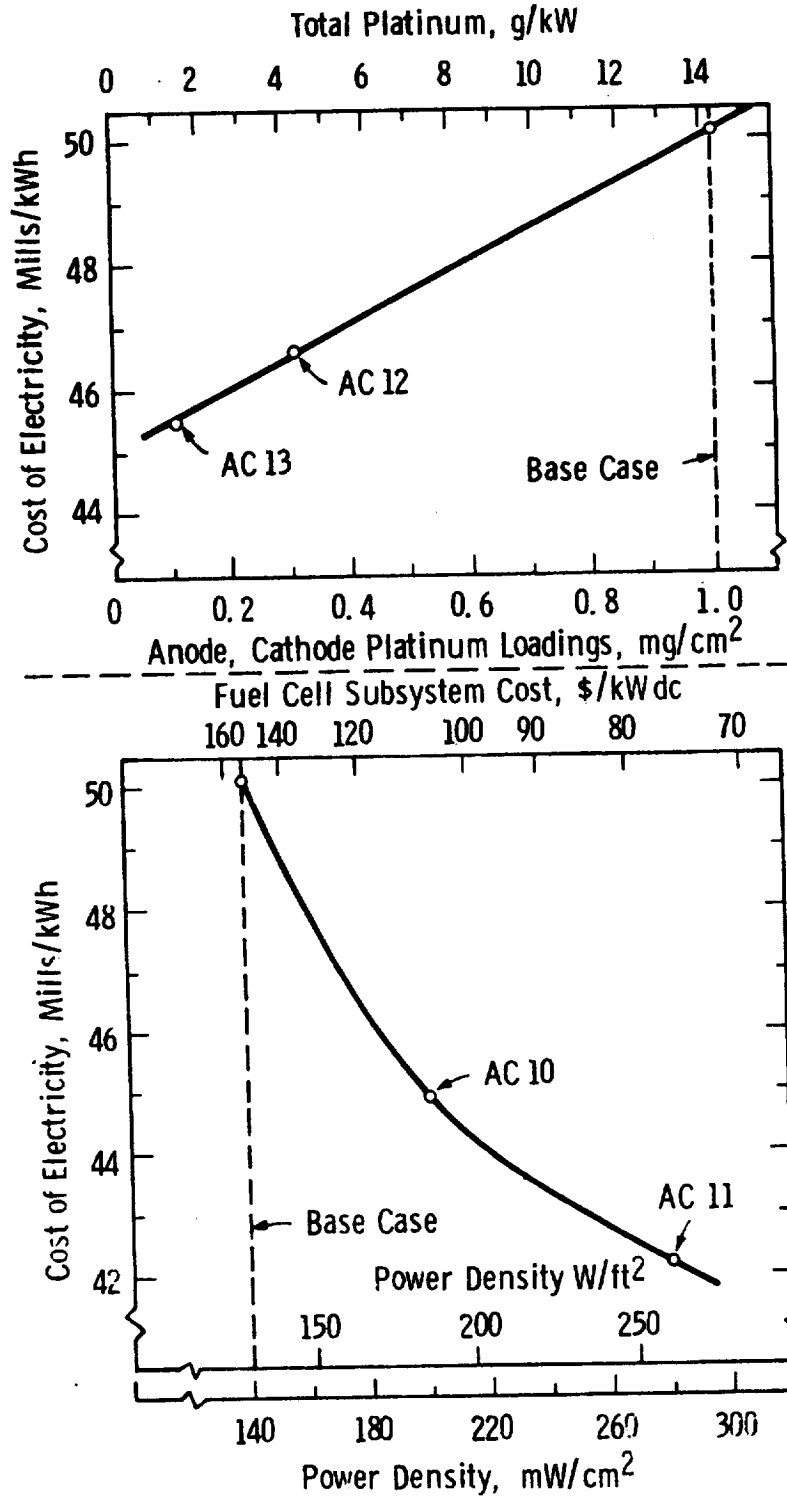


Fig. 13. 13—Dependence of the cost of electricity for phosphoric acid fuel cell power systems on the platinum loadings in the anode and cathode, and on the power density per unit cell area

graph may be used also to compute the effect of reductions in the selling price of the fuel cell subsystem because of the inverse relationship between the selling price and the power density.

Because of the modular nature of the fuel cell and power-conditioning subsystems, there is little, if any, economy of scale in fuel cell power plants. This is illustrated by the data presented in Table 13.9 for 100 MW (AC2) and 250 MW dc (AC3) power plants. Medium-Btu gas is used in the 900 MW dc power plant (corresponding to Point AC4), and because of its relatively lower cost (see Table 13.8), the fuel component of the COE is reduced to 5.44 mills/MJ (19.6 mills/kWh). In contrast, substitution of methanol for high-Btu gas, as in Point AC14, results in an increase in the COE despite a minimal increase in the power plant efficiency. This was caused by the cost of facilities for storing methanol at the power plant, and by the slightly higher charge for this fuel (Table 13.8). The COE is only minimally affected by an electrolyte-thickness reduction (Point AC15). An increase in the versatility of the power-conditioning subsystem to permit handling the lower current or voltage input expected as a result of fuel cell performance deterioration with time (Point AC9) has only a minor effect on the COE.

Oxygen is substituted for air as the oxidant in the fuel cell subsystems of the 25 MW and 250 MW dc power plants corresponding to Points AC5 and AC16, respectively. A comparison of the COE for AC1 and AC5 reveals a 4.5 mill/MJ (16 mill/kWh) penalty for the plant employing oxygen. A further comparison of the data for 250 MW dc power plants (Points AC3 and AC16) indicates that economy of scale in oxygen plants reduces this differential to 2.3 mills/MJ (8.3 mills/kWh). These cost penalties and the efficiency reduction noted above combine to make very unattractive the replacement of air by oxygen from a dedicated oxygen plant.

TABLE 13. 10--VALUES OF ALL RELEVANT PARAMETERS FOR THE PARAMETRIC POINTS OF THE AQUEOUS-ALKALINE FUEL CELL POWER SYSTEM

Parametric Point, AL#	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
Power Output, MWe	22	87	218	770	18.8	22	22	22	22	22	22	22	22	22	22	22
Fuel Cell Rating, MW	25	100	250	900	25	25	25	25	25	25	25	25	25	25	25	25
Fuel																
High-Btu Gas	X	X	X		X	X	X	X	X	X	X	X	X	X	X	X
Medium-Btu Gas				X												
Oxidant																
Air	X	X	X	X		X	X	X	X	X	X	X	X	X	X	X
Oxygen					X											
Fuel Cell Life, 10 ³ hr	10	10	10	10	10	30	50	100	100	10	10	10	10	10	10	10
Voltage Degradation, %	5	5	5	5	5	5	5	5	5	15	5	5	5	5	5	5
Temperature, °C	70	70	70	70	70	70	70	70	70	70	70	70	70	70	70	70
Electrolyte																
Type: 30 w/o KOH	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X
Thickness, cm	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05
Anode																
Type: Pt/C	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X
Type: Raney Ni																
Catalyst Loading: mg Pt/cm ²	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	0.1	0.01	0.01	1.0	1.0
Cathode																
Type: Ag/C	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X
Type: Raney Ni																
Catalyst Loading: mg Ag/cm ²	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0	1.0	5.0
Current Density, mA/cm ²	100	100	100	100	200	100	100	100	100	175	250	100	100	100	100	100
Average Cell Voltage, Volts	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8
Thermodynamic Eff. % (1)																
Powerplant Eff. %	38.1	37.7	37.7	36.6	31.8	38.1	38.1	38.1	38.1	38.1	38.1	38.1	38.1	38.1	38.1	38.3
Overall Eff. %	25.6	25.4	25.4	30.7	21.4	25.6	25.6	25.6	25.6	25.6	25.6	25.6	25.6	25.6	25.6	25.8
Total Capital Cost × 10 ⁻⁶ , \$	13.6	51.9	130	53.6	16.6	13.6	13.6	13.6	13.9	10.9	9.8	11.6	11.3	13.6	13.5	13.6
Capital Costs, \$/kWe	620	599	599	700	890	620	620	620	636	500	448	529	517	620	615	620
Cost of Elect, Mills/kWh																
Capital	19.6	18.9	18.9	22.1	28.1	19.6	19.6	19.6	20.1	15.8	14.2	16.7	16.3	19.6	19.4	19.6
Fuel (2)	23.3	23.6	23.6	18.7	27.9	23.3	23.3	23.3	23.3	23.3	23.3	23.3	23.3	23.3	23.3	23.1
Oper. & Maint.	17.8	17.9	17.9	18.1	11.2	7.2	5.0	3.4	3.6	11.4	8.7	16.5	16.5	17.8	17.4	17.8
Total	60.8	60.4	60.4	58.9	67.2	50.1	48.0	46.4	47.0	50.6	46.2	56.5	56.1	60.8	60.1	60.6
Est. Time of Construction, yr	1.5	2.0	2.5	4.0	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5
Est. Availability Date	1985	1985	1990	1990	1985	1985	1985	1985	1985	1985	1985	1985	1985	1985	1985	1985

Notes:

- ① Where Applicable
- ② Used Base Delivered Fuel Cost

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13.5.2 Alkaline Fuel Cell Power System

The results of the parametric assessment for the alkaline fuel cell power system are shown in Table 13.10. In general, the power plant and overall energy efficiencies are higher than those for the acid power system. The margin is not as great as expected with the higher cell voltages — 0.8 V vs 0.7 V — in this system. The lower margin is due to the parasitic losses associated with operation of the Rectisol units for scrubbing the carbon dioxide from the fuel gas and air streams fed to the power system. The capitalization required for alkaline power plants is greater because of the lower power density and, hence, higher costs of the fuel cell subsystem, and because of the costs of the Rectisol units, which are not required in the acid system. Similarly, the O&M charges are higher because of the greater replacement costs of the fuel cell subsystem. The net effect is that the COE for the alkaline fuel cell power systems are approximately 20% higher than those for the phosphoric acid system.

With high-Btu gas as fuel and air as oxidant, the power plant and overall energy efficiencies lie in general at 38% and 25 to 26%, respectively. When medium-Btu gas is employed as the fuel (AL4), the overall energy efficiency increases to 31%, while the power plant efficiency, 37%, is slightly lower. As discussed earlier for the acid system, the substitution of oxygen for air as the oxidant (AL5) results in a substantial lowering of both efficiency values. The power plant and overall energy efficiencies move sharply downward to 32% and 21%, respectively.

A detailed breakdown of all of the direct costs for the base case, AL1, is provided by Tables A 13.4.4 through A 13.4.6. A simplified breakdown is shown later in Figure 13.16 of Section 13.6. It should be noted that the Rectisol unit for air scrubbing is included in the fuel processing cost in this diagram. Table A 13.4.2 shows the effects of variation of the labor rate, contingency charge, escalation rate, fixed charge rate, interest rate during construction, fuel cost, and capacity

factor on the capital costs and COE for the base case power plant. The cost of electricity breakdown for the values of the construction and operation variables listed in Table 13.8 is as follows: 5.44 mills/MJ (19.6 mills/kWh) for capital, 6.48 mills/MJ (23.3 mills/kWh) for fuel, and 4.95 mills/MJ (17.8 mills/kWh) for O&M. The total COE is 16.9 mills/MJ (60.8 mills/kWh).

As for the acid power system, the high O&M charge is mostly due to the high replacement cost of the fuel cell subsystem. Of the total, 4.4 mills/MJ (16 mills/kWh) is ascribable to the need for total replacement after 36 Ms (10,000 hr) of operation. The data for AL6, AL7, and AL8, shown in Table 13.10, indicate the substantial effect of fuel cell useful life on the COE. Increasing the useful life to 180 Ms (50,000 hr) reduces the COE by 4.0 mills/MJ (14.4 mills/kWh).

An equally profound effect is shown by comparing the data for AL10 and AL11 with those for the base case AL1. A power density increase of 150% results in lowering the COE by 4.05 mills/MJ (14.6 mills/kWh). A lowering of catalyst loadings, however, as in AC12, AC13, and AC15, leads to less significant reductions of the COE, just as in the case of the acid system.

No economy of scale is observed. The COE for the 100 MW dc (AL2) and 250 MW dc (AL3) power systems differs very little from those for the 25 MW dc base case. The full benefit of the use of the cheaper medium-Btu gas is not realized in the 900 MW dc plant (AL4) because of the additional Rectisol process costs associated with the need to eliminate additional carbon dioxide, which now forms 40% of the shift converter effluent instead of the 20% when high-Btu gas is employed as the fuel.

As for the corresponding case in the acid fuel cell parametric assessment, there is COE penalty associated with the replacement of air by oxygen (AL5) from a dedicated liquid-air distillation plant. The penalty of the alkaline system is less than half that for the acid system, however, because acid gas scrubbing of the oxidant stream is no longer

necessary. The effects of electrolyte-thickness reduction (AL16) and increasing the flexibility of the power conditioning subsystem (AL9) are small.

13.5.3 Molten Carbonate Fuel Cell Power System

Table 13.11 is a summary of initial plant and overall efficiencies for the various parametric points. The efficiencies do not represent optimized systems, but allow a comparison of the relative effect of different parameters. The efficiency of the fuel cell for any one fuel is assumed to be the same, with current density being changed by changes in operating variables such as the use of oxygen instead of air or alteration of the electrolyte thickness. Consequently, all systems using high-Btu gas have nearly the same plant and overall efficiencies of 48 to 49% and 32 to 33%, respectively. The small differences are due to different efficiencies of the power-conditioning system for different plant size, as described in Appendix A 13.2.

There will be an efficiency decrease with time from these initial values, depending on the amount of voltage degradation of the cell at constant power.

The fuel type has an important effect on the efficiency. Thus, plant efficiencies of about 36.5% and 45% are obtained for medium-Btu and methanol fuels, respectively, compared to about 49% for high-Btu gas. On the other hand, the overall efficiency of about 30.5% for medium-Btu gas is not much less than the 33 and 31% efficiencies for high-Btu and methanol fuels. This is due to the higher gasification efficiency of the medium-Btu gas.

The use of medium-Btu gas involves much larger heat losses in the cell due to entropy factors in the electrochemical reactions. If we recover this heat and use it to produce additional ac power via a steam turbine generator as in MC4, we increase the power plant efficiency beyond that of the other systems which do not use the waste heat in this way. If we used the analogous combined fuel cell/turbine generator system with high-Btu or methanol fuels we would obtain a plant efficiency of about

TABLE 13.11—VALUES OF ALL RELEVANT PARAMETERS FOR THE PARAMETRIC POINTS OF THE MOLTEN-CARBONATE FUEL CELL POWER SYSTEM

Parametric Point, MC#	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
Power Output, MWe	22.6	88.8	22.2	1190	22.5	22.6	22.6	22.6	22.6	22.6	22.6	22.6	22.6	22.2	22.5	22.6	223.8
Fuel Cell Rating, MWdc	25	100	250	900	25	25	25	25	25	25	25	25	25	25	25	25	250
Fuel																	
High-Btu Gas	X	X	X		X	X	X	X	X	X	X	X	X			X	X
Medium-Btu Gas				X										X			
Methanol															X		
Oxidant																	
Air	X	X	X	X		X	X	X	X	X	X	X	X	X	X	X	X
Oxygen					X												X
Fuel Cell Life, 10 ³ hr	10	10	10	10	10	30	50	100	100	10	10	10	10	10	10	10	10
Voltage Degradation, %	5	5	5	5	5	5	5	5	15	5	5	5	5	5	5	5	5
Temperature, °C	650	650	650	650	650	650	650	650	650	650	650	700	750	650	650	650	650
Electrolyte Type																	
Paste of Li, Na, K, Carbonates and Alkali Aluminates	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X
Thickness, cm	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1
Anode Type																	
Ni	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X
Cathode Type																	
Lithiated NiO	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X
Current Density, Ma/cm ²	200	200	200	200	300	200	200	200	200	150	250	201	202	200	200	203	300
Average Cell Voltage, Volts	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Thermodynamic Eff, % (1)																	
Powerplant Eff, %	48.8	48.0	48.0	54.4	48.8	48.8	48.8	48.8	48.8	48.8	48.8	48.8	48.8	36.4	44.8	48.8	48.0
Overall Eff, %	32.9	32.3	32.3	45.7	32.9	32.9	32.9	32.9	32.9	32.9	32.9	32.9	32.9	30.6	31.2	32.9	32.3
Total Capital Cost × 10 ⁻⁶ , \$	11.5	45.0	114	570	19.9	11.5	11.5	11.5	11.9	14.6	10.7	11.5	11.5	11.9	11.8	11.4	150
Capital Costs, \$/kWe	514	505	511	482	888	514	514	514	530	653	477	513	511	540	528	510	674
Cost of Elect, Mills/kWh																	
Capital	16.3	16.0	16.2	15.2	28.1	16.3	16.3	16.3	16.7	20.7	15.1	16.2	16.2	17.1	16.7	16.1	21.3
Fuel (2)	18.2	18.5	18.5	12.6	18.2	18.2	18.2	18.2	18.2	18.2	18.2	18.2	18.2	18.8	20.6	18.2	18.5
Oper. & Maint.	23.4	23.5	23.6	16.1	20.8	8.6	5.7	3.5	3.6	30.7	19.0	23.2	23.1	23.9	23.5	23.1	18.9
Total	57.8	57.9	58.2	43.9	67.0	43.1	40.1	37.9	38.5	69.6	52.2	57.7	57.4	59.7	60.8	57.4	58.7
Est. Time of Construction, yr	1.5	2.0	2.5	5.0	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	2.5
Est. Availability Date	1990 +	1990 +	1990 +	1990 +	1990 +	1990 +	1990 +	1990 +	1990 +	1990 +	1990 +	1990 +	1990 +	1990 +	1990 +	1990 +	1990 +

Notes:

- ① Where Applicable
- ② Used Base Delivered Fuel Cost

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59% but a lower overall efficiency (40 to 41%) than the 45.5% for MC4 (medium-Btu fuel). This is due to the lower gasification efficiency for the higher-grade fuels.

MC15 uses methanol as fuel. The ~45% power plant efficiency calculated for this system is only 4% less than the value for the corresponding system using high-Btu gas (MC1). Internal reforming of methanol, however, will probably be easier than for high-Btu gas. This might reduce or eliminate efficiency differences between the high-Btu (MC1) and methanol (MC15) systems.

The systems using oxygen, MC5 and MC17, are arbitrarily operated at the same efficiency as MC1 but at higher current densities to reduce fuel cell capital and O&M costs, as explained below. Note that for these systems, power for the air compressor of the oxygen plant is derived from a steam turbine operating with free excess heat from the fuel cell. If an electric drive compressor were used, it would severely reduce the plant efficiency.

In summary, the best overall efficiency of about 45% for the molten carbonate system investigated is attained for a combined system with the fuel cell using medium-Btu fuel and a steam turbine generator to convert excess heat from the fuel cell to electrical energy.

The highest plant efficiencies of about 59% can be obtained for the analogous combined systems using high-Btu or methanol with internal reforming, but their overall efficiencies are only 40 to 41%, due to the inefficiency of the gasification process.

For the fuel cell system only, plant efficiencies of ~49, ~45, and ~36%; and overall efficiencies of ~33, ~31, and ~30.5% are obtained with the high-Btu, methanol and medium-Btu fuels, with little to choose between high-Btu and methanol since internal reforming may be easier with the latter fuel.

The cost input to the computer program and the output data for the power plant corresponding to MC1, the base case, are listed in

Tables A 13.4.7 through A 13.4.9. The cost estimates for COE from the various unoptimized molten carbonate fuel cell plants selected for analysis vary from about 10.8 to 19.4 mills/MJ (39 to 70 mills/kWh), as shown in Table 13.11. If a ten-year fuel cell lifetime with excess heat recovery were possible, a COE value close to 8.3 mills/MJ (30 mills/kWh) could perhaps be realized.

Of the parameters investigated, power plant size has only a small effect of about 0.8% reduction of COE for a factor of ten increase in plant size, going from MC1 (25 MW) to MC3 (250 MW). If oxygen rather than air is used as oxidant, however, plant size is important, since the cost of oxygen varies substantially with the amount of oxygen required. Thus, MC5, the 25 MW plant, has a COE of about 18.6 mills/MJ (67 mills/kWh) compared to 16.3 mills/MJ (58.7 mills/kWh) for the corresponding 250 MW plant (MC17) using oxygen. The latter figure, however, is no improvement over the 25 MW plant using air and assumes the availability of free waste heat from the fuel cell to operate the turbine-drive compressor of the oxygen plant.

The effect of fuel type is surprisingly small, with medium-Btu gas and methanol for the 25 MW plants (MC14 and MC15) having a COE only about 3 to 5% greater, respectively, than for the corresponding plant using high-Btu gas (MC1) and no conversion of waste heat. If the waste heat from the fuel cell, however, is converted to ac electrical energy via a steam turbine generator, as in MC4, the COE is better for medium-Btu than for either high-Btu or methanol fuels, since the overall efficiency is better and capital and O&M costs are about the same.

If the power density can be increased without decreasing the efficiency by any means, this has a strong effect on the COE. This is illustrated by MC10 and 11, in which the power density is reduced by 25% and increased by 25%, respectively, from that of the Base Case, MC1. This results in about a 20% increase and about a 9% decrease, respectively, in the COE.

MC16 shows that the effect of reducing the electrolyte thickness a factor of 2 from an already thin 1 mm (39 mils) is small, only about 0.7% less than the COE for MC1.

MC6, 7, and 8 show that the effect of cell lifetime is the most important factor in reducing the COE. A change from a 36 Ms (10,000 hr) cell lifetime 108, 180, and 360 Ms (30,000, 50,000, and 100,000 hr) gave corresponding reductions of the COE of about 25, 30, and 34%, respectively.

In summary, fuel cell life longer than the presently possible 36 Ms (10,000 hr) production of additional ac energy from excess fuel cell heat, or a substantial increase in power output over that assumed, are the principal factors capable of appreciably reducing the COE below about 16.7 mills/MJ (60 mills/kWh) estimated from available state-of-the-art developments to date for the molten carbonate fuel cell.

Fuel cell size and fuel type (except for combined fuel cell and steam turbine generator systems), electrolyte thickness, and moderate voltage degradation have relatively small effects on the COE.

Oxygen is not competitive with air as the oxidant except possibly in the extreme case of a large (e.g., 250 MW) fuel cell plant, with waste heat being used to run a turbine which operates the air compressor of the oxygen plant.

13.5.4 Solid Electrolyte Fuel Cell Power System

The power plant and overall energy efficiencies for all twenty points of the parametric assessment of the solid electrolyte power system are listed in Table 13.12. This table also provides information pertinent to the operation of the fuel cell subsystem in the power plant corresponding to each of the twenty points.

In the Base Case SE1 high-Btu gas is employed as the fuel. Because of the high average cell voltages, predicted on the direct electrochemical oxidation of methane at the fuel electrode, the power plant and overall energy efficiencies are high, lying at 69.7 and 46.9%, respectively. These values are unlikely to be realized in practice

TABLE 13.12 - VALUES OF ALL RELEVANT PARAMETERS FOR THE PARAMETRIC POINTS OF THE SOLID-ELECTROLYTE FUEL CELL POWER SYSTEM

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20		
Parametric Point, SE #	22.9	87.6	219	1164	18.3	22.3	22.3	22.3	22.3	22.3	22.3	22.3	22.3	22.3	22.3	22.3	22.3	22.3	219	1064	22.7	
Power Output, MW	25	100	250	900	25	25	25	25	25	25	25	25	25	25	25	25	25	25	250	900	25	
Fuel Cell Rating, MW																						
Fuel	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X
High-Btu Gas																						
Medium-Btu Gas																						
Low Btu Gas																						
Oxidant																						
Air	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X
Oxygen																						
Fuel Cell Life, 10 ³ hr	10	10	10	10	30	50	100	100	100	10	10	10	10	10	10	10	10	10	10	10	10	10
Voltage Degradation, %	5	5	5	5	5	5	5	5	5	5	5	5	5	5	5	5	5	5	5	5	5	5
Temperature, °C	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000
Electrolyte																						
(Zr O ₂) 1-x (Ca O) x	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X
(Zr O ₂) 1-x (Y ₂ O ₃) x	4	4	4	4	4	4	4	4	4	4	4	4	4	4	4	4	4	4	4	4	4	4
Thickness, cm 10 ⁻⁵																						
Anode	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X
Ni-Zr O ₂ - Cermet																						
Cathode	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X
In ₂ O ₃ /Pr Co O ₃ -x																						
Sn O ₂ Sb-doped																						
Interconnection																						
Co Cr ₂ O ₂ (Mn-doped)																						
Cr ₂ O ₃	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X
Thickness, cm 10 ⁻³	2	2	2	2	2	2	2	2	2	2	2	2	2	2	2	2	2	2	2	2	2	2
Current Density, mA/cm ²	400	400	400	400	400	400	400	400	400	400	400	400	400	400	400	400	400	400	400	400	400	400
Average Cell Voltage, Volts	0.84	0.66	0.66	0.66	0.76	0.66	0.66	0.66	0.66	0.59	0.51	0.67	0.66	0.64	0.66	0.55	0.77	0.68	0.56	0.96	0.69	0.69
Thermodynamic Eff. % (1)	65.7	40.9	40.9	60.2	39.7	41.6	41.6	41.6	41.6	37.1	32.0	42.3	41.6	40.3	41.6	34.7	48.5	53.0	47.8	56.7	56.7	56.7
Powerplant Eff. %	46.9	34.3	34.3	50.6	33.3	35.0	35.0	35.0	35.0	31.2	26.9	35.5	35.0	33.9	35.0	29.2	40.8	53.0	47.8	38.2	38.2	38.2
Overall Eff. %	9.67	42.0	106	539	17.7	10.7	10.7	10.7	11.0	9.63	9.28	10.5	10.7	10.7	9.04	11.9	9.80	205	893	7.67	7.67	7.67
Total Capital Cost x 10 ⁻⁶ \$	424	482	488	466	972	481	481	481	496	434	418	474	481	483	407	535	442	948	859	340	340	340
Capital Costs, \$/rWe																						
Cost of Elect. Mills/kWh	13.4	15.2	15.4	14.7	30.7	15.2	15.2	15.2	15.7	13.7	13.2	15.0	15.2	15.3	12.9	16.9	14.0	30.0	27.1	10.7	10.7	10.7
Capital	12.7	16.7	16.7	11.3	17.2	16.4	16.4	16.4	16.4	18.4	21.3	16.1	16.4	16.9	16.4	19.7	14.1	5.5	6.3	15.7	15.7	15.7
Fuel (2)	16.0	20.8	21.1	14.2	24.2	7.9	5.3	3.4	3.5	16.2	14.5	20.2	20.6	21.1	20.6	24.3	17.9	12.2	18.4	10.9	10.9	10.9
Oper. & Maint.	42.1	52.7	53.2	40.2	72.2	39.5	36.9	35.0	35.6	48.3	49.0	51.3	52.2	53.3	49.9	60.9	45.9	47.7	51.9	37.3	37.3	37.3
Total	1.5	2.0	2.5	5.0	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5
Est. Time of Construction, yr	2000+	2000+	2000+	2000+	2000+	2000+	2000+	2000+	2000+	2000+	2000+	2000+	2000+	2000+	2000+	2000+	2000+	2000+	2000+	2000+	2000+	2000+
Est. Availability Date																						

Notes:

- ① Where Applicable
- ② Use Base Delivered Fuel Cost

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because of the metastability of methane at this temperature, resulting in either reformation on any available metallic surface, including that of the metallic component of the fuel electrode, in the presence of an adequate supply of water vapor, or carbon deposition (Reference 13.46).

For most points in the parametric assessment, medium-Btu gas is the fuel, so that for comparison purposes, Point SE2, corresponding to the 100 MW dc power plant, is much more useful than Point SE1. With medium-Btu gas as fuel, the power plant efficiency lies in the range 40 to 42% (SE2, 3, 5 through 9, 12 through 15), while the overall energy efficiency moves between 33 and 36%. Because of the lowered cell voltages in the fuel cell subsystems, corresponding to Points SE10 and SE11, increases in the power density cause a reduction in the plant and overall efficiencies.

The heat which must be rejected by the fuel cell subsystem results from thermodynamic and electrochemical inefficiencies. This waste heat may be employed in a steam-bottoming plant (SE4) or to supply the heat required by the endothermic processes occurring in a coal gasification reactor, which then meets the fuel requirements of the fuel cell subsystem (SE18 and SE19). The net effect of waste heat recovery is to substantially raise the efficiency of the power system. For Point SE4, the power plant and overall energy efficiencies are 60.2 and 50.6%, respectively. As the fuel in the power plant corresponding to Point SE18, the Westinghouse Fuel Cell Power System, is coal, the power plant and overall energy efficiencies are identical at 53%. Point SE19, which involves the use of a low-Btu gasification reactor, employing coal, air, and steam as input, the power plant efficiency is 47.8% because of the use of a steam-bottoming plant. This is lower than for Point SE18, in which medium-Btu is generated by recycling fuel gas which has been partially oxidized in the fuel cell subsystem. The overall efficiency for Point SE19 is identical with the power plant efficiency because the low-Btu gasifier is considered to be fully integrated with the power system. Thus, coal and air are the inputs.

The computer outputs for Points SE1 and SE2 is shown in Tables A 13.4.10 through A 13.4.15. The COE and their breakdowns into capital, fuel, and O&M charges, for the values of plant construction and operation shown in Table 13.8, are given in Table 13.12.

As discussed above, the results for Point SE2 provides a better basis for discussion of the parametric assessment of the solid electrolyte power system because of the general use of medium-Btu gas for most points, and also because of the technical uncertainty surrounding the direct use of high-Btu gas as a fuel. The COE for the power plant corresponding to SE2 is 14.6 mills/MJ (52.7 mills/kWh). The portion ascribable to capital is 4.23 mills/MJ (15.2 mills/kWh), to fuel 4.64 mills/MJ (16.7 mills/kWh), and to O&M 5.77 mills/MJ (20.8 mills/kWh). The O&M charge is high because of the high replacement cost — unlike the platinum-laden acid fuel cell modules, the solid electrolyte modules are assumed to have a scrap value of zero. Of the total charge, 5.35 mills/MJ (19.3 mills/kWh) is ascribable to the need to replace the modules after 36 Ms (10,000 hr) of useful life. Points SE6 through 8 explore the effect of increasing the useful life. The electricity costs are 11.0 mills/MJ (39.5 mills/kWh) for 108 Ms (30,000 hr) and 9.73 mills/MJ (35.0 mills/kWh) for 300 Ms (100,000 hr).

Comparison of the COE and efficiency data for SE2, SE10, and SE11, reveal the effect of variations in the power density. An increase of the active cell power density from 264 to 354 mW/cm^2 (245 to 329 W/ft^2) results in a COE reduction from 14.6 to 13.4 mills/MJ (53.7 to 48.3 mills/kWh). A further increase in the power density to 408 mW/cm^2 (379 W/ft^2), however, caused a small increase in the COE, as the increase in the cost component ascribable to fuel (because of the lower cell voltage and thus plant efficiency) more than outweighs the sum of the reductions in the costs ascribable to capital and O&M.

The power density changes, also, when the temperature of operation is changed from 1273°K (1832°F) as in SE2 to 1373°K (2012°F) and 1173°K (1652°F) in SE17 and SE16, respectively. The decline in

electrolyte and interconnection resistivities with increasing temperature is assumed to result in a parallel increase in cell voltages. Thus, an increase in the temperature of 100°K (180°F) causes a 20% increase in cell voltage and power plant efficiency resulting in a COE decrease of approximately 1.9 mills/MJ (6.8 mills/kWh). A lowering of the fuel cell temperature by 100°K (180°F) results in COE penalty of 2.27 mills/MJ (8.2 mills/kWh).

Substitution of oxygen for air as the oxidant, as in Point SE5, results in a substantial COE penalty. The penalty is greater here than for Point MC5 because the power required for the oxygen plant is taken from the ac output of the power-conditioning subsystem instead of being generated by a turbine powered by waste heat from the fuel cell subsystem.

Replacement of tin-doped indium oxide by antimony-doped tin oxide, as in SE15, causes a small decrease in the electricity cost. Even smaller decreases are registered by reducing the electrolyte thickness (SE12) and by substituting manganese-doped cobalt chromite for chromium sesquioxide as the interconnection material (SE13). Despite the lower cost of calcia as a stabilizing agent for zirconia (relative to yttria), the use of calcia-stabilized zirconia as an electrolyte (SE14) results in a small electricity cost penalty. This is caused by the higher resistivity of this electrolyte, which, in turn, causes a cell voltage reduction and, consequently, a loss in plant efficiency.

The use of a steam bottoming plant for waste-heat recovery in the power plant corresponding to Point SE4 results in a dramatic reduction in the electricity cost from 14.6 mills/MJ (52.7 mills/kWh) to 11.2 mills/MJ (40.2 mills/kWh). Similarly, a reduction of 1.4 mills/MJ (5 mills/kWh) is observed for the Westinghouse Fuel Cell Power System (SE18). The results for SE19 indicate that partial thermal coupling of the fuel cell subsystem with a low-Btu gasifier offers little advantage from a COE standpoint, despite the use of a steam-bottoming plant for waste heat recovery.

13.6 Conclusions and Recommendations

Considerable caution must be exercised when a comparison between fuel cell power systems is attempted on the basis of data provided for individual systems in Subsection 13.5. The calculated power plant efficiencies may be relied upon to within a few percent either way, as they compare favorably with estimates available from other fuel cell work. They are, of course, dependent on the correctness of the assumptions of fuel cell subsystem performance, cited in Subsections 13.2 and 13.3.

As will be evident from the approaches taken in the costing of the fuel cell subsystems in Section 13.4, there is a much greater possibility of error in the estimation of COE. The costing procedures employed represent an unbiased effort to estimate the possible costs of the fuel cell subsystems on the basis of a realization of the performance targets, as discussed in subsection 13.2. Because they are founded on so many arbitrary assumptions, the comparison of the different fuel cell power systems based on the COE derived in this study must be approached with care.

The parametric assessment of the four cell power systems was based on a matrix of 69 points - 16 points each for the phosphoric acid and alkaline systems, 17 for molten carbonate, and 20 for solid electrolyte. The parameters of the power systems, which were varied, are listed in Table 13.13.

Table 13.13 - Parameters Varied in Fuel Cell Assessment

Fuel Cell Useful Life	Fuel Cell Plant Rating
Power Density	Electrolyte Thickness
Fuel Type	Voltage Degradation
Oxidant Type	Waste Heat Recovery System**
Catalyst Loading*	Temperature of Operation**

* Applicable for acid and alkaline systems.

** Applicable for molten carbonate and solid electrolyte systems.

For each of the 69 points, power plant and overall energy efficiencies, and COE (broken down into capital, fuel, and O&M components) were calculated. Analysis of these results indicated that four of the parameters listed in Table 13.13 were of particular importance in improving efficiency and reducing COE. The four parameters and their areas of impact are shown in Table 13.14.

Table 13.14 - Important Parameters of Fuel Cell Power Systems

Parameter	Area of Impact
Fuel Cell Useful Life	O&M costs
Fuel Cell Power Density	Capital and O&M costs
Waste Heat Recovery System	Plant and overall efficiencies
Fuel Type	Overall efficiency

The importance of fuel cell subsystems useful life is seen in Figure 13.14. The decrease in COE with increasing life is most pronounced for the molten carbonate and solid electrolyte systems. The effect is least for the phosphoric acid power system because, at \$152/kW dc, the acid fuel cell subsystem not only is the cheapest, but also has a sizeable salvage value due to its platinum content. Although the alkaline system has a similar salvage value, the low power density of the cell results in a higher replacement cost for the final cell subsystem, and, consequently, a greater dependency of the COE on the useful fuel cell life.

Figure 13.15 shows the marked effect of power density, i.e., power output per unit electrode area, on the COE for the acid, alkaline, and molten carbonate systems. Increasing power density at constant efficiency implies advances in the state of the art of cathode and anode fabrication technology. The more conventional technique of power density variation is to increase the current density, accepting a cell voltage reduction and, therefore, an efficiency penalty. This results in an

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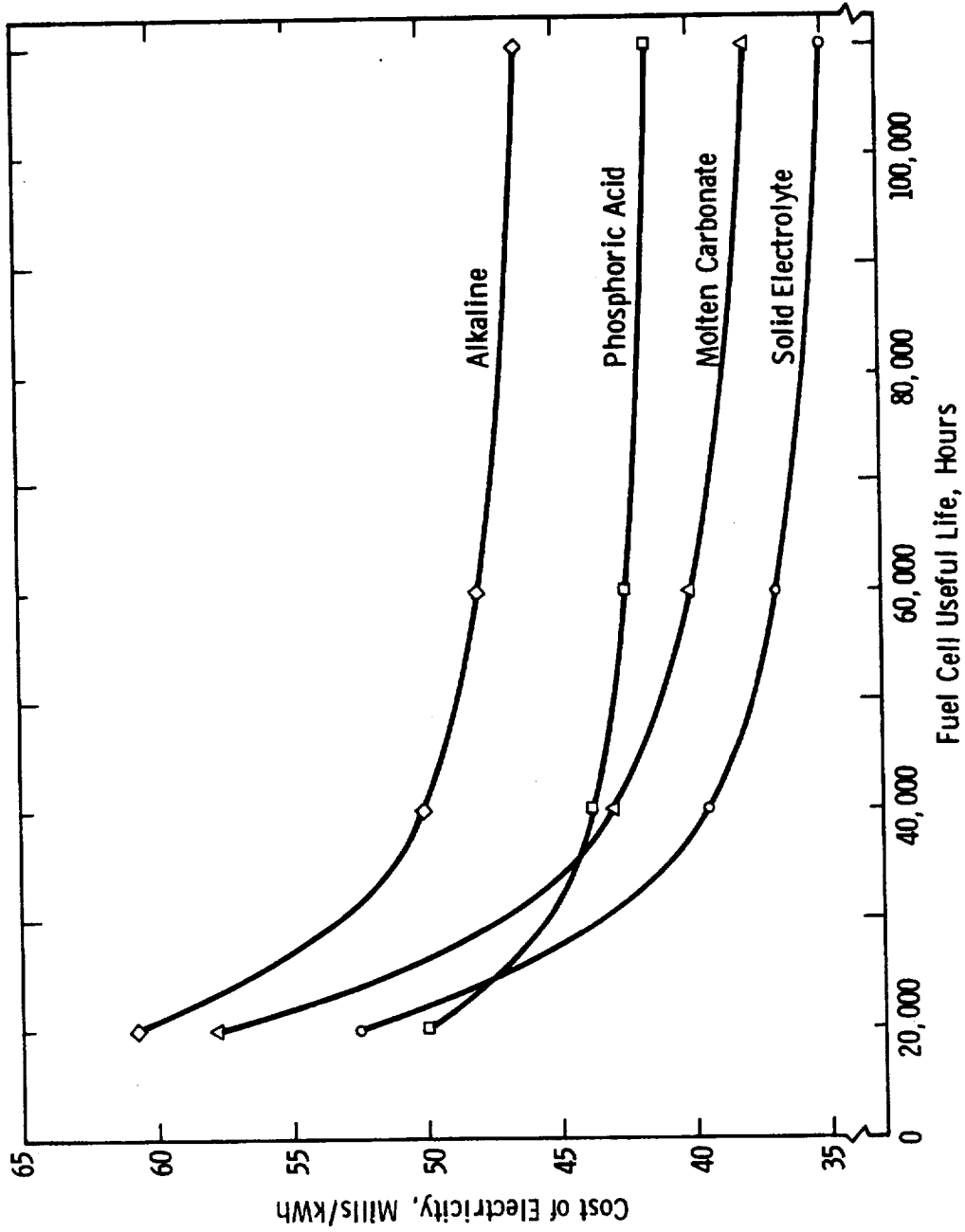


Fig. 13. 14—The effect of the useful life of the fuel cell subsystem on the cost of electricity. The fuel gas for the acid, alkaline and molten carbonate systems is high-Btu gas, while the solid electrolyte system costs were calculated on the basis of the use of medium-Btu gas

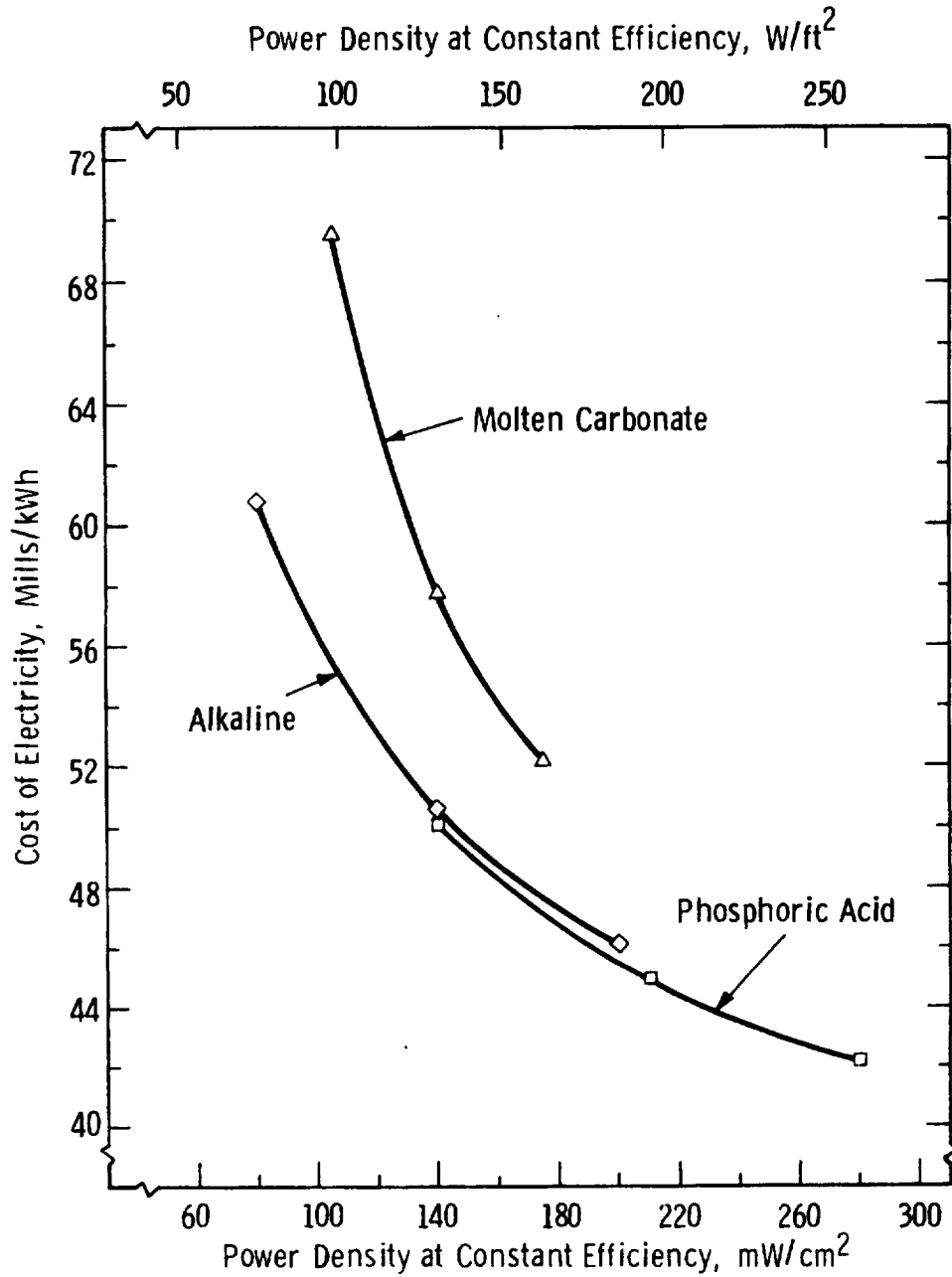


Fig. 13. 15—Dependence of the electricity cost on the power density at constant efficiency for phosphoric acid, alkaline, and molten carbonate fuel cell power systems

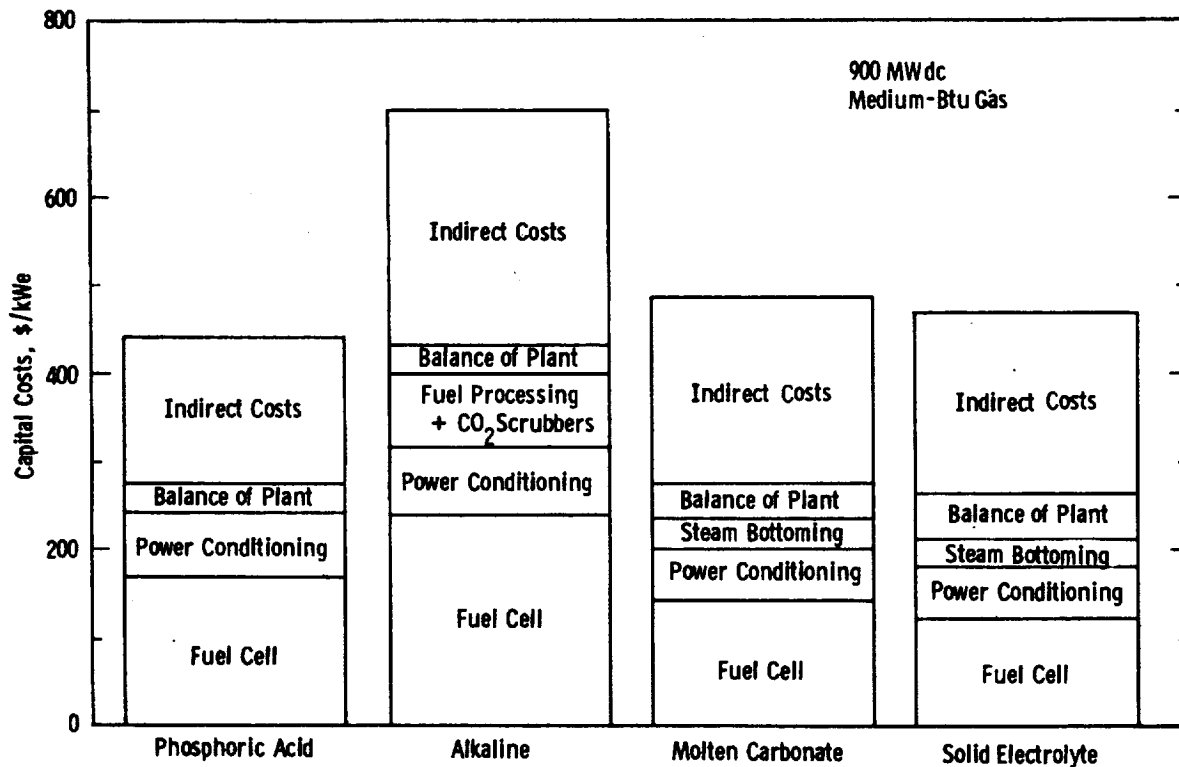
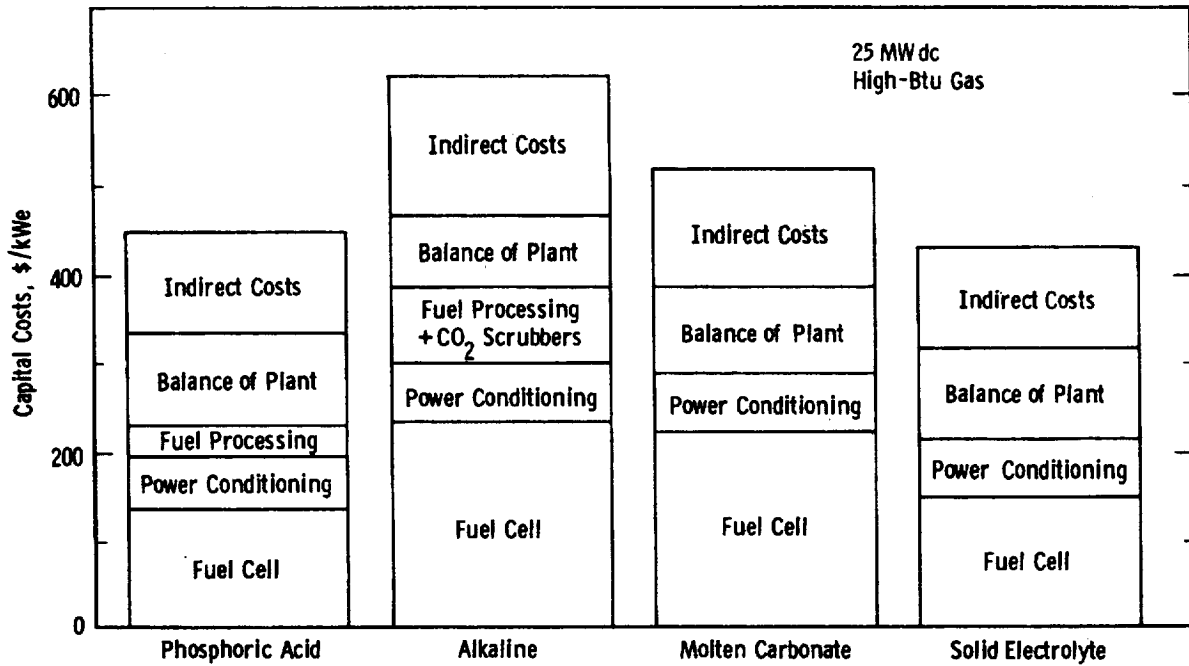


Fig. 13.16—Breakdown of capital costs for the base cases (corresponding to parametric points AC1, AL1, MC1, and SE1) and for the 900 MW fuel cell power plants (AC4, AL4, MC4, and SE4)

increased fuel charge which serves to offset the reductions in the capital and O&M components of the COE (as discussed for Points SE10 and SE11 in Section 13.5.4). A further complication of operation at a higher power density was discussed by Kordesch (Reference 13.47) for alkaline fuel cells. The useful life is inversely proportional to the current density, so that operation at a higher power density would result in more frequent replacement of the fuel cell subsystem, and thus, in a higher O&M charge.

The coupling of a steam bottoming plant to the 900 MW solid electrolyte and molten carbonate fuel cell subsystems of the power plants, corresponding to Points SE4 and MC4, raises the ac outputs to 1164 and 1170 MWe, respectively, and the overall energy efficiencies to 50.6 and 45.7%, respectively. Thermal coupling of the fuel cell subsystem with a coal gasifier (another form of waste heat recovery) results in an overall energy efficiency of 53.0%, the highest derived in this study (Point SE18).

Because of the greater efficiency of the production of medium-Btu gas, relative to high-Btu gas, a 25% gain in overall energy efficiency may be registered by the use of this fuel. The lower cost of medium-Btu gas (Table 13.8) results also in a lowering of the fuel component of the COE by greater than 20%, as shown by comparing the data for Points AC1 and AC4.

A comparison of the capital cost breakdowns for all four base cases is provided in Figure 13.16. The balance of plant was calculated by subtracting the sum of the material and site labor costs for the major components - e.g., for the phosphoric acid system, the fuel cell, power-conditioning, and fuel processing subsystems - from the total direct costs of the power plant. The indirect costs, which include the interest during construction, escalation and contingency charges; and the profit and owner costs, were calculated similarly and represent the difference between the total capitalization and total direct costs.

The dominance of the fuel cell subsystem cost, lying in the range of 35 to 42% of the total capitalization, is apparent for every power system. The indirect costs, averaging 25% of the total, are also

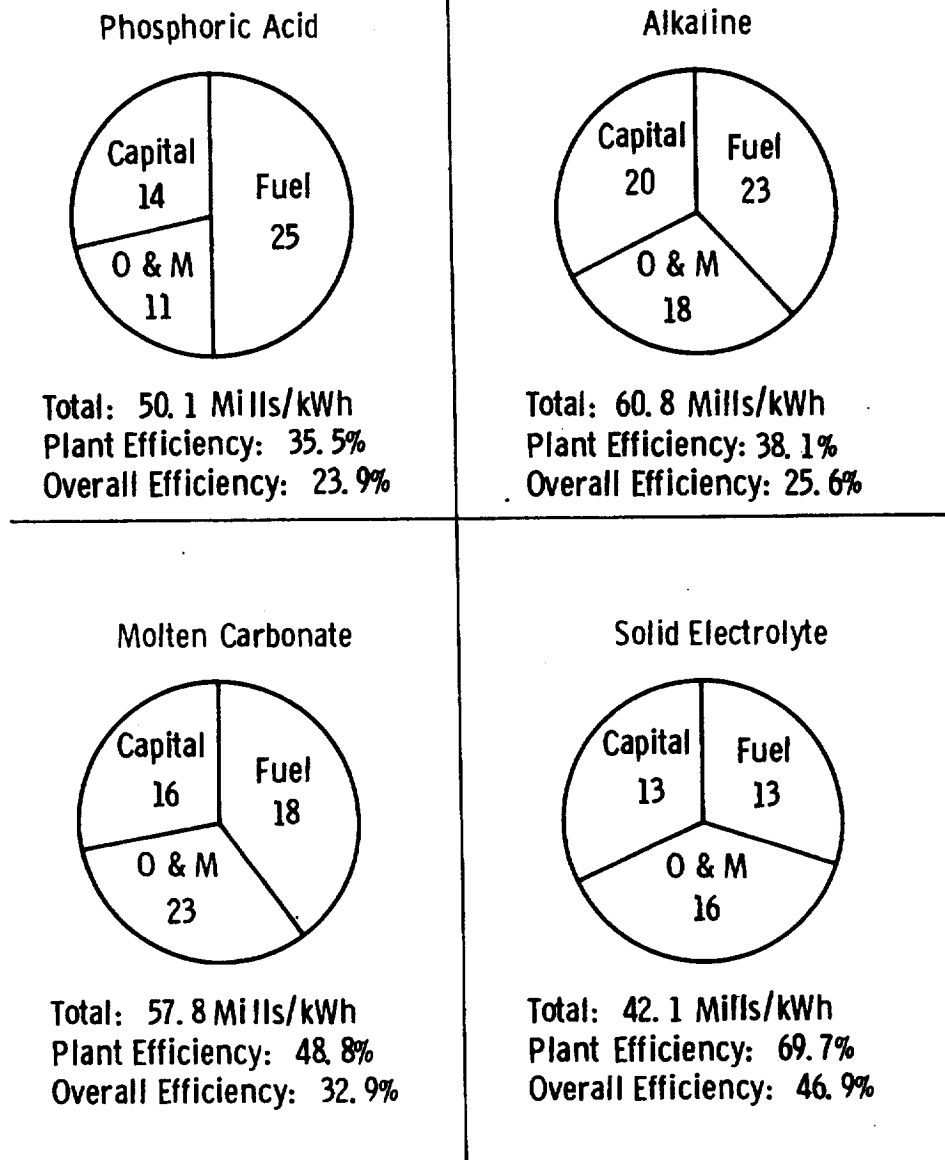


Fig. 13.17 — Breakdown of electricity costs for the base-case fuel cell power plants (25 MWdc; useful life of 10,000 hours). The fuel gas is high-Btu gas (OTF)

of considerable importance. Although the fuel processing cost for the phosphoric acid system is small, approximately \$38/kWe (8.6%), this is not the case for the alkaline system, in which scrubbing of the carbon dioxide from the fuel gas and air before they enter the fuel cell subsystem is necessary. The cost of both scrubbers is included in the total of \$87/kWe shown for the alkaline fuel processing. The power-conditioning costs are similar, and lie in the range of \$62 to 66/kWe. The differences arise from the assumptions of dissimilar parasitic losses for each system. The balance of plant costs are slightly higher for the two high-temperature power systems because they include charges for recuperative heat exchangers necessary for the heating of the input air and fuel gas streams.

From a total capitalization standpoint, the alkaline power system, at \$620/kWe, is the most expensive, due to the low power density assumed for the fuel cell subsystem and the need for carbon dioxide scrubbing. The phosphoric acid and solid electrolyte systems, at \$448 and \$424/kWe, respectively, require the least investment, and the molten carbonate system requires an intermediate capitalization of \$514/kWe. The relative importance of the capitalization, at a fixed charge rate of 18%, is shown in Figure 13.17, which presents a breakdown of the COE for the base cases of all four power systems. The power plant and overall energy efficiencies for each case are also shown. The fuel gas is high-Btu gas costing \$2.46/GJ (\$2.60/10⁶ Btu); the useful life of all fuel cell subsystems was assumed to be 36 Ms (10,000 hr). The fuel charges are less for the high-temperature systems because of their greater efficiencies. Their O&M charges are greater, however, because of their higher fuel cell subsystem replacement costs.

A better basis for comparison is afforded by the data for Point 4 of every system. In every case, the 900 MW dc fuel cell subsystem uses medium-Btu gas as a fuel and is assumed to have a useful life of 36 Ms (10,000 hr). In addition, each of the high-temperature systems includes a steam bottoming plant consisting of steam generators,

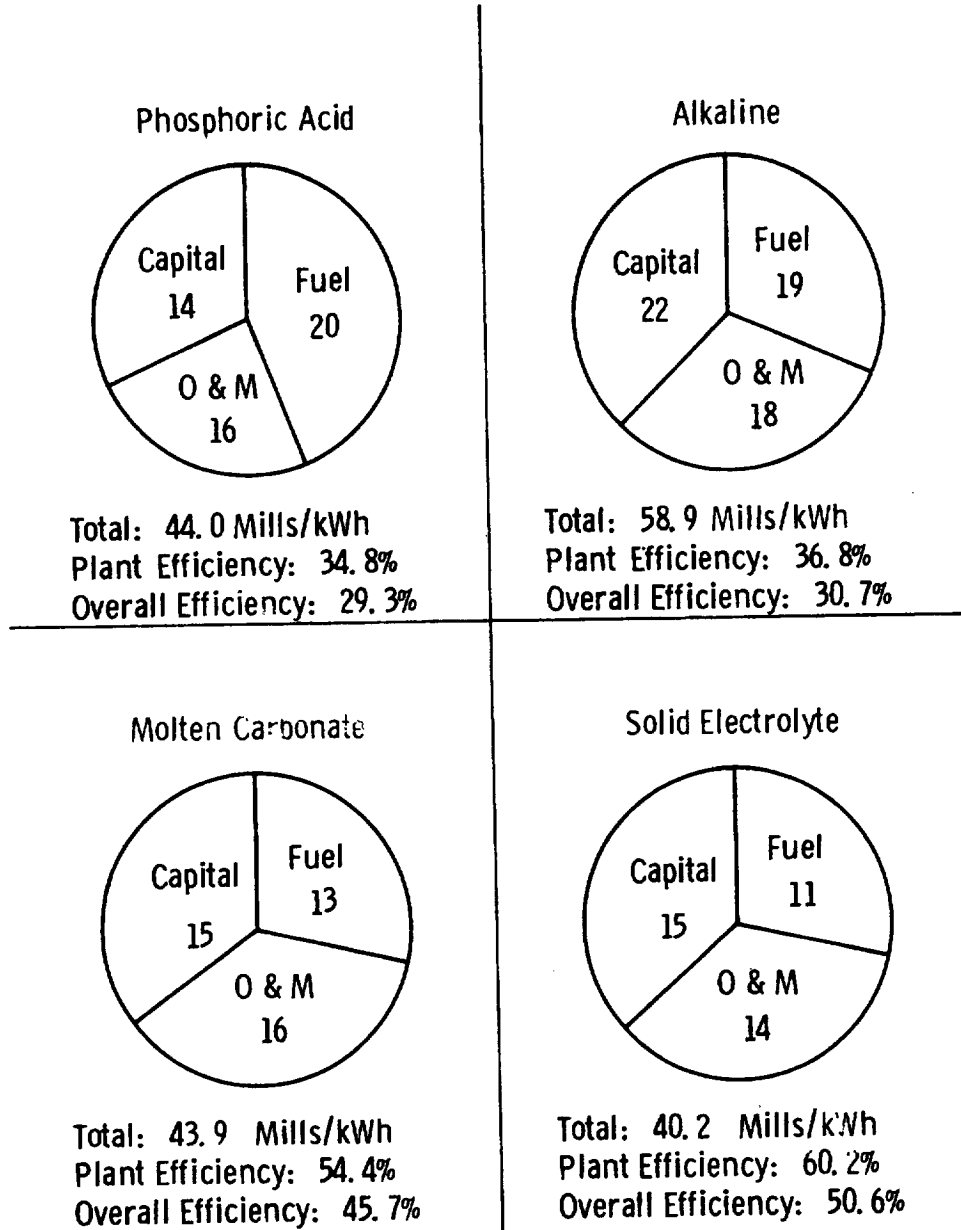


Fig. 13. 18—Breakdown of electricity costs for all 900 MW dc fuel cell power plants. The fuel gas is medium-Btu gas (OTF), and the useful life of all fuel cell subsystems is 10,000 hours

a steam turbine, and a heat rejection facility. This raises the ac output of the molten carbonate and solid electrolyte plants to 1170 and 1164 MW, respectively.

A breakdown of the capital costs for each system is shown in Figure 13.16. The total capitalizations required for the phosphoric acid, molten carbonate, and solid electrolyte systems are seen to be virtually identical, lying in the range of \$440 to 480/kWe. These costs would be even closer but for the assumption of a 157.7 Ms (5 yr) construction period for the high-temperature systems, as against a 126.1 Ms (4 yr) period for the acid system. Accordingly, the indirect costs amount to approximately 37.5% of the total for the acid system versus approximately 44% for the high-temperature systems. The alkaline system, with additional problems posed by the use of medium-Btu gas, necessitating the removal of carbon dioxide at twice the rate of that when high-Btu gas is employed as the fuel, is now noncompetitive, lying at \$700/kWe. For convenience of presentation, the fuel processing costs for the acid system have been included in the balance of plant costs in Figure 13.16.

The COE breakdowns and both efficiencies for each of the 900 MW dc systems are shown in Figure 13.18. The overall energy efficiencies for the high-temperature systems are, as expected, much higher than for either of the low-temperature systems. The higher fuel cost for the acid system, however, is offset by the lower O&M charge due to the lower fuel cell subsystems replacement cost, so that the total COE is essentially the same for the molten carbonate and acid systems at 12.2 mills/MJ (44 mills/kWh). The COE for the solid electrolyte system is lower still at 11.2 mills/MJ (40.2 mills/kWh). The alkaline system displays a COE of 16.4 mills/MJ (58.9 mills/kWh), which is substantially higher than for any of the other systems.

Projections as to the lowest COE possible for each system may be made on the basis of the data shown in Figure 13.18. These projections are highly tentative and are based on the multitude of assumptions

presented and discussed in Subsections 13.4 and 13.5. If medium-Btu gas, costing \$2.46/GJ (\$2.60/10⁶ Btu), is employed as the fuel; air is the oxidant; and the fuel cell subsystem life is at least 144 Ms (40,000 hr); then the COE will be as shown in Table 13.15.

Table 13.15 - Projections of Possible Efficiencies and Electricity Costs of Fuel Cell Power Systems

System Type	Overall Energy Efficiency, %	Possible COE, mills/kWh
Phosphoric Acid	~30	High 30s
Alkaline	~30	Low 40s
Molten Carbonate	~45	Low 30s
Solid Electrolyte	~50	High 20s

The costs and efficiencies for the high-temperature systems are predicated on the use of a waste-heat recovery system. This probably limits the minimum size of fuel cell subsystem to approximately 200 MW dc in order to allow economical and efficient recovery of the rejected heat.

The selection of fuel cell power systems for inclusion in Task II, Conceptual Design Preparation, and in Task III, Implementation Assessment, of this study, was based on the criterion of an overall energy efficiency significantly in excess of 35%. This eliminates all of the low-temperature fuel cell plants, and most of the high-temperature plants, which do not incorporate a waste-heat recovery system. Because of their high overall efficiencies, a solid electrolyte and molten carbonate power plant, as typified by Points SE4 and MC4, is recommended for the further refinement of efficiency and electricity cost estimates specified for Task II.

The Westinghouse Solid Electrolyte Fuel Cell Power System is recommended also for inclusion in Tasks II and III. Inspection of Figure 13.19, which presents all of the data pertinent to the recommended cases,

reveals why. The overall efficiency has been estimated very conservatively in this study. The estimated value of 53%, lower than previously published values of 60% and 57.5% (Reference 13.25) is, nevertheless, the highest determined in this study. The COE of 13.3 mills/MJ (47.7 mills/kWh) is inflated by the capitalization associated with coal gasification. This estimate of \$335/kWe is based on an evaluation performed in late 1968 (Reference 13.45), in which approximately 75% of the installed cost of the special fluidized bed coal gasification reactor was attributable to the cost of Incoloy 800 sheathing, considered necessary for efficient heat transfer from the fuel cell modules. In Task II, alternative materials and methods for efficient and economical thermal coupling of the gasifier and fuel cell subsystem should be explored. The potential for a reduction in the COE, and specifically, its capital component, is obvious from Figure 13.19.

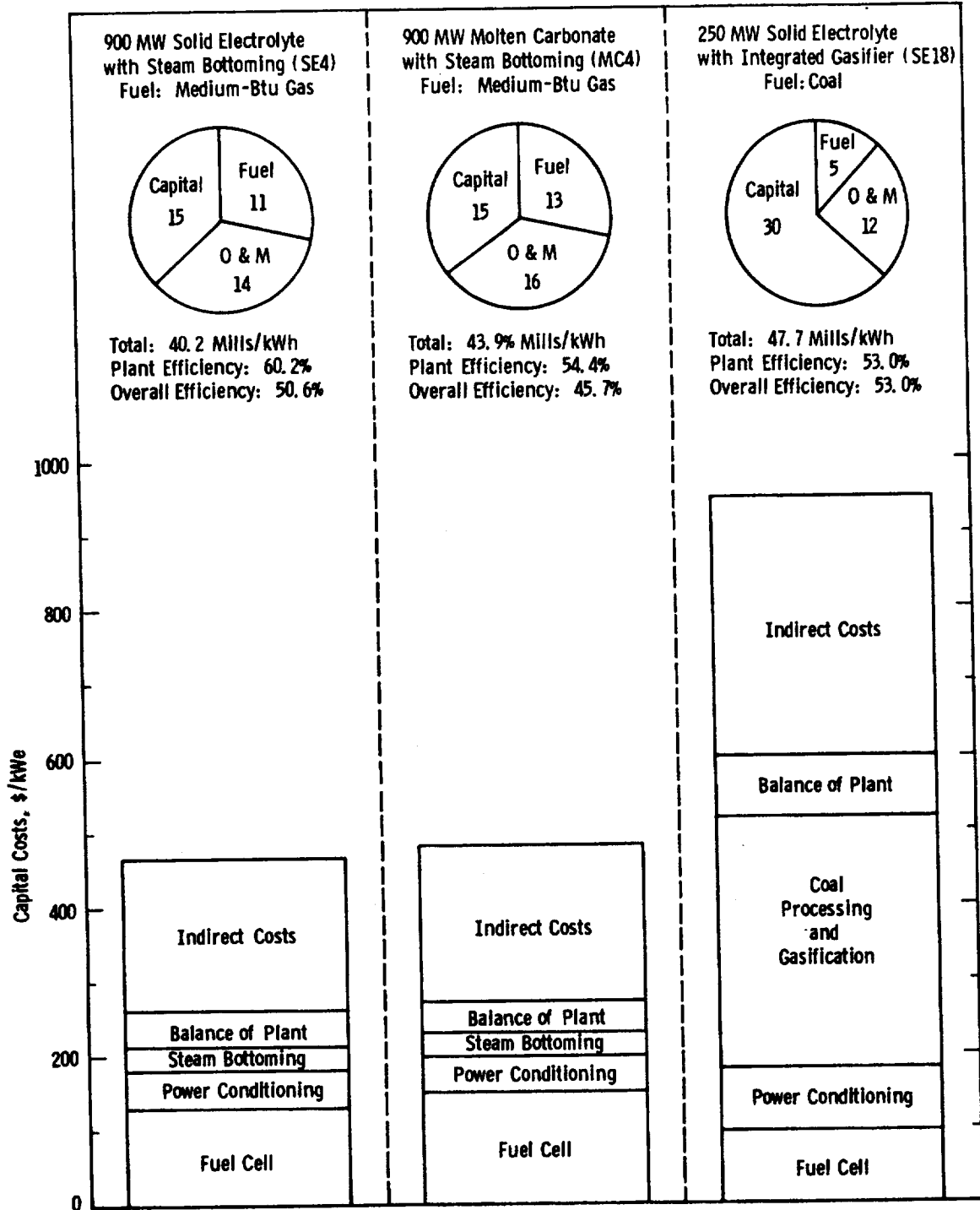


Fig. 13.19—Breakdown of electricity and capital costs for the three power plants (corresponding to parametric points SE4, MC4, and SE18) recommended for further investigation in Tasks II and III. Note that the assumed useful life of all fuel cell subsystems is 10,000 hours

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Appendix A 13.1

FUEL PROCESSING FOR LOW-TEMPERATURE FUEL CELL POWER PLANTS

The fuels employed in the parametric assessment of the low-temperature fuel cell power systems are medium-Btu gas (AC4 and AL4), methanol (AC14), and high-Btu gas (all other points). In order for these fuels to be usable at the anodes of the acid or alkaline fuel cell modules, they must be converted to a fuel gas consisting principally of hydrogen.

For high-Btu gas, consisting principally of methane, steam reformation coupled with shift conversion is the most economical method of producing this fuel gas. Carbon dioxide removal from the fuel gas stream is necessary in the alkaline case to prevent conversion of the potassium hydroxide electrolyte to potassium carbonate. This step is not required in the acid fuel cell system, as phosphoric acid does not react with carbon dioxide. Auxiliary equipment necessary in both power systems includes steam generators to supply the steam requirements of the reformer. Thus, an acid fuel cell power plant, fueled with high-Btu gas, requires a steam reformer, a shift converter, and a steam generator. In addition to these components, the alkaline fuel cell system must include a carbon dioxide removal subsystem.

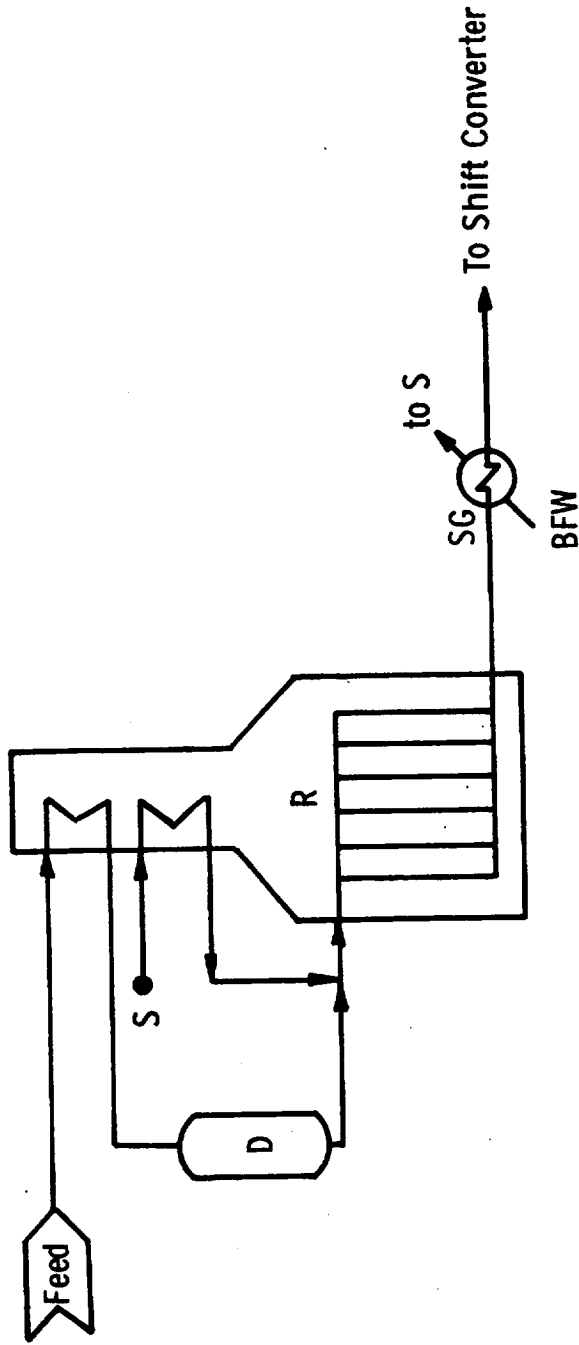
When medium-Btu gas, which is principally comprised of carbon monoxide (approximately 55% by volume) and hydrogen (approximately 33% by volume), is the fuel, there is obviously no need for steam reformation. Shift conversion and steam generation, to meet the steam requirements of the shift converter, are still required. As stated in Subsection 13.3, the carbon dioxide content of the medium-Btu gas stream after shift conversion is approximately double that in the shift-converter effluent for high-Btu gas. Thus, for alkaline power plants

using medium-Btu gas as the fuel carbon dioxide must be scrubbed at twice the rate necessary in plants operating on high-Btu gas. Although methanol may be cracked directly to form hydrogen and carbon monoxide (Reference 13.5), this study assumed that methanol, used in the power plant corresponding to Point AC14, is fed to a steam reformer, just as in the case of high-Btu gas.

To deliver one kilowatt of electrical power from a phosphoric acid fuel cell operating at a terminal voltage of 0.7 V, a cell current of 1429 A is required. Thus, a minimum of 1429/96489 equivalentents or 14.9 mg/s (3.28×10^{-5} lb/s) of hydrogen must be delivered to the anode. The hydrogen requirement for an alkaline fuel cell is lower because of the higher cell voltage and may be calculated by the use of a multiplier, 0.7 V/0.8 V.

The optimum level of hydrogen utilization in a fuel cell is a complex function of the fuel cell performance and the relative costs of the fuel cell subsystem, the fuel processing subsystem, and the fuel, as the unused hydrogen may be employed to provide the thermal requirement of the steam reformer, as described below. The preparation of a detailed conceptual design of the complete power system, coupled with the knowledge of fuel cell performance as a function of hydrogen utilization, would permit accurate estimation of this optimum level. For the purposes of this preliminary study, however, utilization rates of 90 and 80% were assumed for feedstocks of high- and medium-Btu gas, respectively, for the reasons outlined in Subsection 13.3.1. These assumptions lead to the hydrogen requirements for the four ratings of fuel cell subsystems shown in Table A 13.1.1. The 900 MW power plants operate with medium-Btu gas as fuel; all others employ high-Btu gas as a feedstock.

In the following subsections, the procedures employed in the costing of the reformer, the shift converter, the steam generators and the carbon dioxide removal subsystem, necessary for the production of the required hydrogen, are described. In addition, a brief description



D: De-sulfurizer

S: Steam feed to reformer

SG: Steam-Generator

BFW: Boiler feed water

R: Gas-Fired Steam Reformer

Fig. A 13.1.1 --Schematic of a steam-methane reformer system

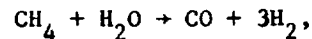
Table A 13.1.1 - Hydrogen Requirements of Low-Temperature Fuel Cell Power Plants

Fuel Cell Rating, MW	Hydrogen Requirements, tons/d	
	Acid	Alkaline
25	39.5	34.5
100	158	138
250	395	345
900	1600	1400

is given of the cost assumptions for the blowers used for the circulation of air through the cathode components of the fuel cell modules.

A 13.1.1 Steam-Methane Reformer

A schematic of a typical steam-reforming unit, is shown in Figure A 13.1.1. The reforming furnace is gas fired, and the convection tube banks are fabricated from carbon steel. The radiant tubes, made from stainless steel, operate at a pressure of 0.689 MPa (100 psi) abs. High-Btu gas is preheated, desulfurized by passage through activated carbon beds, mixed with preheated steam, and fed to the catalyst-filled furnace, which operates at 1033 to 1144°K (1400 to 1600°F). This reaction



$$\Delta H_{1600^\circ\text{F}} = 97,400 \text{ Btu/lb-mole CH}_4$$

is highly endothermic. As one lb-mole of methane produces four moles of hydrogen (after shift conversion) or 42.87 std m³ (1514 scf), the reformer heat duty, Q/P, may be calculated as shown:

Table A 13.1.1.2 - Heat Duty and High-Btu Gas Requirements and Base Costs (Mid-1974) of Steam-Methane Reformers for Hydrogen Production in Fuel Cell Power Plants

Hydrogen Production Rate ton/d $\times 10^{-3}$, scf/hr	Reformer Heat Duty $\times 10^{-6}$, Btu/hr	High-Btu Gas Usage $\times 10^{-3}$, scf/hr	Base Cost $\times 10^{-3}$, \$
34.5	34	200	220
39.5	39	230	250
138	140	800	550
158	160	920	590
345	340	2,000	1,100
395	390	2,300	1,200

Table A 13.1.1.3 - Breakdown of Adjustments to Base Costs (Mid-1974) of Steam-Methane Reformers for Hydrogen Production in Fuel Cell Power Plants

Base Cost $\times 10^{-3}$, \$	Adjustment Factors			Total	Purchased Equipment Base Cost $\times 10^{-3}$, \$
	Furnace Type	Tube Material	Tube Pressure		
220	1.345	0.345	0.05	1.740	380
250	1.345	0.345	0.05	1.740	440
550	1.342	0.341	0.05	1.733	950
590	1.342	0.341	0.05	1.733	1,020
1100	1.341	0.340	0.05	1.731	1,900
1200	1.341	0.340	0.05	1.731	2,100

$$Q/P = \frac{97,400 \text{ Btu}}{1514 \text{ scf}} = 64.3 \text{ Btu/scf of H}_2 \text{ required.}$$

The reformer heat duty for all six hydrogen production levels is presented in Table A 13.1.2. Shown also are the total quantities of high-Btu gas required, based on arguments presented in Section 13.3.1.

Using these heat-duty values, the base cost of the steam-methane reformer may be calculated. The method of cost estimation for all fuel conditioning costs follows an approach outlined by Guthrie (Reference 13.44) which uses mid-1970 prices. The costs have been factored upward by the ratio of the average Marshall and Stevens indices (Reference 13.48) for the second and third quarters of 1974 and 1970 (1.34). The base costs in mid-1974 terms are shown in the last column of Table A 13.1.2.

These base costs must be adjusted upward by a multiplier which is the sum of factors to allow for furnace type, radiant tube material, and pressure. This yields purchased equipment base costs, which are shown in the last column of Table A 13.1.3.

The total direct cost of the reformer includes the purchased equipment base cost computed earlier, plus the materials and labor required for installation. Following Guthrie (Reference 13.44), installation materials (less concrete) typically average 25.4%. These are summarized in Table A 13.1.4. The total cost shown does not include indirect costs associated with construction overhead, engineering, interest during construction, etc.

A 13.1.2 Shift Converter

After leaving the steam reforming unit, the gases are cooled to between 644 and 700°K (700 and 800°F) and passed over a water-gas shift catalyst to convert the carbon monoxide component to carbon dioxide and hydrogen by the reaction



Table A 13.1.1.4 - Reformer Costs (Mid-1974) for Fuel Cell Power Plants

System Type	Fuel Cell Rating, MW	Reformer Costs x 10 ⁻³ , \$			
		Purchased Equipment	Installation Material	Total Material	Site Labor
Alkaline Acid	25	380	97	480	110
Alkaline Acid	25	440	110	550	130
Alkaline Acid	100	950	240	1,200	280
Alkaline Acid	100	1,020	260	1,300	300
Alkaline Acid	250	1,900	480	2,400	550
Alkaline Acid	250	2,100	530	2,600	610

Table A 13.1.1.5 - Shift-Converter Volumes, Dimensions, and Base Costs (Mid-1974)

Hydrogen Required ton/d	Volume, ft ³		Vessel Dimensions, ft		Base Costs x 10 ⁻³ , \$
	Catalyst	Vessel	Length	Diameter	
34.5	270	540	23	5.5	36
39.5	320	640	24	6	45
138	1,100	2,200	35	9	77
158	1,250	2,500	39	9	81
345	2,700	5,400	70	10	140
395	3,200	6,400	80	10	180
1400	11,000	22,000	90	18	610
1600	13,000	26,000	100	18	640

A schematic of the shift conversion unit is shown in Figure A 13.1.2. The shift converter is a pressure vessel loaded with an appropriate catalyst. In the four smallest sizes a length-to-diameter ratio of four is assumed. If a space velocity of 0.556 s^{-1} ($2,000 \text{ hr}^{-1}$) is assumed, the volume of catalyst required may be calculated. The overall volume of the vessel is considered to be twice that of the catalyst.

The costing technique again follows that of Guthrie (Reference 13.44), with an escalation factor of 1.34 to convert mid-1970 costs to those of mid-1974. In the penultimate column of Table A 13.1.6, the base costs (mid-1974) shown are for a carbon steel vessel operating at 0.345 MPa (50 psi) abs. The costs include shell and two heads, nozzles and runways, skirt, base ring, and lugs. The shift converter, however, should operate from 2.76 to 3.45 MPa (400 to 500 psi) abs for compatibility with the steam-methane reformer. All costs at 0.345 MPa (50 psi) abs were multiplied by 2.8 to allow for high-pressure design costs. These costs are shown in the last column of Table A 13.1.7.

Installation costs of vertical pressure vessels are typically 3.05 times the base costs (Reference 13.44). The total material and site labor costs of the shift converter units for the different ratings of acid and alkaline fuel-cell power plants are shown in Table A 13.1.5.

A 13.1.3 Steam Generators

An excess of steam, generally three to five times the stoichiometric requirement, is employed in the steam-methane reformer. Steam is raised in three main locations: 1) between the reformer and shift converter, in the cooling of the fuel gases from 1144 to 700°K (1600 to 800°F), as shown in Figure A 13.1.1; 2) by the use of the heat rejected by the shift converter; and 3) in cooling the hydrogen-rich fuel gas from 700 to 464°K (800 to 375°F), the temperature of operation of the acid fuel-cell modules. The steam-generators of 2 and 3 are shown schematically in Figure A 13.1.2.

For the purposes of calculation, it will be assumed that 3 moles of steam are required in the reformer for every 4 moles of hydrogen

Table A 13.1.6 - Shift-Converter Costs (Mid-1974) for Fuel Cell Power Plants

System Type	Fuel Cell Rating, MW	Shift Converter Costs x 10 ⁻³ , \$			
		Base	Installation Material	Total Material	Site-Labor
Alkaline	25	36	20	56	21
Acid	25	45	25	70	26
Alkaline	100	77	43	120	45
Acid	100	81	46	130	47
Alkaline	250	140	79	220	81
Acid	250	180	100	280	100
Alkaline	900	610	340	950	350
Acid	900	640	360	1,000	370

Table A 13.1.7 - Steam Requirements and Generator Costs (Mid-1974) for Fuel Cell Power Plants

Hydrogen Production Rate (ton/d) (x 10 ³ , scf/hr)	Steam Rate x 10 ⁻³ , lb/hr		Steam Generator Base Cost x 10 ⁻³ , \$
	Hydrogen Production Rate	Steam Rate x 10 ⁻³ , lb/hr	
34.5	530	19	66
39.5	610	22	77
138	2,100	75	199
158	2,400	85	230
345	5,300	190	350
395	6,100	220	420
1400	21,000	750	1,500
1600	24,000	850	1,600

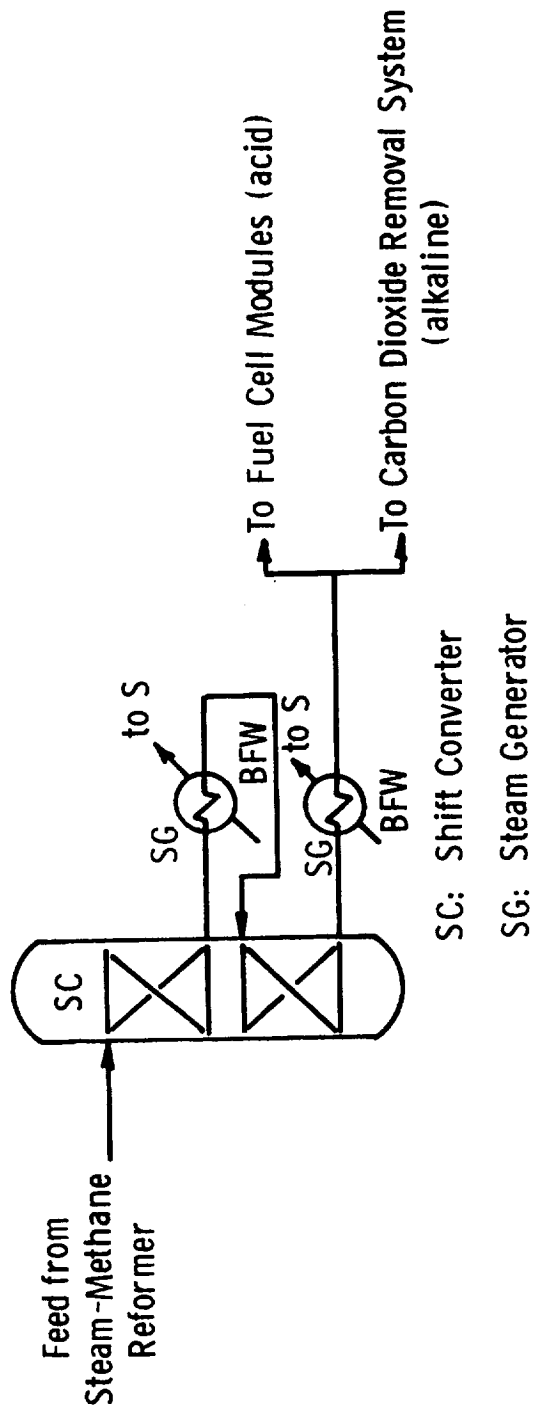


Fig. A 13.1.2 – Schematic of shift conversion unit for fuel-cell power plants

Table A 13.1.1.8 - Steam Generator Costs (Mid-1974) for Fuel Cell Power Plants

System Type	Fuel Cell Rating, MW	Steam Generator Costs x 10 ⁻³ , \$			
		Base	Installation Material	Total Material	Site-Labor
Alkaline Acid	25	66	8.4	74	20
Alkaline Acid	25	77	9.9	87	24
Alkaline Acid	100	190	24	210	58
Alkaline Acid	100	230	29	260	70
Alkaline Acid	250	350	45	400	110
Alkaline Acid	250	420	54	470	130
Alkaline Acid	900	1,500	140	1,700	450
Alkaline Acid	900	1,600	200	1,800	470

Table A 13.1.1.9 - Estimates of Material and Site Labor Costs for Fuel Gas Scrubbing in an Alkaline Fuel Cell Power Plant

Fuel Cell Rating, MW	Hydrogen Required, ton/d	Fuel Gas Flow Rate, lb-moles/hr	Material Costs x 10 ⁻³ , \$	Site Labor Costs x 10 ⁻³ , \$
25	34.5	1,760	420	140
100	138	7,200	1,700	560
250	345	17,600	4,200	1,410
900	1400	72,000	23,000	7,700

reaching the fuel cell modules. The steam requirements are, therefore, 0.570 kg of steam/std m³ of hydrogen (0.0356 lb of steam/scf of hydrogen). Table A 13.1.7 presents the steam rates and steam generator base costs (mid-1974) as a function of hydrogen production. The costing method again follows that of Guthrie (Reference 13.44). The total material and site labor costs are shown in Table A 13.1.8 for all four ratings of acid and alkaline fuel cell power plants.

A 13.1.4 Carbon Dioxide Removal System

Leaving the shift converter, the fuel gas steam is at a temperature of 700°K (800°F) and a pressure of 0.483 MPa (70 psi) abs. The carbon dioxide-carbon monoxide ratio is approximately 50. The gas composition is ~ 70% hydrogen, ~ 10% steam, and the balance carbon oxides and inerts. The steam may be cooled and the steam condensed to yield a gas mixture of 80% H₂, 17% carbon oxides, and 3% inerts. This fuel gas may be fed directly to the acid fuel-cell modules.

In the alkaline system, however, carbon dioxide removal must still be accomplished in order to protect the potassium hydroxide electrolyte. The process considered for this application is shown schematically in Figure A 13.1.3. It consists of the Lurgi Rectisol Process, which uses refrigerated methanol. The total capital investment for a Rectisol System capable of stripping carbon dioxide and hydrogen sulfide from 4.16 kg-moles/s (33,000 lb moles/hr) of fuel gas in the Bituminous Coal Research Bi-Gas Process (Reference 13.49) was \$23.5 million (mid-1970) (Reference 13.50). When factored upward by 1.34 to convert to mid-1974 costs, the total capital investment required is approximately \$0.582 per kg-mole/s (\$950 per lb-mole/hr) of fuel gas.

If it is assumed that the direct installed cost of the scrubbing system represents 28.5% of the total capital investment (Reference 13.52) and that the material-to-site labor ratio is equal to three, approximate estimates of the equipment and site labor costs for

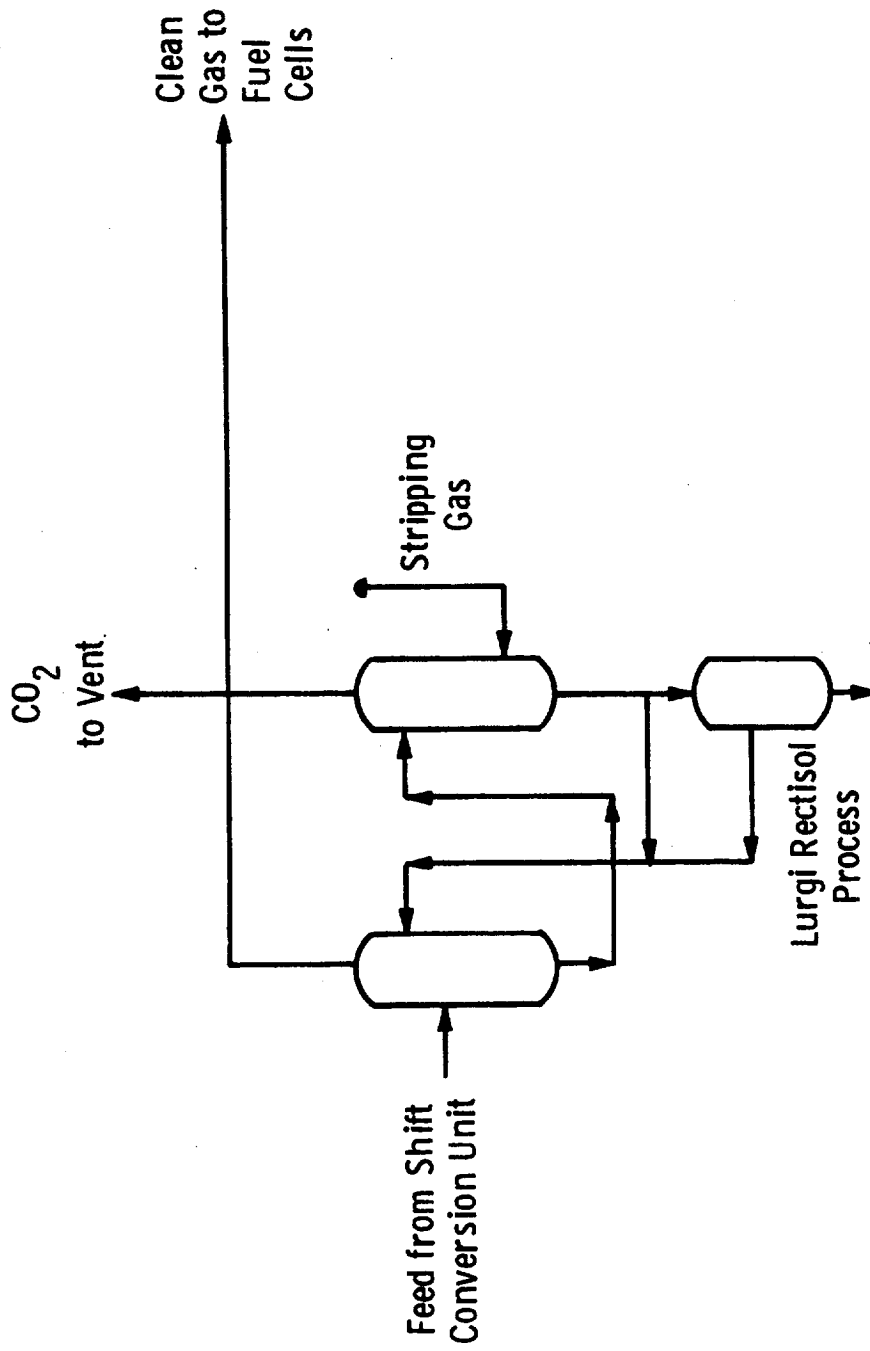


Fig. A 13.1.3—Schematic of carbon dioxide removal system for alkaline fuel-cell power plants

the Rectisol Process may be derived. These are shown for the four ratings of alkaline fuel cell power plants in Table A 13.1.9. A linear relationship is assumed between scrubber cost and volume of gas scrubbed. Because of the additional load on the scrubbing system of the 900 MW plant due to the use of medium-Btu gas as the fuel the costs are multiplied by 1.35.

Two of these scrubbing units are required per power plant, as carbon dioxide must also be removed from air. Although air contains approximately four orders of magnitude less carbon dioxide than does the fuel gas, the costs associated with scrubbing air were assumed to be the same as those given in Table A 13.1.9. In general, the equipment cost was assumed to be a linear function of the quantity of gas passed through the process. For a 50% utilization of oxygen in air, the quantity of throughput air would be approximately a factor of eight greater than that of fuel gas. The advantages associated with the lower concentration of carbon dioxide in air, however, were assumed to approximately negate the eight-to-one flow penalty. The treatment above, though very crude and allowing for no economy of scale, yields reasonable costs for scrubbing equipment to remove carbon dioxide from fuels and air. The value of \$22/kW for air scrubbing in this closed-cycle process is acceptable when compared with \$5/kW for the much simpler process involving the use of potassium hydroxide on a once-through basis (Reference 13.8).

A 13.1.5 Air Blowers

The costing of air blowers for the fuel cell subsystem was performed following the method of Guthrie (Reference 13.44). The air blower cost, C, in mid-1974 dollars is given by

$$C = \$ (9) (1.34) (\text{Required airflow rate in scfm})^{0.68}$$

With the assumption of 50% utilization of the oxygen in the throughput air, a value of \$19,000 is calculated for the air blowers in the 25 MW

phosphoric acid fuel cell power plant. A site labor cost equivalent to 33% of the equipment cost was arbitrarily assumed. The material and site labor costs for all power plants were derived similarly.

Possible errors in these estimates arising from the under-estimation of the pressure drop through the fuel cell modules are likely to have a negligible impact on the total capitalization of the fuel cell power plant. A total material and site labor cost of \$25,000 for the 25 MW acid system air blowers is trivial when compared with, for example, a cost of \$3.8 million for the fuel cell subsystem.

Appendix A 13.2
POWER-CONDITIONING SUBSYSTEM

Power-conditioning subsystems are necessary in fuel cell power plants for the conversion of the dc output of the fuel cell modules to 60 Hz ac power. NASA has specified that the 25 MW plant* should deliver power at 69 kV to the distribution net, but the output from the other three sizes — 100*, 250*, and 900 MW* — should be at 500 kV. The following is a discussion of inverter subsystems and transformers which are necessary to accomplish the required power conversion, and the resultant implications for the selection of fuel cell module sizes.

In general, there are at least seven power conversion schemes that can be considered for this application. These are: (1) chopper-inverter; (2) inverter; (3) buck-boost inverters; (4) complementary inverters; (5) HF link; (6) hybrid (HF link + simple line-commutated inverter), and (7) force-commutated inverter. In a recent study at Westinghouse Research Laboratories (Reference 13.51), scheme 7 was found to be optimum from an economic and technical standpoint. Force-commutated systems not only operate with lower losses (4-1/2 to 5% versus 5 to 6% for the more conventional line-commutated systems) but also offer considerable operational advantages:

- They will ride through a system fault.
- It is possible to control their behavior with respect to reactive power demand and deliver independently of

*25, 100, 250 and 900 MW ratings apply to fuel cell subsystems. The overall plant ratings are less, due to inverter subsystem losses and other parasitic losses in the fuel cell power system.

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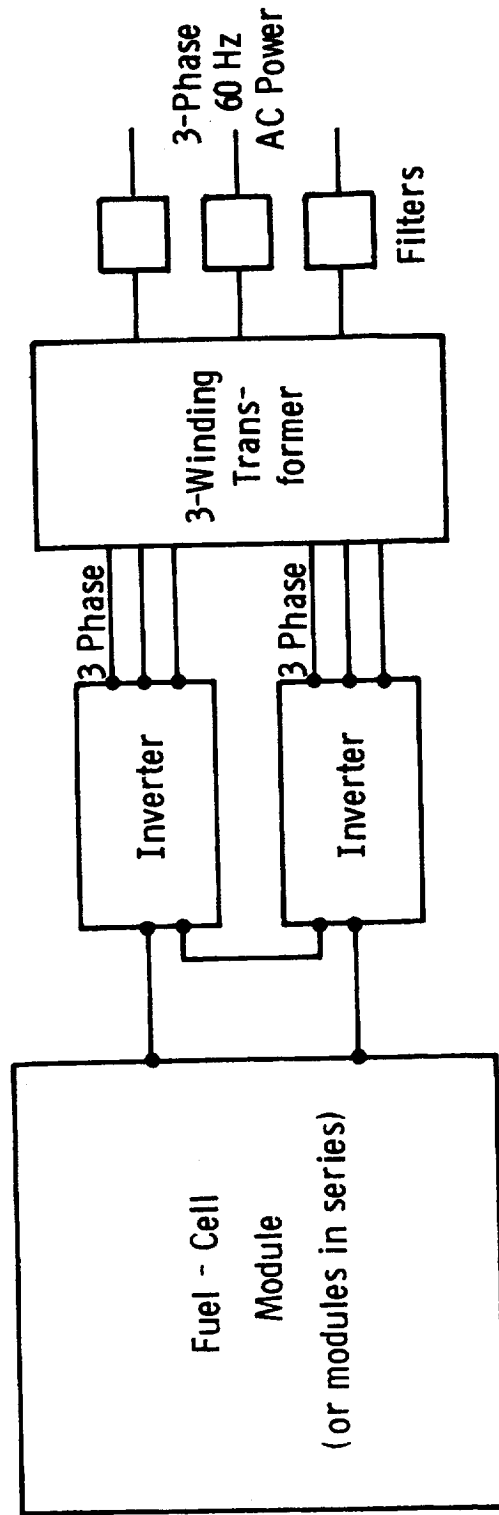


Fig. A 13.2.1 -- Block diagram of power conversion equipment required for fuel-cell power plants

real power, thus conferring great stability on the power conversion system.

- They will start and run into a passive load.

Force-commutated inverter systems, however, suffer from a major limitation in that the state of the art indicates an upper limit of approximately 2 kV output voltage. Because of transformer primary-current considerations, this constraint limits use of this type of inverter system to the 25 MW case.

For the larger power plants (100, 250, and 900 MW) a line-commutated inverter system must be used. Systems 1 through 3 are typical line-commutated schemes and represent relatively simple extensions of high voltage dc (HVDC) technology. Of these three, the buck-boost inverter is optimum from a cost standpoint. Further, it is close to being the most efficient and has no operational disadvantages compared to the other line-commutated schemes. Most significantly, power factor improvement is achievable with this scheme. This results in a considerable savings in cost because auxiliary power-factor correction equipment is not needed. Although this inverter system can run, in theory, into a passive load, the stability of this operational mode is questionable in practice unless a synchronous capacitor is used to supply reactive VA. Another disadvantage of line-commutated inverter systems, which applied also to this scheme, is that the dc side must be quickly interrupted after a system fault in order to protect the overall system. From this standpoint, line-commutated schemes compare unfavorably with force-commutated systems, which, as noted above, are able to ride through system faults.

The basic inverter unit, in the force- and line-commutated systems above, is the 3-phase bridge or Graetz connection. This is illustrated in Figure A 13.2.1. It should be noted that the ac outputs of the two inverters are out of phase by $\pi/6$. The transformer required for step-up of the inverters' output to distribution voltages levels should have a twin core structure. For the line-commutated inverter

scheme, wye-delta and delta-delta windings are specified, while in the force-commutated inverter case, one core should involve a wye-open wye winding plus a delta tertiary, and the other a delta-wye winding. These specifications, plus the poor power factor at which the transformers operate, result in transformer costs which are approximately double those of conventional transformers of the same kVA rating (Reference 13.52).

Table A 13.2.1 outlines the relationship between the dc and three-phase ac currents and voltages for the force- and line-commutated systems, where α is the firing angle delay of the line-commutated inverter, and ϕ is the phase-shift angle of the force-commutated inverter.

Table A 13.2.1 - The Relationships Between Electrical Input and Output of Inverter Systems

System Parameters	Inverter System Type	
	Force-Commutated	Line-Commutated
Voltage	$V_{ac} = \left(\frac{\sqrt{6}}{\pi}\right) V_{dc}$	$V_{ac} = \left(\frac{\pi}{3\sqrt{2}}\right) \cos\alpha V_{dc}$
Current	$I_{ac} = \left(\frac{\pi}{3\sqrt{2}}\right) (\cos\phi) (I_{dc})$	$I_{ac} = \left(\frac{\sqrt{6}}{\pi}\right) I_{dc}$
Power	$\sqrt{3} (V_{ac}) (I_{ac}) (\cos\phi)$	$\sqrt{3} (V_{ac}) (I_{ac}) (\cos\alpha)$

The sizes, weights, and selling prices of inverter systems for the four power plant ratings are shown in Table A 13.2.2. All the data shown are based on the results of a previous study of inverter systems for 25 MW fuel cell power plants. The plan area/MW (Reference 13.51) for the buck-boost systems at the various rating levels were calculated

by multiplication of the plan area requirement at the 25 MW level by the cube root of the ratio of the ratings of the larger system to the 25 MW unit. This factoring upwards ensures adequate clearance at the higher ratings, thus minimizing problems associated with creepage paths. The weights are taken from the above mentioned study (Reference 13.51) The selling price data are derived from estimates of \$60/kW and \$55/kW for 25 MW line-commutated and force-commutated systems respectively, and from an estimate of \sim \$65/kW for the buck-boost system at the 900 MW level. The values for the 100 and 250 MW systems were derived by linear interpolation. The increases in capital costs on going to the larger plant sizes is due to higher costs of filters and the greater need for grading networks for the valves.

Because of the lack of demonstrated reliability of fuel cell modules in extended service, it is not unreasonable to specify for all fuel cell power plants that a fault on the dc side should result in outages of not more than 10% of the rated capacity. Thus, in all cases, ten banks of fuel-cell modules are envisioned, each with its own inverter pair and transformer. In general, only one set of filters and switch-gear will be used on the high sides of the transformers.

In the 25 MW plants, each of the banks of fuel cell module may be run at \sim 2200 A dc (1.15 kV dc). Single transformation from \sim 900 V ac to the required 69 kV ac level is achievable. In the 900 MW plants, ten banks of fuel cell modules are envisioned, each operating at 2900 A dc and \sim 31 kV dc. The \sim 23 kV ac output of the inverter pair permits single transformation to the required level of 500 kV ac.

Single transformation, however, is not possible without a significant inverter cost penalty in the 100 MW plants. Under the 10% outage constraint, each bank would contain 10 MW of fuel cell modules. As the ac output from the inverters must be at least 16 kV to facilitate single transformation to 500 kV ac, the fuel cell dc voltage requirements could be as high as \sim 27 kV, assuming a worst-case inverter power factor of 0.8. This would result in a dc current input to the inverters of

Table A 13.2.2 - Sizes, Weights, and Costs of Inverter Systems for Fuel Cell Power Plants

Fuel Cell Rating, MW	Inverter Type	Plan Area,* ft ² /MW	Weight, lb/MW	Selling Price,** \$/MW
25	Force-commutated	29	7,900	55,000
100	Buck-boost	23	4,800	60,000
250	Buck-boost	32	4,800	61,000
900	Buck-boost	48	4,800	65,000

* 8 ft total height excluding bushings, for 25 MW; approximately 30 ft for 100, 250, and 900 MW plants because of 500 kV tie.

** Prices include cost of single transformation. If double transformation is required, the costs shown must be increased by \$10,000/MW.

Table A 13.2.3 - Vendor, Site Labor, and Total Costs (Mid-1974) of Power Conditioning Subsystems for Fuel Cell Power Plants

Fuel Cell Rating, MW	Power-Conditioning Costs x 10 ⁻³ , \$		Total
	Equipment	Site Labor	
25	1,380	75	1,455
100*	7,000	300	7,300
250*	17,500	750	18,250
900	59,000	2,700	61,700

* Costs include additional transformation step at \$10,000/MW

~ 370 A which is too low for the inverters to be economical (Reference 13.51). Rather than pay inverter costs more than twice those for systems operating at ~ 2200 A dc, it is more reasonable to pay the additional price for double transformation. This is estimated (Reference 13.52) to be ~ \$10,000/MW. An intermediate ac voltage of 34.5 kV will optimize the system with a dc voltage in each 10 MW bank of 4.6 kV (2200 A dc).

Similar considerations apply to the 250 MW plant. The 25 MW banks would supply less than 1 kA dc under the worst-case condition mentioned above. Once again, double transformation is economically more attractive than paying for increased inverter capability. Here, an intermediate ac voltage of 69 kV is envisioned with a dc voltage of ~ 11.5 kV (2200 A dc) in each 25 MW bank.

For the purposes of this study, an efficiency of 95.5% is assumed for the power-conditioning subsystems of all 25 MW power plants, based on the use of a force-commutated inverter system. Line-commutated inverter systems must be employed in the 100, 250, and 900 MW power plants. Here, an efficiency of 95% is assumed. Double transformation, however, is required in all 100 and 250 MW power plants for the reasons outlined above. The second transformer is assumed to have an efficiency of 99.5%, so the net efficiency of the power-conditioning subsystem of the 100 and 250 MW plants is $(95\%)(0.995)$ or 94.5%.

The site labor costs are based on recent experience (mid-1974) of the Westinghouse Electric Corporation with the installation of a VAR generator (a device to inhibit voltage flickering in an arc furnace) in a steel-making facility of Akron Steel (Reference 13.53). The installation of this equipment, which is very similar in character to the power-conditioning subsystems described above, involved a site labor of approximately \$3/kW. This value has been employed in this study.

The equipment, site labor, and total costs of the power-conditioning subsystems for the four fuel cell power plant ratings are presented in Table A 13.2.3. The costs for the 100 and 250 MW power plants reflect the cost of double transformation. All these designs and

costs are based on a maximum input variation of the dc current and voltage to the power-conditioning subsystem of 5%. Point 9 in all four fuel cell power systems explores the effect of the variation of the voltage degradation at constant power for 5 to 15% in 25 MW plants. The power-conditioning subsystem must, therefore, have the capability of handling this greater variation of current and voltage characteristic of fuel cell subsystems. The cost of the power-conditioning equipment may be assumed to be proportional to the dc current (Reference 13.51). At constant power the system with 15% voltage degradation would have to carry 20% more current than the system with a 5% voltage degradation. The costs of the power-conditioning subsystem for AC9, AL9, MC9, and SE9 were estimated by adding 20% to the value for the base case.

Appendix A 13.3

OXYGEN PLANTS FOR FUEL CELL POWER SYSTEMS

A fuel cell plant may be integrated with an oxygen plant as well as with a coal gasification plant. If oxygen is supplied to both the coal gasification unit (for production of medium-or high-Btu gas) and to the fuel cell plant, the total oxygen cost will be of similar magnitude to the coal cost.

Oxygen requirements depend on the type and size of the fuel cell and source of the coal, and are roughly 0.8 and 1.4 Mg of oxygen per Mg of coal, respectively, for the gasification and fuel cell plants. Thus, substitution of oxygen for air at the fuel cell cathode would have to give a considerable improvement in the power density at a given efficiency to justify the high oxygen costs involved.

The cost of oxygen from an on-site plant as operated by the utility is a strong function of plant size and the cost of energy for the compressor. Requirement for a 25 MW fuel cell would be about 3.15 kg/s (300 ton/d) of oxygen. The maximum size plant design commercially available is 21 kg/s (~ 2000 ton/d) of oxygen, so that large fuel cell plants (e.g., 900 MW) would require multiple plants of this size. Only minor cost reductions would be realized for multiple plants on the same site, beyond the single 21 kg/s (2000 ton/d) plant size.

The cost of oxygen produced by different size commercial plants (utility operated) was estimated (Reference 13.54) on the basis of the approximate commercial plant cost and operational data supplied by Air Products Corporation (Reference 13.55). The costs are adjusted to July 1974 dollars. Table A 13.3.1 shows oxygen costs for two plant sizes,

Table A 13.3.1 - Oxygen Production Costs (330 use days per year)

Size, ton/d	300	2000
A. Motor-Driven Plants		
Capital charges at 18%	1,008,000	3,412,800
Maintenance at 2% of investment	112,000	379,200
Power at 3.0¢/kWh	946,600	5,132,000
Water at 10¢/1000 gal	63,100	342,100
Labor at \$30,000/man year	<u>180,000</u>	<u>240,000</u>
Total	2,309,700	9,506,100
\$/ton	23.33	14.41
B. Turbine-Driven Plants		
Capital charges at 18%	1,064,200	3,603,200
Maintenance at 2% of investment	118,200	400,400
Power at 3.0¢/kWh	17,900	121,200
Water at 10¢/1000 gal	248,800	1,344,400
Labor at \$30,000/man year	180,000	240,000
Steam - free	<u>0</u>	<u>0</u>
Total	1,629,000	5,709,200
\$/ton	16.45	8.65

3.15 and 21 kg/s (300 and 2000 ton/d) and for an electric motor driven compressor or a turbine driven compressor. The calculations assume that the turbine driven compressor is operated on steam generated from the free waste heat from the fuel cell and that the electric motor drive compressor uses electricity from the fuel cell costing 8.3 mills/MJ (30 mills/kWh), which would appear to be about the minimum cost that could be expected from a 315 Ms (10 yr) life optimized fuel cell from the present study. A recirculating water cost of \$ 0.026/m³ (\$ 0.10/1000 gal) is assumed. Also included are 18% per year capital based charges for depreciation, interest, local taxes, etc., plus 2% per year for maintenance.

We should note that the costs do not include the cost of the steam generating equipment which is used to remove waste heat from the fuel cell for the turbine-drive compressor plants, and that oxygen costs for electric motor-driven compressor plants will vary with the cost of electricity produced by the fuel cell.

Electric-driven compressors are assumed for Point 5 for the 25 MW acid, alkaline, and high-temperature solid electrolyte fuel cells, and the fuel cell electrical output is derated by the energy required. A turbine-driven compressor using steam generated by the waste heat is assumed for the molten carbonate Point 5 (25 MW dc fuel cell) and Point 17 (250 MW dc fuel cell), so the electrical output is not derated for those system points.

Appendix A 13.4
COMPUTER OUTPUT SHEETS

Table A 13.4.1 PHOSPHORIC ACID FUEL CELL SYSTEM ACCOUNT LISTING
PARAMETRIC POINT NO. 1

ACCOUNT NO. & NAME	UNIT	AMOUNT	MAT \$/UNIT	INS \$/UNIT	MAT COST,\$	INS COST,\$
SITE DEVELOPMENT						
1. 1 LAND COST	ACRE	1.5	100000.00	.00	150000.00	.00
1. 2 CLEARING LAND	ACRE	1.5	.00	600.00	.00	298.97
1. 3 GRADING LAND	ACRE	1.5	.00	3000.00	.00	4500.00
1. 4 ACCESS RAILROAD	MILE	.0	115000.00	110000.00	.00	.00
1. 5 LOOP RAILROAD TRACK	MILE	.0	120000.00	70000.00	.00	.00
1. 6 SIDING R R TRACK	MILE	.0	125000.00	80000.00	.00	.00
1. 7 OTHER SITE COSTS	ACRE	.0	2.063	.00	3736.39	3736.39
PERCENT TOTAL DIRECT COST IN ACCOUNT		1	= 2.063	ACCOUNT TOTAL,\$	153736.39	8536.36
EXCAVATION & PILING						
2. 1 COMMON EXCAVATION	YD3 FT	9600.0	.00	3.00	.00	28800.00
2. 2 PILING	FT	.0	6.50	8.50	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		2	= .366	ACCOUNT TOTAL,\$.00	28800.00
PLANT ISLAND CONCRETE						
3. 1 PLANT IS. CONCRETE	YD3	3200.0	70.00	80.00	224000.00	256000.00
3. 2 SPECIAL STRUCTURES	YD3	.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		3	= 6.103	ACCOUNT TOTAL,\$	224000.00	256000.00
HEAT REJECTION SYSTEM						
4. 1 COOLING TOWERS	EACH	.0	.00	.00	.00	.00
4. 2 CIRCULATING H2O SYS	EACH	1.0	.00	.00	.00	.00
4. 3 SURFACE CONDENSER	FT2	.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		4	= .000	ACCOUNT TOTAL,\$.00	.00
STRUCTURAL FEATURES						
5. 1 STAT. STRUCTURAL ST. ION	TON	.0	650.00	175.00	.00	.00
5. 2 SILOS & BUNKERS	TPH	.0	1800.00	750.00	.00	.00
5. 3 CHIMNEY	FT	.0	.00	.00	.00	.00
5. 4 STRUCTURAL FEATURES EACH	EACH	1.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		5	= .000	ACCOUNT TOTAL,\$.00	.00
BUILDINGS						
6. 1 STATION BUILDINGS	FT3	.0	.16	.16	.00	.00
6. 2 ADMINISTRATION	FT2	5000.0	16.00	14.00	80000.00	70000.00
6. 3 WAREHOUSE & SHOP	FT2	.0	12.00	8.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		6	= 1.907	ACCOUNT TOTAL,\$	80000.00	70000.00
FUEL HANDLING & STORAGE						
7. 1 COAL HANDLING SYS	TPH	.0	.00	.00	.00	.00
7. 2 DOLOMITE HAND. SYS	TPH	.0	.00	.00	.00	.00
7. 3 FUEL OIL HAND. SYS	GAL	.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		7	= .000	ACCOUNT TOTAL,\$.00	.00

Table A 13.4.1 PHOSPHORIC ACID FUEL CELL SYSTEM ACCOUNT LISTING
 Continued PARAMETRIC POINT NO. 1

ACCOUNT NO. & NAME	UNIT	AMOUNT	MAT \$/UNIT	INS \$/UNIT	MAT COST,\$	INS COST,\$
FUEL PROCESSING						
1	CCAL GASIFIER	EACH	.00	.00	.00	.00
2	DESULFURIZER	EACH	.00	.00	.00	.00
3	VENTURI SCRUBBER	EACH	.00	.00	.00	.00
4	COAL PRETREATMENT	EACH	58000.00	114000.00	58000.00	114000.00
5	REFORMER	EACH	67900.00	25300.00	57900.00	25300.00
6	SHIFT CONVERTER	EACH	1.00	1.00	19000.00	3000.00
7	CO2 SCRUBBER	EACH	1.00	5000.00	19000.00	3000.00
8	BLOWER	EACH	1.00	19000.00	649000.00	145700.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		2	10.047	ACCOUNT TOTAL,\$		
FIRING SYSTEM						
3	1	PERCENT TOTAL DIRECT COST IN ACCOUNT	9	.000	ACCOUNT TOTAL,\$.00
VAPOR GENERATOR (FIRED)						
10	1	PERCENT TOTAL DIRECT COST IN ACCOUNT	10	.000	ACCOUNT TOTAL,\$.00
ENERGY CONVERTER SYSTEM						
11	1	FUEL CELL GENERATOR EACH	1.00	330000.00	123000.00	125000.00
11	2	STM TURBINE GENERATOR EACH	.00	.00	.00	.00
11	3	OXYGEN PLANT EACH	.25	.25	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT	11	50.287	ACCOUNT TOTAL,\$		383000.00	125000.00
COUPLING HEAT EXCHANGER						
12	1	HEAT REC STEAM GEN EACH	1.00	96000.00	23000.00	23000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT	12	1.766	ACCOUNT TOTAL,\$		96000.00	23000.00
HEAT RECOVERY HEAT EXCH.						
13	1	RECUPERATOR EACH	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT	13	=	.000	ACCOUNT TOTAL,\$.00
WATER TREATMENT						
14	1	DEMINERALIZER GPM	.00	2500.00	700.00	.00
14	2	CONDENSATE POLISHING KWE	.00	1.25	.30	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT	14	=	.000	ACCOUNT TOTAL,\$.00
POWER CONDITIONING						
15	1	POWER CONDITIONING EQ EA	1.00	138000.00	75000.00	75000.00
15	2	STD TRANSFORMER KVA	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT	15	18.500	ACCOUNT TOTAL,\$		138000.00	75000.00
AUXILIARY MECH EQUIPMENT						
16	1	BOILER FEED PUMP 8DR KWE	.00	.55	.04	.00
16	2	OTHER PUMPS KWE	.00	.88	.12	.00
16	3	MISC SERVICE SYS KWE	13855.00	1.17	32583.42	24700.47
16	4	AUXILIARY BOILER PPH	.00	4.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT	16	.317	ACCOUNT TOTAL,\$		32583.42	24700.47

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Table A 13.4.1
Continued

PHOSPHORIC ACID FUEL CELL SYSTEM ACCOUNT LISTING
PARAMETRIC POINT NO. 1

ACCOUNT NO. & NAME	UNIT	AMOUNT	MAT \$/UNIT	INS \$/UNIT	MAT COST.\$	INS COST.\$
PIPE & FITTINGS						
17. 1 CONVENTIONAL PIPING	TON	10.0	3000.00	1800.00	30000.00	18000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 17 =			.610 ACCOUNT TOTAL.\$		30000.00	18000.00
AUXILIARY ELEC EQUIPMENT						
18. 1 MISC MOTORS, ETC		6767.3	1.40	.17	9474.15	1150.43
18. 2 SWITCHGEAR & MCC PAN KWE		33836.3	1.95	.45	65980.70	15226.32
18. 3 CONDUIT, CABLES, TRAYS FT		100000.0	1.32	1.36	132000.00	136000.00
18. 4 ISOLATED PHASE BUS FT			510.00	450.00		
18. 5 LIGHTING & COMMUN KWE		33836.3	.35	.43	11842.69	14549.59
PERCENT TOTAL DIRECT COST IN ACCOUNT 18 =			4.911 ACCOUNT TOTAL.\$		219297.54	166926.34
CONTROL, INSTRUMENTATION						
19. 1 COMPUTER	EACH	1.0	100000.00	3000.00	100000.00	3000.00
19. 2 OTHER CONTROLS	EACH	1.0	83000.00	50000.00	83000.00	50000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 19 =			3.001 ACCOUNT TOTAL.\$		183000.00	53000.00
PROCESS WASTE SYSTEMS						
20. 1 BOTTOM ASH	TPH	.0	.00	.00	.00	.00
20. 2 DRY ASH	TPH	.0	.00	.00	.00	.00
20. 3 WET SLURRY	TPH	.0	.00	.00	.00	.00
20. 4 ONSITE DISPOSAL	ACRE		7676.49	11070.93		
PERCENT TOTAL DIRECT COST IN ACCOUNT 20 =			.000 ACCOUNT TOTAL.\$.00	.00
STACK GAS CLEANING						
21. 1 PRECIPITATOR	EACH	.0	.00	.00	.00	.00
21. 2 SCRUBBER	KWE	.0	.00	.00	.00	.00
21. 3 MISC STEEL & DUCTS		.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 21 =			.000 ACCOUNT TOTAL.\$.00	.00
TOTAL DIRECT COSTS.\$					6870522.25	994263.16

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Table A 13.4.2 PHOSPHORIC ACID FUEL CELL SYSTEM CCGI OF ELECTRICITY, MILLS/KW·HR
 PARAMETRIC POINT NO. 1

ACCOUNT	RATE, PERCENT	LABOR RATE, \$/HR	21.5F
TOTAL DIRECT COSTS, \$	6.00	15.00	8927138.
INDIRECT COSTS, \$	74.3713	7864785.	10228500.
PROF & OWNER COSTS, \$	2670223.	507074.	7100975.
CONTINGENCY COST, \$	394355.	629183.	3999233.
SUB TOTAL, \$	374499.	353915.	11025585.
ESCALATION COST, \$	849500.	3354958.	572894.
INTREST DURING CONST, \$	434055.	354500.	59597.
TOTAL CAPITALIZATION, \$	908372.	547132.	12307277.
TOTAL OF ELEC-CAPITAL	1043649.	1043649.	15.70197
COST OF ELEC-FUEL	11.51214	14.16395	15.13205
COST OF ELEC-OP & MAIN	10.89330	25.01226	10.89330
TOTAL COST OF ELEC	42.00689	50.07551	12.60753

ACCOUNT	RATE, PERCENT	CONTINGENCY, PERCENT	20.00
TOTAL DIRECT COSTS, \$	5.00	4.50	7864785.
INDIRECT COSTS, \$	7864785.	7864785.	507074.
PROF & OWNER COSTS, \$	507074.	507074.	629183.
CONTINGENCY COST, \$	629183.	629183.	333239.
SUB TOTAL, \$	532239.	353915.	533915.
ESCALATION COST, \$	9001042.	9354958.	3354958.
INTREST DURING CONST, \$	514135.	547132.	549437.
TOTAL CAPITALIZATION, \$	526433.	1043649.	1043649.
TOTAL OF ELEC-CAPITAL	1043649.	14.16395	16.01226
COST OF ELEC-FUEL	15.61228	25.01226	10.89330
COST OF ELEC-OP & MAIN	10.89330	10.89330	50.13550
TOTAL COST OF ELEC	48.24379	50.07551	51.92199

ACCOUNT	RATE, PERCENT	ESCALATION RATE, PERCENT	10.00
TOTAL DIRECT COSTS, \$	5.00	8.00	10.00
INDIRECT COSTS, \$	7864785.	7864785.	7864785.
PROF & OWNER COSTS, \$	507074.	507074.	507074.
CONTINGENCY COST, \$	629183.	629183.	629183.
SUB TOTAL, \$	353915.	353915.	353915.
ESCALATION COST, \$	9354958.	9354958.	3354958.
INTREST DURING CONST, \$	534350.	657580.	821842.
TOTAL CAPITALIZATION, \$	547132.	552921.	580621.
TOTAL OF ELEC-CAPITAL	1043649.	14.34513	14.34513
COST OF ELEC-FUEL	15.61228	25.01226	25.01226
COST OF ELEC-OP & MAIN	10.89330	10.89330	10.89330
TOTAL COST OF ELEC	49.90028	50.25068	50.48416

ACCOUNT	RATE, PERCENT	INT DURING CONST, PERCENT	10.00
TOTAL DIRECT COSTS, \$	5.00	10.00	10.00
INDIRECT COSTS, \$	7864785.	7864785.	7864785.
PROF & OWNER COSTS, \$	507074.	507074.	507074.
CONTINGENCY COST, \$	629183.	629183.	629183.
SUB TOTAL, \$	353915.	353915.	353915.
ESCALATION COST, \$	9354958.	9354958.	3354958.
INTREST DURING CONST, \$	534350.	547132.	534350.
TOTAL CAPITALIZATION, \$	547132.	1043649.	681641.
TOTAL OF ELEC-CAPITAL	1043649.	14.34513	14.34513
COST OF ELEC-FUEL	15.61228	25.01226	25.01226
COST OF ELEC-OP & MAIN	10.89330	10.89330	10.89330
TOTAL COST OF ELEC	49.92855	50.07551	50.48416

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Table PHOSPHORIC ACID FUEL CELL SYSTEM COST OF ELECTRICITY, MILLS/KW·HR
 A 13.4.2 Continued PARAMETRIC POINT NO. 1

ACCOUNT	RATE, PERCENT	FIXED CHARGE RATE, PCT	25.00
TOTAL DIRECT COSTS,\$	14.40	18.00	7864785.
INDIRECT COST,\$	51.0	7864785.	507074.
PROF & OWNER COSTS,\$	8.0	507074.	629183.
CONTINGENCY COST,\$	4.5	629183.	353915.
SUB TOTAL,\$	0	353915.	9354958.
ESCALATION COST,\$	6.5	9354958.	534350.
INTEREST DURING CONST,\$	10.0	534350.	547132.
TOTAL CAPITALIZATION,\$	0	547132.	10436439.
COST OF ELEC-CAPITAL	25.0	10436439.	14.16995
COST OF ELEC-FUEL	0	14.16995	25.01226
COST OF ELEC-OP & MAIN	0	25.01226	10.89330
TOTAL COST OF ELEC	0	10.89330	47.24152

ACCOUNT	RATE, PERCENT	FUEL COST, \$/10 ⁶ BTU	2 C8
TOTAL DIRECT COSTS,\$	1.50	4.00	7864785.
INDIRECT COST,\$	7864785.	7864785.	507074.
PROF & OWNER COSTS,\$	51.0	507074.	629183.
CONTINGENCY COST,\$	4.5	629183.	353915.
SUB TOTAL,\$	0	353915.	9354958.
ESCALATION COST,\$	6.5	9354958.	534350.
INTEREST DURING CONST,\$	10.0	534350.	547132.
TOTAL CAPITALIZATION,\$	0	547132.	10436439.
COST OF ELEC-CAPITAL	18.0	10436439.	14.16995
COST OF ELEC-FUEL	0	14.16995	38.48040
COST OF ELEC-OP & MAIN	0	38.48040	10.89330
TOTAL COST OF ELEC	0	10.89330	63.54365

ACCOUNT	RATE, PERCENT	CAPACITY FACTOR, PERCENT	80.00
TOTAL DIRECT COSTS,\$	12.00	50.00	7864785.
INDIRECT COST,\$	7864785.	7864785.	507074.
PROF & OWNER COSTS,\$	51.0	507074.	629183.
CONTINGENCY COST,\$	8.0	629183.	353915.
SUB TOTAL,\$	4.5	353915.	9354958.
ESCALATION COST,\$	6.5	9354958.	534350.
INTEREST DURING CONST,\$	10.0	534350.	547132.
TOTAL CAPITALIZATION,\$	0	547132.	10436439.
COST OF ELEC-CAPITAL	18.0	10436439.	14.16995
COST OF ELEC-FUEL	0	14.16995	25.01226
COST OF ELEC-OP & MAIN	0	25.01226	11.16484
TOTAL COST OF ELEC	0	11.16484	54.59804

Table A 13.4.3 PHOSPHORIC ACID FUEL CELL SYSTEM

ACCOUNT NO	AUX POWER, MWE	PERC PLANT POW	OPERATION COST	MAINTENANCE COST
11	.00000	.00000	221.00000	120000.00000
18	.11700	100.00000	.00000	.00000
TOTALS	.11700	50251	232.55377	120000.00000
PHOSPHORIC ACID FUEL CELL SYSTEM				
NOMINAL POWER, MWE	23.4000			23.2830
NET HEAT RATE, BTU/KW-HR	9571.9998			9620.1001
SI TURB HEAT RATE CHANGE				
CONDENSER				
DESIGN PRESSURE, IN HG A	2.0000			.0000
NUMBER OF TUBES/SHELL	.0000			.0000
U, BTU/HR-FI2-F	.0000			.0000
HEAT REJECTION RANGE, F	.0000			.0000
DESIGN TEMP, F	.0000			.0000
OFF DESIGN PRES, IN HG A	.0000			.0000
	23.400	2	.000	9572.000
6	.000	7	.000	.000
11	.000	12	1.000	2.000
16	.000	17	5.000	.000
21	.000	22	.000	.000
26	.000	27	.000	.000
31	1.000	32	.000	.200
36	100000.000	37	1.000	1.000
41	.000	42	300.000	83000.000
46	.000	47	1.000	1.000
51	.000	52	5.350	.000
	.000	53	.000	.000
6	1.000	7	.000	.000
11	.000	12	.000	.000
16	.000	17	.000	.000
21	.000	22	.000	.000
26	.000	27	.000	.000
31	23000.000	32	.000	.000
36	.000	37	.000	.000
41	.000	42	.000	.000
46	.000	47	.000	.000
51	.200	52	.000	.000
		53	.000	.000
		54	.000	.000
		55	.000	.000
		56	.000	.000
		57	.000	.000
		58	.000	.000
		59	.000	.000
		60	.000	.000
		61	.000	.000
		62	.000	.000
		63	.000	.000
		64	.000	.000
		65	.000	.000
		66	.000	.000
		67	.000	.000
		68	.000	.000
		69	.000	.000
		70	.000	.000
		71	.000	.000
		72	.000	.000
		73	.000	.000
		74	.000	.000
		75	.000	.000
		76	.000	.000
		77	.000	.000
		78	.000	.000
		79	.000	.000
		80	.000	.000
		81	.000	.000
		82	.000	.000
		83	.000	.000
		84	.000	.000
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		89	.000	.000
		90	.000	.000
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		92	.000	.000
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		102	.000	.000
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		114	.000	.000
		115	.000	.000
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		118	.000	.000
		119	.000	.000
		120	.000	.000
		121	.000	.000
		122	.000	.000
		123	.000	.000
		124	.000	.000
		125	.000	.000
		126	.000	.000
		127	.000	.000
		128	.000	.000
		129	.000	.000
		130	.000	.000
		131	.000	.000
		132	.000	.000
		133	.000	.000
		134	.000	.000
		135	.000	.000
		136	.000	.000
		137	.000	.000
		138	.000	.000
		139	.000	.000
		140	.000	.000
		141	.000	.000
		142	.000	.000
		143	.000	.000
		144	.000	.000
		145	.000	.000
		146	.000	.000
		147	.000	.000
		148	.000	.000
		149	.000	.000
		150	.000	.000
		151	.000	.000
		152	.000	.000
		153	.000	.000
		154	.000	.000
		155	.000	.000
		156	.000	.000
		157	.000	.000
		158	.000	.000
		159	.000	.000
		160	.000	.000
		161	.000	.000
		162	.000	.000
		163	.000	.000
		164	.000	.000
		165	.000	.000
		166	.000	.000
		167	.000	.000
		168	.000	.000
		169	.000	.000
		170	.000	.000
		171	.000	.000
		172	.000	.000
		173	.000	.000
		174	.000	.000
		175	.000	.000
		176	.000	.000
		177	.000	.000
		178	.000	.000
		179	.000	.000
		180	.000	.000
		181	.000	.000
		182	.000	.000
		183	.000	.000
		184	.000	.000
		185	.000	.000
		186	.000	.000
		187	.000	.000
		188	.000	.000
		189	.000	.000
		190	.000	.000
		191	.000	.000
		192	.000	.000
		193	.000	.000
		194	.000	.000
		195	.000	.000
		196	.000	.000
		197	.000	.000
		198	.000	.000
		199	.000	.000
		200	.000	.000

Table A 13.4.4 ALKALINE FUEL CELL SYSTEM ACCOUNT LISTING
 PARAMETRIC POINT NO. 1

ACCOUNT NO. & NAME	UNIT	AMOUNT	MAT \$/UNIT	INS \$/UNIT	MAT COST,\$	INS COST,\$
SITE DEVELOPMENT						
1. 1 LAND COST	ACRE	1.5	100000.00	.00	150000.00	.00
1. 2 CLEARING LAND	ACRE	1.5	.00	500.00	.00	299.97
1. 3 GRADING LAND	ACRE	1.5	.00	3000.00	.00	4500.00
1. 4 ACCESS RAILROAD	MILE	.0	115000.00	110000.00	.00	.00
1. 5 LOOP RAILROAD TRACK	MILE	.0	120000.00	70000.00	.00	.00
1. 6 SIDING R.R. TRACK	MILE	.0	125000.00	80000.00	.00	.00
1. 7 OTHER SITE COSTS	ACRE	.0	.00	.00	3736.39	3736.39
PERCENT TOTAL DIRECT COST IN ACCOUNT		1 =	1.536	ACCOUNT TOTAL,\$	153736.39	8536.36
EXCAVATION & PILING						
2. 1 COMMON EXCAVATION	YD3	9600.0	.00	3.00	.00	28800.00
2. 2 PILING	FT	.0	6.50	8.50	.00	28800.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		2 =	.281	ACCOUNT TOTAL,\$.00	28800.00
PLANT ISLAND CONCRETE						
3. 1 PLANT IS. CONCRETE	YD3	3200.0	70.00	80.00	224000.00	256000.00
3. 2 SPECIAL STRUCTURES	YD3	.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		3 =	4.691	ACCOUNT TOTAL,\$	224000.00	256000.00
HEAT REJECTION SYSTEM						
4. 1 COOLING TOWERS	EACH	.0	.00	.00	.00	.00
4. 2 CIRCULATING H2O SYS	EACH	1.0	.00	.00	.00	.00
4. 3 SURFACE CONDENSER	FT2	.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		4 =	.000	ACCOUNT TOTAL,\$.00	.00
STRUCTURAL FEATURES						
5. 1 STAT. STRUCTURAL ST. TON	TON	.0	650.00	175.00	.00	.00
5. 2 SILOS & BUNKERS	TPH	.0	1800.00	750.00	.00	.00
5. 3 CHIMNEY	FT	.0	.00	.00	.00	.00
5. 4 STRUCTURAL FEATURES EACH	EACH	1.0	170000.00	30000.00	170000.00	30000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		5 =	1.954	ACCOUNT TOTAL,\$	170000.00	30000.00
BUILDINGS						
6. 1 STATION BUILDINGS	FT3	.0	.16	.16	.00	.00
6. 2 ADMINISTRATION	FT2	5000.0	16.00	14.00	80000.00	70000.00
6. 3 WAREHOUSE & SHOP	FT2	.0	12.00	8.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		6 =	1.466	ACCOUNT TOTAL,\$	80000.00	70000.00
FUEL HANDLING & STORAGE						
7. 1 COAL HANDLING SYS	TPH	.0	.00	.00	.00	.00
7. 2 DOLOMITE HAND. SYS	TPH	.0	.00	.00	.00	.00
7. 3 FUEL OIL HAND. SYS	GAL	.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		7 =	.000	ACCOUNT TOTAL,\$.00	.00

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Table A 13.4.4 ALKALINE FUEL CELL SYSTEM ACCOUNT LISTING
Continued

ACCOUNT NO. & NAME	UNIT	AMOUNT	MAT \$/UNIT	INS \$/UNIT	MAT COST,\$	INS COST,\$
FUEL PROCESSING						
1	COAL GASIFIER	EACH	.00	.00	.00	.00
2	DESULFURIZER	EACH	.00	.00	.00	.00
3	VENTURI SCRUBBER	EACH	.00	.00	.00	.00
4	COAL PRETREATMENT	EACH	115000.00	115000.00	500000.00	115000.00
5	REFORMER	EACH	100000.00	115000.00	500000.00	210000.00
6	SHIFT CONVERTER	EACH	350000.00	210000.00	840000.00	280000.00
7	CO2 SCRUBBER	EACH	420000.00	141000.00	560000.00	280000.00
8	BLOWER	EACH	1.00	1.00	1355000.00	416000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 9 = 17.777 ACCOUNT TOTAL,\$						
FIRING SYSTEM						
1	PERCENT TOTAL DIRECT COST IN ACCOUNT 9 =	.00	.00	.00	.00	.00
VAPOR GENERATOR (FIRED)						
1	PERCENT TOTAL DIRECT COST IN ACCOUNT 10 =	.00	.00	.00	.00	.00
ENERGY CONVERTER						
1	FUEL CELL SYSTEM	EACH	1.00	125000.00	500000.00	125000.00
2	STM TURBINE GENERATOR	EACH	.00	.00	.00	.00
3	OXYGEN PLANT	EACH	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 11 = 0.004 ACCOUNT TOTAL,\$						
COUPLING HEAT EXCHANGER						
1	HEAT REC STEAM GEN	EACH	1.00	74000.00	28000.00	20000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 12 = .915 ACCOUNT TOTAL,\$						
HEAT RECOVERY HEAT EXCH.						
1	RECUPERATOR	EACH	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 13 = .000 ACCOUNT TOTAL,\$						
WATER TREATMENT						
1	DEMINERALIZER	GPM	.00	2500.00	700.00	.00
2	CONDENSATE POLISHING	KWE	1.00	1.25	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 14 = .000 ACCOUNT TOTAL,\$						
POWER CONDITIONING						
1	POWER CONDITIONING EQ	EA	1.00	170000.00	70000.00	70000.00
2	SID TRANSFORMER	KVA	1.00	130000.00	130000.00	75000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 15 = 14.019 ACCOUNT TOTAL,\$						
AUXILIARY MECH EQUIPMENT						
1	BOILER FEED PUMP	SPH-KWE	.00	.55	.00	.00
2	OTHER PUMPS	KWE	.00	.88	.00	.00
3	MISC SERVICE SYS	KWE	31041.7	1.17	57254.79	23244.44
4	AUXILIARY BOILER	SPH	1.00	4.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 16 = .531 ACCOUNT TOTAL,\$						

Table A 13.4.4 ALKALINE FUEL CELL SYSTEM ACCOUNT LISTING
 Continued PARAMETRIC POINT NO. 1

ACCOUNT NO. & NAME,	UNIT	AMOUNT	MAT \$/UNIT	INS \$/UNIT	MAT COST,\$	INS COST,\$
PIPE & FITTINGS						
17. 1 CONVENTIONAL PIPING ION		10.0	3000.00	1900.00	30000.00	13000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 17 =		.469	ACCOUNT TOTAL,\$		30000.00	18000.00
AUXILIARY ELEC EQUIPMENT						
18. 1 MISC MOTORS, ETC		6368.3	1.40	.17	8915.68	1082.62
18. 2 SWITCHGEAR & MCC PAN KWE		31841.7	1.95	.45	62091.32	14328.77
18. 3 CONDUIT, CABLES, TRAYS FT		100000.0	1.32	1.36	132000.00	136000.00
18. 4 ISOLATED PHASE BUS FT		.0	510.00	450.00	.00	.00
18. 5 LIGHTING & COMMUN KWE		31841.7	.35	.43	11144.60	13691.93
PERCENT TOTAL DIRECT COST IN ACCOUNT 18 =		3.706	ACCOUNT TOTAL,\$		214151.59	165103.31
CONTROL, INSTRUMENTATION						
19. 1 COMPUTER EACH		1.0	100000.00	3000.00	100000.00	3000.00
19. 2 OTHER CONTROLS EACH		1.0	83000.00	50000.00	83000.00	50000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 19 =		2.306	ACCOUNT TOTAL,\$		183000.00	53000.00
PROCESS WASTE SYSTEMS						
20. 1 BOTTOM ASH TPH		.0	.00	.00	.00	.00
20. 2 DRY ASH TPH		.0	.00	.00	.00	.00
20. 3 WET SLURRY TPH		.0	.00	.00	.00	.00
20. 4 ONSITE DISPOSAL ACRE		.0	7676.49	11070.89	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 20 =		.000	ACCOUNT TOTAL,\$.00	.00
STACK GAS CLEANING						
21. 1 PRECIPITATOR EACH		.0	.00	.00	.00	.00
21. 2 SCRUBBER KWE		.0	.00	.00	.00	.00
21. 3 MISC STEEL & DUCYS		.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 21 =		.000	ACCOUNT TOTAL,\$.00	.00
TOTAL DIRECT COSTS,\$					8942142.62	1290684.11

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Table ALKALINE FUEL CELL SYSTEM COST OF ELECTRICITY • MILLS/KW • HR
A 13.4.5 PARAMETRIC POINT NO. 1

ACCOUNT	RATE, PERCENT	LABOR RATE, \$/HR	
TOTAL DIRECT COSTS, \$	100.00	10232827.0	
INDIRECT COSTS, \$	11.00	11256033.9	
PROF & OWNER COSTS, \$	3.00	11580033.9	
CONTINGENCY COST, \$	4.00	12275127.0	
SUB TOTAL, \$	118.00	12469827.0	
ESCALATION COST, \$	0.00	520202.0	
INTEREST DURING CONST, \$	10.00	880586.0	
TOTAL CAPITALIZATION, \$	10.00	13045139.0	
COST OF ELEC-CAPITAL	18.00	748189.0	
COST OF ELEC-FUEL	0.00	838695.0	
COST OF ELEC-OP & MAINT	0.00	763013.0	
TOTAL COST OF ELEC	0.00	15997070.0	
		23.00	103331.0
		23.00	31116.0
		17.00	83717.0
		62.00	16693.0

ACCOUNT	RATE, PERCENT	CONTINGENCY, PERCENT	
TOTAL DIRECT COSTS, \$	100.00	10232827.0	
INDIRECT COSTS, \$	11.00	11256033.9	
PROF & OWNER COSTS, \$	3.00	11580033.9	
CONTINGENCY COST, \$	4.00	12275127.0	
SUB TOTAL, \$	118.00	12469827.0	
ESCALATION COST, \$	0.00	520202.0	
INTEREST DURING CONST, \$	10.00	880586.0	
TOTAL CAPITALIZATION, \$	10.00	13045139.0	
COST OF ELEC-CAPITAL	18.00	748189.0	
COST OF ELEC-FUEL	0.00	838695.0	
COST OF ELEC-OP & MAINT	0.00	763013.0	
TOTAL COST OF ELEC	0.00	15997070.0	
		25.00	31091.0
		19.00	8583.0
		23.00	31116.0
		17.00	83717.0
		50.00	75553.0

ACCOUNT	RATE, PERCENT	ESCALATION RATE, PERCENT	
TOTAL DIRECT COSTS, \$	100.00	10232827.0	
INDIRECT COSTS, \$	11.00	11256033.9	
PROF & OWNER COSTS, \$	3.00	11580033.9	
CONTINGENCY COST, \$	4.00	12275127.0	
SUB TOTAL, \$	118.00	12469827.0	
ESCALATION DURING CONST, \$	12.00	65468.0	
TOTAL CAPITALIZATION, \$	10.00	719313.0	
COST OF ELEC-CAPITAL	18.00	13744960.0	
COST OF ELEC-FUEL	0.00	19.84965.0	
COST OF ELEC-OP & MAINT	0.00	23.31116.0	
TOTAL COST OF ELEC	0.00	17.83717.0	
		60.00	99797.0
		10.00	10232827.0
		65.00	249.0
		81.00	628.0
		46.00	477.0
		121.00	179.0
		69.00	154.0
		71.00	115.0
		19.00	80725.0
		23.00	31116.0
		17.00	83717.0
		60.00	75553.0

ACCOUNT	RATE, PERCENT	INT DURING CONST, PERCENT	
TOTAL DIRECT COSTS, \$	100.00	10232827.0	
INDIRECT COSTS, \$	11.00	11256033.9	
PROF & OWNER COSTS, \$	3.00	11580033.9	
CONTINGENCY COST, \$	4.00	12275127.0	
SUB TOTAL, \$	118.00	12469827.0	
ESCALATION DURING CONST, \$	6.00	695154.0	
TOTAL CAPITALIZATION, \$	15.00	868770.0	
COST OF ELEC-CAPITAL	18.00	13744960.0	
COST OF ELEC-FUEL	0.00	19.84965.0	
COST OF ELEC-OP & MAINT	0.00	23.31116.0	
TOTAL COST OF ELEC	0.00	17.83717.0	
		50.00	75553.0
		10.00	10232827.0
		65.00	249.0
		81.00	628.0
		46.00	477.0
		121.00	179.0
		69.00	154.0
		71.00	115.0
		19.00	80725.0
		23.00	31116.0
		17.00	83717.0
		60.00	75553.0

Table ALKALINE FUEL CELL SYSTEM COST OF ELECTRICITY, MILLS/KW·HR
 A 13.4.5 Continued PARAMETRIC POINT NO. 1

ACCOUNT	RATE, PERCENT	FIXED CHARGE RATE, PCY	21.50	25.00
TOTAL DIRECT COSTS, \$	10.00	18.00	10232827.	10232827.
INDIRECT COST, \$	51.00	658249.	658249.	658249.
PROF & OWNER COSTS, \$	8.00	818626.	818626.	818626.
CONTINGENCY COST, \$	4.50	460477.	460477.	460477.
SUB TOTAL, \$	0.00	12170179.	12170179.	12170179.
ESCALATION COST, \$	6.50	695154.	695154.	695154.
INTEREST DURING CONST, \$	10.00	711782.	711782.	711782.
TOTAL CAPITALIZATION, \$	0.00	13577115.	13577115.	13577115.
COST OF ELEC-CAPITAL	25.00	10.83292.	23.52870.	27.23230.
COST OF ELEC-FUEL	0.00	23.31116.	23.31116.	23.31116.
COST OF ELEC-OP & MAIN	0.00	17.83717.	17.83717.	17.83717.
TOTAL COST OF ELEC	0.00	56.83413.	64.67703.	68.38062.

ACCOUNT	RATE, PERCENT	FUEL COST, \$/10**6 BTU	2.08	3.12
TOTAL DIRECT COSTS, \$	10.00	10232827.	10232827.	10232827.
INDIRECT COST, \$	51.00	658249.	658249.	658249.
PROF & OWNER COSTS, \$	8.00	818626.	818626.	818626.
CONTINGENCY COST, \$	4.50	460477.	460477.	460477.
SUB TOTAL, \$	0.00	12170179.	12170179.	12170179.
ESCALATION COST, \$	6.50	695154.	695154.	695154.
INTEREST DURING CONST, \$	10.00	711782.	711782.	711782.
TOTAL CAPITALIZATION, \$	0.00	13577115.	13577115.	13577115.
COST OF ELEC-CAPITAL	18.00	19.60725.	19.60725.	19.60725.
COST OF ELEC-FUEL	0.00	13.44874.	18.64892.	27.97339.
COST OF ELEC-OP & MAIN	0.00	17.83717.	17.83717.	17.83717.
TOTAL COST OF ELEC	0.00	60.75558.	56.09335.	65.41781.

ACCOUNT	RATE, PERCENT	CAPACITY FACTOR, PERCENT	50.00	80.00
TOTAL DIRECT COSTS, \$	12.00	10232827.	10232827.	10232827.
INDIRECT COST, \$	51.00	658249.	658249.	658249.
PROF & OWNER COSTS, \$	8.00	818626.	818626.	818626.
CONTINGENCY COST, \$	4.50	460477.	460477.	460477.
SUB TOTAL, \$	0.00	12170179.	12170179.	12170179.
ESCALATION COST, \$	6.50	695154.	695154.	695154.
INTEREST DURING CONST, \$	10.00	711782.	711782.	711782.
TOTAL CAPITALIZATION, \$	0.00	13577115.	13577115.	13577115.
COST OF ELEC-CAPITAL	18.00	28.32159.	19.60725.	15.93089.
COST OF ELEC-FUEL	0.00	23.31116.	23.31116.	23.31116.
COST OF ELEC-OP & MAIN	0.00	18.43622.	17.83717.	17.58445.
TOTAL COST OF ELEC	0.00	70.06896.	60.75558.	56.82650.

Table A 13.4.6 ALKALINE FUEL CELL SYSTEM

ACCOUNT NO	AUX POWER, MWE	PERC PLANT FOW	OPERATION COST	MAINTENANCE COST
11	.00000	.00000	350.00000	168000.00000
18	.11000	100.00000	.00000	.00000
TOTALS	.11000	.50251	360.95094	168000.00000
NOMINAL POWER, MWE	22.0000		BASE CASE INPUT	21.8900
NOM HEAT RATE, BTU/KW-HR	8920.9999		NET POWER, MWE	3965.8289
ST TURB HEAT RATE CHANGE	.0000		NET HEAT RATE, BTU/KW-HR	
CONDENSER			NUMBER OF SHELLS	.0000
DESIGN PRESSURE, IN HG A	2.0000		TUBE LENGTH, FT	.0000
NUMBER OF TUBES/SHELL	.0000		TERMINAL TEMP DIFF, F	.0000
U, BTU/HR-FI2-F			APPROACH, F	.0000
HEAT REJECTION	.0000		OFF DESIGN TEMP, F	.0000
DESIGN TEMP, F	.0000		LP TURBINE BLADE LEN, IN	.0000
RANGE, F	.0000			
OFF DESIGN PRES, IN HG A				
1	22.000	3	.000	8921.000
6	.000	8	.000	.000
11	.000	13	1.000	2.000
15	.000	18	5.000	.000
21	.000	23	.000	.000
26	.000	28	.000	.000
31	1.000	33	.000	.000
36	100000.000	38	1.000	1.000
41	30000.000	43	300.000	83000.000
46	.000	48	1.000	1.000
51	.000	53	.000	.000
6	1.000	8	1.000	.000
11	.000	13	.000	.000
16	.000	18	115000.000	56000.000
21	420000.000	23	500000.000	125000.000
26	.000	28	.000	1.000
31	.000	33	.000	.000
36	20000.000	38	.000	.000
41	.000	43	75000.000	.000
46	.000	48	.000	.000
51	.200	53	.000	.000
6	1.000	8	1.000	.000
11	.000	13	.000	.000
16	.000	18	115000.000	56000.000
21	420000.000	23	500000.000	125000.000
26	.000	28	.000	1.000
31	.000	33	.000	.000
36	20000.000	38	.000	.000
41	.000	43	75000.000	.000
46	.000	48	.000	.000
51	.200	53	.000	.000
6	1.000	8	1.000	.000
11	.000	13	.000	.000
16	.000	18	115000.000	56000.000
21	420000.000	23	500000.000	125000.000
26	.000	28	.000	1.000
31	.000	33	.000	.000
36	20000.000	38	.000	.000
41	.000	43	75000.000	.000
46	.000	48	.000	.000
51	.200	53	.000	.000
6	1.000	8	1.000	.000
11	.000	13	.000	.000
16	.000	18	115000.000	56000.000
21	420000.000	23	500000.000	125000.000
26	.000	28	.000	1.000
31	.000	33	.000	.000
36	20000.000	38	.000	.000
41	.000	43	75000.000	.000
46	.000	48	.000	.000
51	.200	53	.000	.000
6	1.000	8	1.000	.000
11	.000	13	.000	.000
16	.000	18	115000.000	56000.000
21	420000.000	23	500000.000	125000.000
26	.000	28	.000	1.000
31	.000	33	.000	.000
36	20000.000	38	.000	.000
41	.000	43	75000.000	.000
46	.000	48	.000	.000
51	.200	53	.000	.000
6	1.000	8	1.000	.000
11	.000	13	.000	.000
16	.000	18	115000.000	56000.000
21	420000.000	23	500000.000	125000.000
26	.000	28	.000	1.000
31	.000	33	.000	.000
36	20000.000	38	.000	.000
41	.000	43	75000.000	.000
46	.000	48	.000	.000
51	.200	53	.000	.000
6	1.000	8	1.000	.000
11	.000	13	.000	.000
16	.000	18	115000.000	56000.000
21	420000.000	23	500000.000	125000.000
26	.000	28	.000	1.000
31	.000	33	.000	.000
36	20000.000	38	.000	.000
41	.000	43	75000.000	.000
46	.000	48	.000	.000
51	.200	53	.000	.000
6	1.000	8	1.000	.000
11	.000	13	.000	.000
16	.000	18	115000.000	56000.000
21	420000.000	23	500000.000	125000.000
26	.000	28	.000	1.000
31	.000	33	.000	.000
36	20000.000	38	.000	.000
41	.000	43	75000.000	.000
46	.000	48	.000	.000
51	.200	53	.000	.000
6	1.000	8	1.000	.000
11	.000	13	.000	.000
16	.000	18	115000.000	56000.000
21	420000.000	23	500000.000	125000.000
26	.000	28	.000	1.000
31	.000	33	.000	.000
36	20000.000	38	.000	.000
41	.000	43	75000.000	.000
46	.000	48	.000	.000
51	.200	53	.000	.000
6	1.000	8	1.000	.000
11	.000	13	.000	.000
16	.000	18	115000.000	56000.000
21	420000.000	23	500000.000	125000.000
26	.000	28	.000	1.000
31	.000	33	.000	.000
36	20000.000	38	.000	.000
41	.000	43	75000.000	.000
46	.000	48	.000	.000
51	.200	53	.000	.000
6	1.000	8	1.000	.000
11	.000	13	.000	.000
16	.000	18	115000.000	56000.000
21	420000.000	23	500000.000	125000.000
26	.000	28	.000	1.000
31	.000	33	.000	.000
36	20000.000	38	.000	.000
41	.000	43	75000.000	.000
46	.000	48	.000	.000
51	.200	53	.000	.000
6	1.000	8	1.000	.000
11	.000	13	.000	.000
16	.000	18	115000.000	56000.000
21	420000.000	23	500000.000	125000.000
26	.000	28	.000	1.000
31	.000	33	.000	.000
36	20000.000	38	.000	.000
41	.000	43	75000.000	.000
46	.000	48	.000	.000
51	.200	53	.000	.000
6	1.000	8	1.000	.000
11	.000	13	.000	.000
16	.000	18	115000.000	56000.000
21	420000.000	23	500000.000	125000.000
26	.000	28	.000	1.000
31	.000	33	.000	.000
36	20000.000	38	.000	.000
41	.000	43	75000.000	.000
46	.000	48	.000	.000
51	.200	53	.000	.000
6	1.000	8	1.000	.000
11	.000	13	.000	.000
16	.000	18	115000.000	56000.000
21	420000.000	23	500000.000	125000.000
26	.000	28	.000	1.000
31	.000	33	.000	.000
36	20000.000	38	.000	.000
41	.000	43	75000.000	.000
46	.000	48	.000	.000
51	.200	53	.000	.000
6	1.000	8	1.000	.000
11	.000	13	.000	.000
16	.000	18	115000.000	56000.000
21	420000.000	23	500000.000	125000.000
26	.000	28	.000	1.000
31	.000	33	.000	.000
36	20000.000	38	.000	.000
41	.000	43	75000.000	.000
46	.000	48	.000	.000
51	.200	53	.000	.000
6	1.000	8	1.000	.000
11	.000	13	.000	.000
16	.000	18	115000.000	56000.000
21	420000.000	23	500000.000	125000.000
26	.000	28	.000	1.000
31	.000	33	.000	.000
36	20000.000	38	.000	.000
41	.000	43	75000.000	.000
46	.000	48	.000	.000
51	.200	53	.000	.000
6	1.000	8	1.000	.000
11	.000	13	.000	.000
16	.000	18	115000.000	56000.000
21	420000.000	23	500000.000	125000.000
26	.000	28	.000	1.000
31	.000	33	.000	.000
36	20000.000	38	.000	.000
41	.000	43	75000.000	.000
46	.000	48	.000	.000
51	.200	53	.000	.000
6	1.000	8	1.000	.000
11	.000	13	.000	.000
16	.000	18	115000.000	56000.000
21	420000.000	23	500000.000	125000.000
26	.000	28	.000	1.000
31	.000	33	.000	.000
36	20000.000	38	.000	.000
41	.000	43	75000.000	.000
46	.000	48	.000	.000
51	.200	53	.000	.000
6	1.000	8	1.000	.000
11	.000	13	.000	.000
16	.000	18	115000.000	56000.000
21	420000.000	23	500000.000	125000.000
26	.000	28	.000	1.000
31	.000	33	.000	.000
36	20000.000	38	.000	.000
41	.000	43	75000.000	.000
46	.000	48	.000	.000
51	.200	53	.000	.000
6	1.000	8	1.000	.000
11	.000	13	.000	.000
16	.000	18	115000.000	56000.000
21	420000.000	23	500000.000	125000.000
26	.000	28	.000	1.000
31	.000	33	.000	.000
36	20000.000	38	.000	.00

Table
A 13.4.7

MOLTEN CARBONATE FUEL CELL SYSTEM ACCOUNT LISTING

PARAMETRIC POINT NO. 1

ACCOUNT NO. & NAME UNIT AMOUNT MAT \$/UNIT INS \$/UNIT MAT COST \$ INS COST \$

SITE DEVELOPMENT

1.	1	LAND COST	ACRE	1.5	100000.00	.00	150000.00	.00	.00
1.	2	CLEARING LAND	ACRE	1.5	.00	800.00	.00	.00	299.97
1.	3	GRADING LAND	ACRE	1.5	.00	3000.00	.00	.00	4500.00
1.	4	ACCESS RAILROAD	MILE	.0	115000.00	.00	.00	.00	.00
1.	5	LOOP RAILROAD TRACK	MILE	.0	120000.00	.00	.00	.00	.00
1.	6	SIDING R.R. TRACK	MILE	.0	125000.00	.00	.00	.00	.00
1.	7	OTHER SITE COSTS	ACRE	.0	.00	80000.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 1 =				1.876	ACCOUNT TOTAL \$.00	3736.39	.00	3736.39
PERCENT TOTAL DIRECT COST IN ACCOUNT 2 =				1.876	ACCOUNT TOTAL \$.00	152776.39	.00	8536.36

EXCAVATION & PILING

2.	1	COMMON EXCAVATION	YD3	9600.0	.00	3.00	.00	.00	28800.00
2.	2	PILING	FT	.0	6.50	8.50	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 2 =				.333	ACCOUNT TOTAL \$.00	.00	.00	28800.00

PLANT ISLAND CONCRETE

3.	1	PLANT IS. CONCRETE	YD3	3200.0	70.00	80.00	224000.00	.00	256000.00
3.	2	SPECIAL STRUCTURES	YD3	.0	.00	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 3 =				5.549	ACCOUNT TOTAL \$.00	224000.00	.00	256000.00

HEAT REJECTION SYSTEM

4.	1	COOLING TOWERS	EACH	.0	.00	.00	.00	.00	.00
4.	2	CIRCULATING H2O SYS	EACH	1.0	.00	.00	.00	.00	.00
4.	3	SURFACE CONDENSER	FT2	.0	.00	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 4 =				.000	ACCOUNT TOTAL \$.00	.00	.00	.00

STRUCTURAL FEATURES

5.	1	SILOS & BUNKERS	TPH	.0	650.00	175.00	.00	.00	.00
5.	2	CHIMNEY	FT	.0	1800.00	750.00	.00	.00	.00
5.	3	STRUCTURAL FEATURES	EACH	1.0	170000.00	30000.00	170000.00	.00	30000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 5 =				2.312	ACCOUNT TOTAL \$.00	170000.00	.00	30000.00

BUILDINGS

6.	1	STATION BUILDINGS	FT3	.0	16.00	14.00	.00	.00	.00
6.	2	ADMINISTRATION	FT2	5000.0	16.00	14.00	30000.00	.00	70000.00
6.	3	WAREHOUSE & SHOP	FT2	.0	12.00	8.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 6 =				1.734	ACCOUNT TOTAL \$.00	80000.00	.00	70000.00

FUEL HANDLING & STORAGE

7.	1	COAL HANDLING SYS	TPH	.0	.00	.00	.00	.00	.00
7.	2	PELOMITE HAND. SYS	TPH	.0	.00	.00	.00	.00	.00
7.	3	FUEL OIL HAND. SYS	GAL	.0	.00	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 7 =				.000	ACCOUNT TOTAL \$.00	.00	.00	.00

Table MOLIEN CARBONATE FUEL CELL SYSTEM NO. 1 ACCOUNT LISTING
 A 13.4.7 Continued PARAMETRIC POINT NO. 1

ACCOUNT NO. & NAME	UNIT	AMOUNT	MAT \$/UNIT	INS \$/UNIT	MAT COST,\$	INS COST,\$
FUEL PROCESSING						
1	COAL GASIFIER	EACH	.00	.00	.00	.00
2	DESULFURIZER	EACH	.00	.00	.00	.00
3	VENTURI SCRUBBER	EACH	.00	.00	.00	.00
4	COAL PRETREATMENT	EACH	.00	.00	.00	.00
5	REFORMER	EACH	.00	.00	.00	.00
6	SHIFT CONVERTER	EACH	.00	.00	.00	.00
7	CO2 SCRUBBER	EACH	.00	.00	.00	.00
8	BLOWER	EACH	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		1.0	25000.00	7000.00	32000.00	3000.00
PERCENT ACCOUNT TOTAL,\$		1.0	.270	.700	29000.00	3000.00
FIRING SYSTEM						
1	PERCENT TOTAL DIRECT COST IN ACCOUNT	3.0	.000	.000	.000	.000
PERCENT ACCOUNT TOTAL,\$		3.0	.000	.000	.000	.000
VAPOR GENERATOR (FIRED)						
1	PERCENT TOTAL DIRECT COST IN ACCOUNT	10.0	.000	.000	.000	.000
PERCENT ACCOUNT TOTAL,\$		10.0	.000	.000	.000	.000
ENERGY CONVERTER						
1	FUEL CELL SYSTEM	EACH	1.0	475000.00	475000.00	200000.00
2	STEAM TURBINE GENERATOR	EACH	.0	.00	.00	.00
3	OXYGEN PLANT	EACH	.0	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		11.0	475.001	.000	475000.00	200000.00
PERCENT ACCOUNT TOTAL,\$		11.0	475.001	.000	475000.00	200000.00
COUPLING HEAT EXCHANGER						
1	HEAT REC STEAM GEN	EACH	1.0	102000.00	102000.00	51000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		12.0	1.769	.000	102000.00	51000.00
PERCENT ACCOUNT TOTAL,\$		12.0	1.769	.000	102000.00	51000.00
HEAT RECOVERY HEAT EXCH.						
1	RECUPERATOR	EACH	1.0	136000.00	136000.00	170000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		13.0	3.607	.000	136000.00	170000.00
PERCENT ACCOUNT TOTAL,\$		13.0	3.607	.000	136000.00	170000.00
WATER TREATMENT						
1	DEMINERALIZER	GPM	.0	2500.00	700.00	.00
2	CONDENSATE POLISHING	KWE	1.25	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		14.0	.000	.000	700.00	.00
PERCENT ACCOUNT TOTAL,\$		14.0	.000	.000	700.00	.00
POWER CONDITIONING						
1	POWER CONDITIONING EQ	EA	1.0	75000.00	75000.00	75000.00
2	STD TRANSFORMER	KVA	.0	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		15.0	15.020	.000	75000.00	75000.00
PERCENT ACCOUNT TOTAL,\$		15.0	15.020	.000	75000.00	75000.00
AUXILIARY MECH EQUIPMENT						
1	BOILER FEED PUMP	8DR. KWE	.0	.00	.00	.00
2	OTHER PUMPS	KWE	.0	.00	.00	.00
3	MIXTURE SERV SYS	KWE	225.34	1.17	260.88	23764.77
4	AUXILIARY BOILER	PPH	.0	4.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		16.0	.715	.000	260.88	23764.77
PERCENT ACCOUNT TOTAL,\$		16.0	.715	.000	260.88	23764.77

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Table
A 13.4.7 Continued

MOLTEN CARBONATE FUEL CELL SYSTEM
PARAMETRIC POINT NO. 1

ACCOUNT LISTING

ACCOUNT NO. & NAME, UNIT AMOUNT MAT \$/UNIT INS \$/UNIT MAT COST,\$ INS COST,\$

PIPE & FITTINGS
17. 1 CONVENTIONAL PIPING ION 10 C 3000.00 1300.00 30000.00 18000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 17 = .555 ACCOUNT TOTAL,\$ 30000.00

AUXILIARY ELEC EQUIPMENT
18. 1 MISC MOTORS, ETC 6510.9 1.40 9115.25 1106.85
18. 2 SWITCHGEAR & MCC PAN KWE 32554.5 1.95 63481.23 14599.51
18. 3 CONDUIT, CABLES, TRAYS FT 100000.0 1.32 132000.00 136000.00
18. 4 ISOLATED PHASE BUS FT 510.00 450.00
18. 5 LIGHTING & COMMUN KWE 32554.5 .35 11394.07 13998.42
PERCENT TOTAL DIRECT COST IN ACCOUNT 18 = 4.413 ACCOUNT TOTAL,\$ 215990.54 165754.79

CONTROL, INSTRUMENTATION
19. 1 COMPUTER EACH 1.0 100000.00 3000.00 3000.00
19. 2 OTHER CONTROLS EACH 1.0 83000.00 50000.00 50000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 19 = 2.728 ACCOUNT TOTAL,\$ 183000.00

PROCESS WASTE SYSTEMS
20. 1 BOTTOM ASH IPH .0 .00 .00 .00
20. 2 DRY ASH IPH .0 .00 .00 .00
20. 3 WET SLURRY IPH .0 .00 .00 .00
20. 4 ONSITE DISPOSAL ACRE .0 7676.49 11070.89 .00
PERCENT TOTAL DIRECT COST IN ACCOUNT 20 = .000 ACCOUNT TOTAL,\$.00

STACK GAS CLEANING
21. 1 PRECIPITATOR EACH .0 .00 .00 .00
21. 2 SCRUBBER KWE .0 .00 .00 .00
21. 3 MISC STEEL & DUCTS .0 .00 .00 .00
PERCENT TOTAL DIRECT COST IN ACCOUNT 21 = .000 ACCOUNT TOTAL,\$.00

TOTAL DIRECT COSTS,\$ 7491815.56 1158855.91

MOLTEN CARBONATE FUEL CELL SYSTEM COST OF ELECTRICITY • MILLS/KW • HR
 PARAMETRIC POINT NO. 1

Table A 13.4.8

ACCOUNT	RATE, PERCENT	LABOR RATE, \$/HR	15.00	21.45
TOTAL DIRECT COSTS, \$	14777.2	10.50	15.00	21.45
INDIRECT COST, \$	14777.2	34210.87	34210.87	11727.60
PROF & OWNER COSTS, \$	28192.22	47329.8	47329.8	15755.20
CONTINGENCY COST, \$	25068.00	69205.4	69205.4	22871.36
SUB TOTAL, \$	50078.00	77994.9	77994.9	44290.5
ESCALATION COST, \$	54250.00	93479.1	93479.1	12271.76
INTEREST DURING CONSTRUCTION, \$	55560.00	103230.22	111061.13	71777.00
TOTAL CAPITALIZATION, \$	109210.00	58964.00	64974.00	128900.13
TOTAL OF ELEC-CAPITAL	194368.47	115154.17	127401.8	194368.47
COST OF ELEC-FUEL MAIN	18.00	19.1884	19.1884	18.00
COST OF ELEC-OP & MAINT	18.00	22.37816	22.37816	18.00
TOTAL COST OF ELEC	50.00	57.00853	57.00853	50.00

ACCOUNT	RATE, PERCENT	CONTINGENCY, PERCENT	4.50	36.50
TOTAL DIRECT COSTS, \$	14777.2	38506.71	38506.71	12500.71
INDIRECT COST, \$	14777.2	50101.6	50101.6	16910.16
PROF & OWNER COSTS, \$	28192.22	92205.4	92205.4	30205.4
CONTINGENCY COST, \$	25068.00	38228.00	41253.74	17301.74
SUB TOTAL, \$	50078.00	95374.1	103230.22	116923.98
ESCALATION COST, \$	54250.00	56741.0	55211.6	82170.0
INTEREST DURING CONSTRUCTION, \$	55560.00	58964.00	60627.9	82170.0
TOTAL CAPITALIZATION, \$	109210.00	115154.17	115154.17	137391.9
TOTAL OF ELEC-CAPITAL	194368.47	19.1884	19.1884	19.1884
COST OF ELEC-FUEL MAIN	18.00	18.00	18.00	18.00
COST OF ELEC-OP & MAINT	18.00	22.37816	22.37816	18.00
TOTAL COST OF ELEC	50.00	57.00853	57.00853	50.00

ACCOUNT	RATE, PERCENT	ESCALATION RATE, PERCENT	8.00	36.50
TOTAL DIRECT COSTS, \$	14777.2	38506.71	38506.71	12500.71
INDIRECT COST, \$	14777.2	50101.6	50101.6	16910.16
PROF & OWNER COSTS, \$	28192.22	92205.4	92205.4	30205.4
CONTINGENCY COST, \$	25068.00	38228.00	38228.00	13333.8
SUB TOTAL, \$	50078.00	103230.22	103230.22	103230.22
ESCALATION DURING CONSTRUCTION, \$	54250.00	60371.0	61013.7	62956.4
INTEREST DURING CONSTRUCTION, \$	55560.00	58964.00	60627.9	82170.0
TOTAL CAPITALIZATION, \$	109210.00	115154.17	115154.17	137391.9
TOTAL OF ELEC-CAPITAL	194368.47	19.1884	19.1884	19.1884
COST OF ELEC-FUEL MAIN	18.00	18.00	18.00	18.00
COST OF ELEC-OP & MAINT	18.00	22.37816	22.37816	18.00
TOTAL COST OF ELEC	50.00	57.00853	57.00853	50.00

ACCOUNT	RATE, PERCENT	INT DURING CONSTRUCTION, PERCENT	10.00	36.50
TOTAL DIRECT COSTS, \$	14777.2	38506.71	38506.71	12500.71
INDIRECT COST, \$	14777.2	50101.6	50101.6	16910.16
PROF & OWNER COSTS, \$	28192.22	92205.4	92205.4	30205.4
CONTINGENCY COST, \$	25068.00	38228.00	38228.00	13333.8
SUB TOTAL, \$	50078.00	103230.22	103230.22	103230.22
ESCALATION DURING CONSTRUCTION, \$	54250.00	58964.00	58964.00	82170.0
INTEREST DURING CONSTRUCTION, \$	55560.00	48451.3	48451.3	82170.0
TOTAL CAPITALIZATION, \$	109210.00	115154.17	115154.17	137391.9
TOTAL OF ELEC-CAPITAL	194368.47	19.1884	19.1884	19.1884
COST OF ELEC-FUEL MAIN	18.00	18.00	18.00	18.00
COST OF ELEC-OP & MAINT	18.00	22.37816	22.37816	18.00
TOTAL COST OF ELEC	50.00	57.00853	57.00853	50.00

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MOLTEN CARBONATE FUEL CELL SYSTEM COST OF ELECTRICITY, MILLS/KW.HR
 Table A 13.4.8 Continued

ACCOUNT	10.00	14.00	21.00	FIXED CHARGE RATE, PCT	25.00
TOTAL DIRECT COSTS, \$	8650671.	8650671.	8650671.	18.00	8650671.
INDIRECT COST, \$	591016.	591016.	591016.	3650671.	591016.
PROF & OWNER COSTS, \$	692054.	692054.	692054.	591016.	692054.
CONTINGENCY COST, \$	389280.	389280.	389280.	692054.	389280.
SUB TOTAL, \$	10323022.	10323022.	10323022.	389280.	10323022.
ESCALATION COST, \$	589646.	589646.	589646.	10323022.	589646.
INTREST DURING CONST, \$	603750.	603750.	603750.	589646.	603750.
TOTAL CAPITALIZATION, \$	11516417.	11516417.	11516417.	603750.	11516417.
COST OF ELEC-CAPITAL	9.03430	13.00939	19.51408	11516417.	22.58574
COST OF ELEC-FUEL	18.16864	18.16864	18.16864	19.51408	19.16864
COST OF ELEC-OP & MAIN	23.37816	23.37816	23.37816	18.16864	23.37816
TOTAL COST OF ELEC	50.58109	54.55618	57.90853	23.37816	61.13254

ACCOUNT	1.50	4.00	2.00	FUEL COST, \$/10**6 BTU	3.12
TOTAL DIRECT COSTS, \$	8650671.	8650671.	8650671.	4.00	8650671.
INDIRECT COST, \$	591016.	591016.	591016.	8650671.	591016.
PROF & OWNER COSTS, \$	692054.	692054.	692054.	591016.	692054.
CONTINGENCY COST, \$	389280.	389280.	389280.	692054.	389280.
SUB TOTAL, \$	10323022.	10323022.	10323022.	389280.	10323022.
ESCALATION COST, \$	589646.	589646.	589646.	10323022.	589646.
INTREST DURING CONST, \$	603750.	603750.	603750.	589646.	603750.
TOTAL CAPITALIZATION, \$	11516417.	11516417.	11516417.	603750.	11516417.
COST OF ELEC-CAPITAL	16.26173	16.26173	16.26173	11516417.	16.26173
COST OF ELEC-FUEL	10.48191	18.16664	27.95176	16.26173	21.80237
COST OF ELEC-OP & MAIN	23.37816	23.37816	23.37816	18.16664	23.37816
TOTAL COST OF ELEC	50.12180	57.80853	67.59165	23.37816	61.44226

ACCOUNT	12.00	45.00	50.00	CAPACITY FACTOR, PERCENT	80.00
TOTAL DIRECT COSTS, \$	8650671.	8650671.	8650671.	50.00	8650671.
INDIRECT COST, \$	591016.	591016.	591016.	8650671.	591016.
PROF & OWNER COSTS, \$	692054.	692054.	692054.	591016.	692054.
CONTINGENCY COST, \$	389280.	389280.	389280.	692054.	389280.
SUB TOTAL, \$	10323022.	10323022.	10323022.	389280.	10323022.
ESCALATION COST, \$	589646.	589646.	589646.	10323022.	589646.
INTREST DURING CONST, \$	603750.	603750.	603750.	589646.	603750.
TOTAL CAPITALIZATION, \$	11516417.	11516417.	11516417.	603750.	11516417.
COST OF ELEC-CAPITAL	18.08438	23.04891	21.14025	11516417.	11.516417
COST OF ELEC-FUEL	18.16864	19.16864	18.16864	21.14025	18.16864
COST OF ELEC-OP & MAIN	20.77361	27.71984	23.00879	18.16864	23.37816
TOTAL COST OF ELEC	133.02663	65.37765	62.91768	23.00879	54.61531

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Table A 13.4.9

MOLTEN CARBONATE FUEL CELL SYSTEM

ACCOUNT NO	AUX POWER, MWE	PERC PLANT POW	OPERATION COST	MAINTENANCE COST					
11	.00000	.00000	495.00000	98000.00000					
18	.11250	100.00000	.00000	.00000					
TOTALS	.11250	.50251	506.16736	98000.00000					
MOLTEN CARBONATE FUEL CELL SYSTEM	BASE CASE INPUT			22.3875					
NOMINAL POWER, MWE	22.5000			6987.9395					
NET HEAT RATE, BTU/KW-HR	6952.9999								
NET TURB HEAT RATE CHANGE	.0000								
CONDENSER									
DESIGN PRESSURE, IN HG A	2.0000			.0000					
NUMBER OF TUBES/SHELL	.0000			.0000					
U, BTU/HR-FI2-F	.0000			.0000					
HEAT REJECTION	.0000			.0000					
DESIGN TEMP, F	.0000			.0000					
OFF DESIGN PRES, IN HG A	.0000			.0000					
1	22.500	2	.000	3	.000	4	5953.000	5	1.500
6	.000	7	2.000	8	.000	9	.000	10	.000
11	.000	12	.000	13	.000	14	1.000	15	.000
16	.000	17	1.500	18	.000	19	5.000	20	.000
21	.000	22	3200.000	23	.000	24	.000	25	.000
26	.000	27	5000.000	28	.000	29	.000	30	.000
31	1.000	32	10.000	33	.000	34	.200	35	1.000
36	100000.000	37	10.000	38	.000	39	1.000	40	170000.000
41	30000.000	42	100000.000	43	.000	44	3000.000	45	50000.000
46	.000	47	.000	48	.000	49	1.000	50	6.000
51	.000	52	5.350	53	.000	54	.000	55	.000
1	.000	2	.000	3	.000	4	1.000	5	.000
6	.000	7	.000	8	.000	9	.000	10	.000
11	.000	12	.000	13	.000	14	.000	15	.000
16	.000	17	.000	18	.000	19	.000	20	.000
21	.000	22	.000	23	.000	24	29000.000	25	3000.000
26	.000	27	.000	28	.000	29	4750000.000	30	200000.000
31	.000	32	.000	33	.000	34	1.000	35	102000.000
36	51000.000	37	1.000	38	.000	39	136000.000	40	1.000
41	.000	42	1380000.000	43	.000	44	175000.000	45	.000
46	.000	47	.000	48	.000	49	.000	50	.000
51	.200	52	.000	53	.000	54	.000	55	.000
56	.000	57	.000	58	.000	59	495.000	60	98000.000

Table SOLID ELECTROLYTE FUEL CELL ACCOUNT LISTING
 A 13.4.10 PARAMETRIC POINT NO. 1

ACCOUNT NO. & NAME	UNIT	AMOUNT	MAT \$/UNIT	IMS \$/UNIT	MAT COST \$	IMS COST \$
SITE DEVELOPMENT						
1. 1 LAND COST	ACRE	1.5	100000.00		150000.00	.00
1. 2 CLEARING LAND	ACRE	1.5		EGC .00		299.97
1. 3 GRADING LAND	ACRE	1.5		3000.00		4500.00
1. 4 ACCESS RAILROAD	MILE	.0	115000.00			.00
1. 5 LOOP RAILROAD TRACK	MILE	.0	120000.00			.00
1. 6 SIDING R R TRACK	MILE	.0	125000.00			.00
1. 7 OTHER SITE COSTS	ACRE	.0			3736.39	3736.39
PERCENT TOTAL DIRECT COST IN ACCOUNT		1 = 2.260	ACCOUNT TOTAL \$		153736.39	8536.36
EXCAVATION & PILING						
2. 1 COMMON EXCAVATION	YD3	9600.0		3.00		28800.00
2. 2 PILING	FT	.0	6.50	8.50		.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		2 = .401	ACCOUNT TOTAL \$			28800.00
PLANT ISLAND CONCRETE						
3. 1 PLANT IS. CONCRETE	YD3	3200.0	70.00	80.00	224000.00	256000.00
3. 2 SPECIAL STRUCTURES	YD3	.0				.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		3 = 6.685	ACCOUNT TOTAL \$		224000.00	256000.00
HEAT REJECTION SYSTEM						
4. 1 COOLING TOWERS	EACH	.0				.00
4. 2 CIRCULATING H2O SYS	EACH	1.0				.00
4. 3 SURFACE CONDENSER	FI2	.0				.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		4 = .000	ACCOUNT TOTAL \$.00
STRUCTURAL FEATURES						
5. 1 STAT. STRUCTURAL ST.	TON	.0	650.00	175.00		.00
5. 2 SILOS & BUNKERS	TPH	.0	1800.00	750.00		.00
5. 3 CHIMNEY	FT	.0				.00
5. 4 STRUCTURAL FEATURES EACH		1.0	170000.00	30000.00	170000.00	30000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		5 = 2.786	ACCOUNT TOTAL \$		170000.00	30000.00
BUILDINGS						
6. 1 STATION BUILDINGS	FI3	.0		16		.00
6. 2 ADMINISTRATION	FI2	5000.0	16.00	14.00	80000.00	70000.00
6. 3 WAREHOUSE & SHOP	FI2	.0	12.00	8.00		.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		6 = 2.089	ACCOUNT TOTAL \$		80000.00	70000.00
FUEL HANDLING & STORAGE						
7. 1 COAL HANDLING SYS	TPH	.0				.00
7. 2 POLYMETE HAND. SYS	TPH	.0				.00
7. 3 FUEL OIL HAND. SYS	GAL	.0				.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		7 = .000	ACCOUNT TOTAL \$.00

ACCOUNT LISTING

SOLID ELECTROLYTE FUEL CELL PARAMETRIC POINT NO. 1

A 13.4.10 Continued

ACCOUNT NO. & NAME UNIT AMOUNT MAT \$/UNIT INS \$/UNIT MAT COST,\$ INS COST,\$

FUEL PROCESSING
 3. 1 COAL GASIFIER EACH .00
 8. 2 DESULFURIZER EACH .00
 8. 3 VENTURI SCRUBBER EACH .00
 8. 4 COAL PRETREATMENT EACH .00
 8. 5 REFORMER EACH .00
 8. 6 SHIFT CONVERTER EACH .00
 8. 7 CO2 SCRUBBER EACH .00
 8. 8 BLOWER EACH .00
 3 PERCENT TOTAL DIRECT COST IN ACCOUNT 1.0 67000.00 50000.00
 3 = 1.530 ACCOUNT TOTAL,\$ 67000.00 50000.00

FIRING SYSTEM
 9. 1 PERCENT TOTAL DIRECT COST IN ACCOUNT 9.0 .000 ACCOUNT TOTAL,\$.00
 9 = .000 ACCOUNT TOTAL,\$.00

VAPOR GENERATOR (FIRED)
 10. 1 PERCENT TOTAL DIRECT COST IN ACCOUNT 10.0 .00 ACCOUNT TOTAL,\$.00
 10 = .000 ACCOUNT TOTAL,\$.00

ENERGY CONVERTER
 11. 1 FUEL CELL SYSTEM EACH 1.0 310000.00 250000.00 310000.00 250000.00
 11. 2 STIM TURBINE GENERATOR EA .00 .00 .00 .00
 11. 3 OXYGEN PLANT EACH .00 .00 .00 .00
 11 PERCENT TOTAL DIRECT COST IN ACCOUNT 11.0 =46.659 ACCOUNT TOTAL,\$ 310000.00 250000.00

COUPLING HEAT EXCHANGER
 12. 1 HEAT REC STEAM GEN EACH 1.0 44000.00 18000.00 44000.00 18000.00
 12 PERCENT TOTAL DIRECT COST IN ACCOUNT 12.0 = .854 ACCOUNT TOTAL,\$ 44000.00 18000.00

HEAT RECOVERY HEAT EXCH.
 13. 1 RECUPERATOR EACH 1.0 334000.00 110000.00 334000.00 110000.00
 13 PERCENT TOTAL DIRECT COST IN ACCOUNT 13.0 = 6.184 ACCOUNT TOTAL,\$ 334000.00 110000.00

WATER TREATMENT
 14. 1 DEMINERALIZER GPM .00 2500.00 700.00 .00
 14. 2 CONDENSATE POLISHING KWE 1.25 1.25 .30 .30
 14 PERCENT TOTAL DIRECT COST IN ACCOUNT 14.0 = .000 ACCOUNT TOTAL,\$ 2500.00 700.00

POWER CONDITIONING
 15. 1 POWER CONDITIONING EQ EA 1.0 1380000.00 75000.00 1380000.00 75000.00
 15. 2 STD TRANSFORMER KVA .00 .00 .00 .00
 15 PERCENT TOTAL DIRECT COST IN ACCOUNT 15.0 =20.265 ACCOUNT TOTAL,\$ 1380000.00 75000.00

AUXILIARY MECH EQUIPMENT
 16. 1 BOILER FEED PUMP 8DR KWE .00 .00 .00 .00
 16. 2 OTHER PUMPS KWE .00 .00 .00 .00
 16. 3 MISC SERVICE SYS KWE 33124.3 1.17 .73 .73
 16. 4 AUXILIARY BOILER PPH 4.00 .80 .80 .80
 16 PERCENT TOTAL DIRECT COST IN ACCOUNT 16.0 = .877 ACCOUNT TOTAL,\$ 32755.49 24180.77

Table SOLID ELECTROLYTE FUEL CELL ACCOUNT LISTING
 A 13.4.10 Continued PARAMETRIC POINT NO. 1

ACCOUNT NO. & NAME,	UNIT	AMOUNT	MAT \$/UNIT	INS \$/UNIT	MAT COST,\$	INS COST,\$
PIPE & FITTINGS						
17. 1 CONVENTIONAL PIPING	TON	10.0	3000.00	1800.00	30000.00	18000.00
PERCENT TOTAL DIRECT COST	IN ACCOUNT 17 =	.669	ACCOUNT TOTAL,\$		30000.00	18000.00
AUXILIARY ELEC EQUIPMENT						
18. 1 MISC MOTORS, ETC		6624.9	1.00	.17	9274.82	1126.23
18. 2 SWITCHGEAR & MCC PAN	KWE	33124.3	1.95	.45	64592.48	14905.96
18. 3 CONDUIT, CABLES, TRAYS	FT	100000.0	1.32	1.36	132000.00	136000.00
18. 4 ISOLATED PHASE BUS	FT		510.00	450.00		
18. 5 LIGHTING & COMMON	KWE	33124.3	1.35	.43	11593.52	14243.47
PERCENT TOTAL DIRECT COST	IN ACCOUNT 18 =	5.345	ACCOUNT TOTAL,\$		217460.81	166275.65
CONTROL, INSTRUMENTATION						
19. 1 COMPUTER	EACH	1.0	100000.00	3000.00	100000.00	3000.00
19. 2 OTHER CONTROLS	EACH	1.0	83000.00	50000.00	83000.00	50000.00
PERCENT TOTAL DIRECT COST	IN ACCOUNT 19 =	3.287	ACCOUNT TOTAL,\$		183000.00	53000.00
PROCESS WASTE SYSTEMS						
20. 1 BOTTOM ASH	TPH	0	0.00	0.00	0.00	0.00
20. 2 DRY ASH	TPH	0	0.00	0.00	0.00	0.00
20. 3 WET SLURRY	TPH	0	0.00	0.00	0.00	0.00
20. 4 ONSITE DISPOSAL	ACRE		7676.49	11070.89		
PERCENT TOTAL DIRECT COST	IN ACCOUNT 20 =	.000	ACCOUNT TOTAL,\$		0.00	0.00
STACK GAS CLEANING						
21. 1 PRECIPITATOR	EACH	0	0.00	0.00	0.00	0.00
21. 2 SCRUBBER	KWE	0	0.00	0.00	0.00	0.00
21. 3 MISC STEEL & DUCTS		0	0.00	0.00	0.00	0.00
PERCENT TOTAL DIRECT COST	IN ACCOUNT 21 =	.000	ACCOUNT TOTAL,\$		0.00	0.00

TOTAL DIRECT COSTS,\$ 6021952.62 1157792.77

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SOLID ELECTROLYTE FUEL CELL PARAMETRIC POINT NO. 1 COST OF ELECTRICITY, MILLS/KW·HR

Table A 13.4.11

ACCOUNT	RATE, PERCENT	LABOR RATE, \$/HR	CONTINGENCY, PERCENT	ESCALATION RATE, PERCENT	INT DURING CONST, PERCENT
TOTAL DIRECT COSTS, \$	6.00	8.50	5.00	6.50	8.00
INDIRECT COST, \$	677307.	1179745.	7179745.	7179745.	7179745.
PROF & OWNER COSTS, \$	51.00	473434.	590474.	590474.	590474.
CONTINGENCY COST, \$	4.00	312767.	323089.	323089.	323089.
SUB TOTAL, \$	7846201.	8292661.	9344599.	8667688.	8667688.
ESCALATION COST, \$	6.00	485094.	485094.	485094.	485094.
INTREST DURING CONST, \$	10.00	485094.	485094.	485094.	485094.
TOTAL CAPITALIZATION, \$	18.00	970188.	970188.	970188.	970188.
COST OF ELEC-CAPITAL	12.72563	12.72563	12.72563	12.72563	12.72563
COST OF ELEC-FUEL	16.00160	16.00160	16.00160	16.00160	16.00160
TOTAL COST OF ELEC	40.87136	41.56233	42.14283	42.30868	42.30868

ACCOUNT	RATE, PERCENT	LABOR RATE, \$/HR	CONTINGENCY, PERCENT	ESCALATION RATE, PERCENT	INT DURING CONST, PERCENT
TOTAL DIRECT COSTS, \$	5.00	8.50	5.00	6.50	8.00
INDIRECT COST, \$	7179745.	1179745.	7179745.	7179745.	7179745.
PROF & OWNER COSTS, \$	51.00	473434.	590474.	590474.	590474.
CONTINGENCY COST, \$	20.00	312767.	323089.	323089.	323089.
SUB TOTAL, \$	7985612.	8292661.	9344599.	8667688.	8667688.
ESCALATION COST, \$	6.00	485094.	485094.	485094.	485094.
INTREST DURING CONST, \$	10.00	485094.	485094.	485094.	485094.
TOTAL CAPITALIZATION, \$	16.00	970188.	970188.	970188.	970188.
COST OF ELEC-CAPITAL	12.72563	12.72563	12.72563	12.72563	12.72563
COST OF ELEC-FUEL	16.00160	16.00160	16.00160	16.00160	16.00160
TOTAL COST OF ELEC	41.08713	41.64276	42.14283	42.30868	42.30868

ACCOUNT	RATE, PERCENT	LABOR RATE, \$/HR	CONTINGENCY, PERCENT	ESCALATION RATE, PERCENT	INT DURING CONST, PERCENT
TOTAL DIRECT COSTS, \$	5.00	8.50	5.00	6.50	8.00
INDIRECT COST, \$	7179745.	1179745.	7179745.	7179745.	7179745.
PROF & OWNER COSTS, \$	51.00	473434.	590474.	590474.	590474.
CONTINGENCY COST, \$	4.00	312767.	323089.	323089.	323089.
SUB TOTAL, \$	8667688.	8292661.	9344599.	8667688.	8667688.
ESCALATION COST, \$	12.00	485094.	485094.	485094.	485094.
INTREST DURING CONST, \$	10.00	485094.	485094.	485094.	485094.
TOTAL CAPITALIZATION, \$	19.00	970188.	970188.	970188.	970188.
COST OF ELEC-CAPITAL	12.72563	12.72563	12.72563	12.72563	12.72563
COST OF ELEC-FUEL	16.00160	16.00160	16.00160	16.00160	16.00160
TOTAL COST OF ELEC	41.97693	42.14283	42.30868	42.30868	42.30868

ACCOUNT	RATE, PERCENT	LABOR RATE, \$/HR	CONTINGENCY, PERCENT	ESCALATION RATE, PERCENT	INT DURING CONST, PERCENT
TOTAL DIRECT COSTS, \$	6.00	8.50	5.00	6.50	8.00
INDIRECT COST, \$	7179745.	1179745.	7179745.	7179745.	7179745.
PROF & OWNER COSTS, \$	51.00	473434.	590474.	590474.	590474.
CONTINGENCY COST, \$	4.00	312767.	323089.	323089.	323089.
SUB TOTAL, \$	8667688.	8292661.	9344599.	8667688.	8667688.
ESCALATION COST, \$	6.00	485094.	485094.	485094.	485094.
INTREST DURING CONST, \$	15.00	485094.	485094.	485094.	485094.
TOTAL CAPITALIZATION, \$	18.00	970188.	970188.	970188.	970188.
COST OF ELEC-CAPITAL	12.72563	12.72563	12.72563	12.72563	12.72563
COST OF ELEC-FUEL	16.00160	16.00160	16.00160	16.00160	16.00160
TOTAL COST OF ELEC	41.86332	42.14283	42.30868	42.30868	42.30868

SOLID ELECTROLYTE FUEL CELL COST OF ELECTRICITY, MILLS/KW-HR
 Table A 13.4.11 Continued PARAMETRIC POINT NC. 1

ACCOUNT	RATE, PERCENT	FIXED CHARGE RATE, PCT	25.00
TOTAL DIRECT COSTS,\$	10.00	18.00	71.79745
INDIRECT COST,\$	71.79745	71.79745	590474
PROF & OWNER COSTS,\$	51.0	590474	574380
CONTINGENCY COST,\$	8.0	574380	323089
SUB TOTAL,\$	4.5	323089	8667688
ESCALATION COST,\$.0	8667688	495094
INTEREST DURING CONST,\$	6.5	495094	506936
TOTAL CAPITALIZATION,\$	10.0	506936	9669718
COST OF ELEC-CAPITAL	7.45311	9669718	13.41560
COST OF ELEC-FUEL	25.0	13.41560	12.72563
COST OF ELEC-OP & MAIN	.0	12.72563	16.00160
TOTAL COST OF ELEC	.0	16.00160	39.45971
		42.14283	

ACCOUNT	RATE, PERCENT	FUEL COST, \$/10**6 BTU	3.12
TOTAL DIRECT COSTS,\$	1.50	4.00	71.79745
INDIRECT COST,\$	71.79745	71.79745	590474
PROF & OWNER COSTS,\$	51.0	590474	574380
CONTINGENCY COST,\$	8.0	574380	323089
SUB TOTAL,\$	4.5	323089	8667688
ESCALATION COST,\$.0	8667688	495094
INTEREST DURING CONST,\$	6.5	495094	506936
TOTAL CAPITALIZATION,\$	10.0	506936	9669718
COST OF ELEC-CAPITAL	7.34171	9669718	13.41560
COST OF ELEC-FUEL	18.0	13.41560	19.57789
COST OF ELEC-OP & MAIN	.0	12.72563	16.00160
TOTAL COST OF ELEC	.0	16.00160	42.14283
		48.99509	

ACCOUNT	RATE, PERCENT	CAPACITY FACTOR, PERCENT	80.00
TOTAL DIRECT COSTS,\$	12.00	50.00	71.79745
INDIRECT COST,\$	71.79745	71.79745	590474
PROF & OWNER COSTS,\$	51.0	590474	574380
CONTINGENCY COST,\$	8.0	574380	323089
SUB TOTAL,\$	4.5	323089	8667688
ESCALATION COST,\$.0	8667688	495094
INTEREST DURING CONST,\$	6.5	495094	506936
TOTAL CAPITALIZATION,\$	10.0	506936	9669718
COST OF ELEC-CAPITAL	17.44029	9669718	13.41560
COST OF ELEC-FUEL	18.0	13.41560	12.72563
COST OF ELEC-OP & MAIN	.0	12.72563	16.00160
TOTAL COST OF ELEC	.0	16.00160	42.14283
		46.40799	

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Table A 13.4.12

SOLID ELECTROLYTE FUEL CELL

ACCOUNT NO	AUX POWER, MWE	PERC PLANT POW	OPERATION COST	MAINTENANCE COST
11	.00000	.00000	335.00000	10.00000
18	.11450	.1000000	.00000	.00000
TOTALS	.11450	.50251	335.00000	10.00000
SOLID ELECTROLYTE FUEL CELL			BASE CASE INPUT	22.7855
NOMINAL POWER, MWE	22.9000		NET POWER, MWE	489.4722
NOM HEAT RATE, BTU/KW-HR	4869.9999		NET HEAT RATE, BTU/KW-HR	
ST TURB HEAT RATE CHANGE	.0000			
CONDENSER DESIGN PRESSURE, IN HG A	2.0000		NUMBER OF SHELLS	.0000
NUMBER OF TUBES/SHELL	.0000		TUBE LENGTH, FT	.0000
U, BTU/HR-FI2-F	.0000		TERMINAL TEMP DIFF, F	.0000
HEAT REJECTION DESIGN TEMP, F	.0000		APPROACH, F	.0000
RANGE, F	.0000		OFF DESIGN TEMP, F	.0000
OFF DESIGN PRES, IN HG A	.0000		LP TURBINE BLADE LEN, IN	.0000
1	22.900	2	.000	4870.000
6	.000	7	.000	.000
11	.000	12	1.000	10
16	.000	17	5.000	15
21	.000	22	3200.000	20
26	.000	27	5000.000	25
31	1.000	32	10.000	30
36	100000.000	37	1.000	35
41	300000.000	42	3000.000	40
46	.000	47	1.000	45
51	.000	52	1.000	50
6	.000	7	1.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	.000	20
26	.000	27	.000	25
31	.000	32	.000	30
36	18000.000	37	330000.000	35
41	.000	42	750000.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	67000.000	20
26	.000	27	3100000.000	25
31	.000	32	.000	30
36	.000	37	330000.000	35
41	.000	42	750000.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	50000.000	20
26	.000	27	2500000.000	25
31	.000	32	.000	30
36	.000	37	1100000.000	35
41	.000	42	.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	50000.000	20
26	.000	27	2500000.000	25
31	.000	32	.000	30
36	.000	37	1100000.000	35
41	.000	42	.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	50000.000	20
26	.000	27	2500000.000	25
31	.000	32	.000	30
36	.000	37	1100000.000	35
41	.000	42	.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	50000.000	20
26	.000	27	2500000.000	25
31	.000	32	.000	30
36	.000	37	1100000.000	35
41	.000	42	.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	50000.000	20
26	.000	27	2500000.000	25
31	.000	32	.000	30
36	.000	37	1100000.000	35
41	.000	42	.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	50000.000	20
26	.000	27	2500000.000	25
31	.000	32	.000	30
36	.000	37	1100000.000	35
41	.000	42	.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	50000.000	20
26	.000	27	2500000.000	25
31	.000	32	.000	30
36	.000	37	1100000.000	35
41	.000	42	.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	50000.000	20
26	.000	27	2500000.000	25
31	.000	32	.000	30
36	.000	37	1100000.000	35
41	.000	42	.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	50000.000	20
26	.000	27	2500000.000	25
31	.000	32	.000	30
36	.000	37	1100000.000	35
41	.000	42	.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	50000.000	20
26	.000	27	2500000.000	25
31	.000	32	.000	30
36	.000	37	1100000.000	35
41	.000	42	.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	50000.000	20
26	.000	27	2500000.000	25
31	.000	32	.000	30
36	.000	37	1100000.000	35
41	.000	42	.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	50000.000	20
26	.000	27	2500000.000	25
31	.000	32	.000	30
36	.000	37	1100000.000	35
41	.000	42	.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	50000.000	20
26	.000	27	2500000.000	25
31	.000	32	.000	30
36	.000	37	1100000.000	35
41	.000	42	.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	50000.000	20
26	.000	27	2500000.000	25
31	.000	32	.000	30
36	.000	37	1100000.000	35
41	.000	42	.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	50000.000	20
26	.000	27	2500000.000	25
31	.000	32	.000	30
36	.000	37	1100000.000	35
41	.000	42	.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	50000.000	20
26	.000	27	2500000.000	25
31	.000	32	.000	30
36	.000	37	1100000.000	35
41	.000	42	.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	50000.000	20
26	.000	27	2500000.000	25
31	.000	32	.000	30
36	.000	37	1100000.000	35
41	.000	42	.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	50000.000	20
26	.000	27	2500000.000	25
31	.000	32	.000	30
36	.000	37	1100000.000	35
41	.000	42	.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15
21	.000	22	50000.000	20
26	.000	27	2500000.000	25
31	.000	32	.000	30
36	.000	37	1100000.000	35
41	.000	42	.000	40
46	.000	47	.000	45
51	.200	52	.000	50
6	.000	7	.000	5
11	.000	12	.000	10
16	.000	17	.000	15

Table A 13.4.13 SOLID ELECTROLYTE FUEL CELL PARAMETRIC POINT NO. 2 ACCOUNT LISTING

ACCOUNT NO. & NAME	UNIT	AMOUNT	MAT \$/UNIT	INS \$/UNIT	MAT COST,\$	INS COST,\$
SITE DEVELOPMENT						
1. 1 LAND COST	ACRE	2.2	10000.00	.00	22500.00	.00
1. 2 CLEARING LAND	ACRE	2.7	.00	600.00	.00	.00
1. 3 GRADING LAND	ACRE	2.2	.00	3000.00	.00	.00
1. 4 ACCESS RAILROAD	MILE	.0	115000.00	110000.00	.00	.00
1. 5 LOOP RAILROAD TRACK	MILE	.0	120000.00	70000.00	.00	.00
1. 6 SIDING R. TRACK	MILE	.0	125000.00	80000.00	.00	.00
1. 7 OTHER SITE COSTS	ACRE	.0	.00	.00	5599.76	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		1 =	.134	ACCOUNT TOTAL,\$	28099.76	12799.72

EXCAVATION & PILING						
2. 1 COMMON EXCAVATION	YD3	12150.0	.00	3.00	.00	36450.00
2. 2 PILING	FT	.0	6.50	8.50	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT	2 =	.119	ACCOUNT TOTAL,\$		36450.00	

PLANT ISLAND CONCRETE						
3. 1 PLANT IS. CONCRETE	YD3	4050.0	70.00	80.00	283500.00	324000.00
3. 2 SPECIAL STRUCTURES	YD3	.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT	3 =	1.987	ACCOUNT TOTAL,\$		283500.00	324000.00

HEAT REJECTION SYSTEM						
4. 1 COOLING TOWERS	EACH	.0	.00	.00	.00	.00
4. 2 CIRCULATING H2O SYS	EACH	1.0	.00	.00	.00	.00
4. 3 SURFACE CONDENSER	FT2	.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT	4 =	.000	ACCOUNT TOTAL,\$.00	.00

STRUCTURAL FEATURES						
5. 1 STAT. STRUCTURAL ST.	TON	.0	650.00	175.00	.00	.00
5. 2 SILOS & BUNKERS	FT	.0	1800.00	750.00	.00	.00
5. 3 CHIMNEY	FT	.0	.00	.00	.00	.00
5. 4 STRUCTURAL FEATURES EACH	EACH	1.0	280000.00	50000.00	280000.00	50000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT	5 =	1.079	ACCOUNT TOTAL,\$		280000.00	50000.00

BUILDINGS						
6. 1 STATION BUILDINGS	FT3	.0	.16	.16	.00	.00
6. 2 ADMINISTRATION	FT2	5000.0	16.00	14.00	80000.00	70000.00
6. 3 WAREHOUSE & SHOP	FT2	.0	12.00	8.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT	6 =	.491	ACCOUNT TOTAL,\$		80000.00	70000.00

FUEL HANDLING & STORAGE						
7. 1 COAL HANDLING SYS	TPH	.0	.00	.00	.00	.00
7. 2 DOLOMITE HAND. SYS	TPH	.0	.00	.00	.00	.00
7. 3 FUEL OIL HAND. SYS	GAL	.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT	7 =	.000	ACCOUNT TOTAL,\$.00	.00

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Table SOLID ELECTROLYTE FUEL CELL PARAMETRIC POINT NO. 2 ACCOUNT LISTING
 A 13.4.13 Continued

ACCOUNT NO. & NAME	UNIT	AMOUNT	MAT \$/UNIT	INS \$/UNIT	MAT COST,\$	INS COST,\$
FUEL PROCESSING						
1	COAL GASIFIER	EACH	.00	.00	.00	.00
2	DESULFURIZER	EACH	.00	.00	.00	.00
3	VENTURI SCRUBBER	EACH	.00	.00	.00	.00
4	COAL PRETREATMENT	EACH	.00	.00	.00	.00
5	REFORMER CONVERTER	EACH	.00	.00	.00	.00
6	SHIFT CONVERTER	EACH	.00	.00	.00	.00
7	CO2 SCRUBBER	EACH	.00	.00	.00	.00
8	FLOWER	EACH	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			1.0	269000.00	200000.00	200000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			8	1.531 ACCOUNT TOTAL,\$	269000.00	200000.00
FIRING SYSTEM						
9	PERCENT TOTAL DIRECT COST IN ACCOUNT	9	.000 ACCOUNT TOTAL,\$.00	.00	.00
10	VAPOR GENERATOR (FIRED)	10	.000 ACCOUNT TOTAL,\$.00	.00	.00
ENERGY CONVERTER						
11	FUEL CELL SYSTEM	EACH	1.0	1500000.00	1000000.00	1000000.00
11.1	2 STM TURBINE GENERATOR	EACH	.00	.00	.00	.00
11.2	3 OXYGEN PLANT	EACH	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			11	54.947 ACCOUNT TOTAL,\$	1580000.00	1000000.00
COUPLING HEAT EXCHANGER						
12	HEAT REC STEAM GEN	EACH	1.0	518000.00	247000.00	247000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			12	2.829 ACCOUNT TOTAL,\$	618000.00	247000.00
HEAT RECOVERY HEAT EXCH.						
13	RECUPERATOR	EACH	1.0	1910000.00	630000.00	630000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			13	3.308 ACCOUNT TOTAL,\$	1910000.00	630000.00
WATER TREATMENT						
14	DEMINERALIZER	SPM	.00	2500.00	700.00	.00
14.1	CONDENSATE POLISHING	KWE	.00	1.25	.30	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			14	.000 ACCOUNT TOTAL,\$.00	.00
POWER CONDITIONING						
15	POWER CONDITIONING EQ	EA	1.0	700000.00	300000.00	300000.00
15.1	STD TRANSFORMER	KVA	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			15	23.876 ACCOUNT TOTAL,\$	700000.00	300000.00
AUXILIARY MECH EQUIPMENT						
16	BOILER FEED PUMP	8DR. KWE	.00	.55	.04	.00
16.1	OTHER PUMPS	KWE	.00	.88	.12	.00
16.2	MISC SERVICE SYS	KWE	1.1402	1.17	142040.82	89623.76
16.3	AUXILIARY BOILER	PPH	.00	4.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			16	.754 ACCOUNT TOTAL,\$	147040.92	89623.76

Table SOLID ELECTROLYTE FUEL CELL ACCOUNT LISTING
 A 13.4.13 Continued PARAMETRIC POINT NO. 2

ACCOUNT NO. & NAME UNIT AMOUNT MAT \$/UNIT INS \$/UNIT MAT COST.\$ INS COST.\$

PIPE & FITTINGS		TON	3D.C	3000.00	1800.00	90000.00	54000.00
17.	1 CONVENTIONAL PIPING	IN ACCOUNT	17 =	.971	ACCOUNT	TOTAL.\$	90000.00
PERCENT TOTAL DIRECT COST							54000.00
AUXILIARY ELEC EQUIPMENT							
18.	1 MISC MOTORS,ETC	29280.5		1.40	.17	33992.67	4127.68
18.	2 SWITCHGEAR & MCC PAN	121402.4		1.95	.45	236734.69	54631.08
18.	3 CONDUIT,CABLES,TRAYS	150000.0		1.32	1.36	198000.00	204000.00
18.	4 ISOLATED PHASE BUS			510.00	450.00		
18.	5 LIGHTING & COMMUN	121402.4		.35	.43	42490.84	52203.03
PERCENT TOTAL DIRECT COST		IN ACCOUNT	18 =	2.702	ACCOUNT	TOTAL.\$	314951.79

CONTROL & INSTRUMENTATION							
19.	1 COMPUTER	EACH	1.0	100000.00	3000.00	100000.00	3000.00
19.	2 OTHER CONTROLS	EACH	1.0	83000.00	50000.00	83000.00	50000.00
PERCENT TOTAL DIRECT COST		IN ACCOUNT	19 =	.772	ACCOUNT	TOTAL.\$	183000.00

PROCESS WASTE SYSTEMS							
20.	1 BOTTOM ASH	TPH	.0	.00	.00	.00	.00
20.	2 DRY ASH	TPH	.0	.00	.00	.00	.00
20.	3 WET SLURRY	TPH	.0	.00	.00	.00	.00
20.	4 ONSITE DISPOSAL	ACRE		7676.49	11070.89		
PERCENT TOTAL DIRECT COST		IN ACCOUNT	20 =	.000	ACCOUNT	TOTAL.\$.00

STACK GAS CLEANING							
21.	1 PRECIPITATOR	EACH	.0	.00	.00	.00	.00
21.	2 SCRUBBER	KWE	.0	.00	.00	.00	.00
21.	3 MISC STEEL & DUCTS			.00	.00	.00	.00
PERCENT TOTAL DIRECT COST		IN ACCOUNT	21 =	.000	ACCOUNT	TOTAL.\$.00

TOTAL DIRECT COSTS.\$ 27193858.50 3380835.25

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Table A 13.4.14
SOLID ELECTROLYTE FUEL CELL PARAMETRIC POINT NO. 2 COST OF ELECTRICITY, MILLS/KW·HR

ACCOUNT	RATE, PERCENT	6.00	8.50	LABOR RATE, \$/HR	15.00	21.50
TOTAL DIRECT COSTS, \$	51.00	22107539.	29904905.	30574693.	31978059.	34051212.
INDIRECT COST, \$	5.00	975977.	1382624.	1724226.	2430942.	349751.
PROF & OWNER COSTS, \$	8.00	2392392.	2392392.	2445975.	2532245.	2724097.
CONTINGENCY COST, \$	5.00	1495245.	1528735.	1528735.	1598903.	1702561.
SUB TOTAL, \$	6.50	35877495.	35273629.	35273629.	38575148.	41975120.
ESCALATION COST, \$	10.00	2606385.	2707022.	2908720.	3091212.	3363868.
INTEREST DURING CONST, \$	18.00	2713965.	2810954.	41974965.	4635036.	49569122.
TOTAL CAPITALIZATION, \$	18.00	39187844.	40700547.	15.22254	16.18839	17.51122
TOTAL OF ELEC-CAPITAL	18.00	14.21279	15.22254	16.70352	16.70352	17.70352
COST OF ELEC-FUEL	18.00	16.70352	16.70352	20.76495	20.76495	20.76495
COST OF ELEC-OP & MAIN	18.00	16.70352	16.70352	20.76495	20.76495	20.76495
TOTAL COST OF ELEC	18.00	51.68125	52.23003	52.69101	53.65686	55.08369

ACCOUNT	RATE, PERCENT	5.00	10.00	CONTINGENCY, PERCENT	5.00	20.00
TOTAL DIRECT COSTS, \$	51.00	30574693.	30574693.	30574693.	30574693.	30574693.
INDIRECT COST, \$	5.00	1724226.	1724226.	1724226.	1724226.	1724226.
PROF & OWNER COSTS, \$	8.00	2445975.	2445975.	2445975.	2445975.	2445975.
CONTINGENCY COST, \$	5.00	1528735.	1528735.	1528735.	1528735.	1528735.
SUB TOTAL, \$	6.50	34744804.	36273629.	36273629.	36273629.	36273629.
ESCALATION COST, \$	10.00	2573918.	2731557.	2906780.	2906780.	31144504.
INTEREST DURING CONST, \$	18.00	2784275.	2906780.	41971965.	41971965.	47278830.
TOTAL CAPITALIZATION, \$	18.00	40203079.	41971965.	15.22254	15.22254	17.14718
TOTAL OF ELEC-CAPITAL	18.00	14.21279	15.22254	16.70352	16.70352	17.70352
COST OF ELEC-FUEL	18.00	16.70352	16.70352	20.76495	20.76495	20.76495
COST OF ELEC-OP & MAIN	18.00	16.70352	16.70352	20.76495	20.76495	20.76495
TOTAL COST OF ELEC	18.00	51.68125	52.04946	52.69101	52.69101	54.61565

ACCOUNT	RATE, PERCENT	5.00	6.50	ESCALATION RATE, PERCENT	10.00	12.50
TOTAL DIRECT COSTS, \$	51.00	30574693.	30574693.	30574693.	30574693.	30574693.
INDIRECT COST, \$	5.00	1724226.	1724226.	1724226.	1724226.	1724226.
PROF & OWNER COSTS, \$	8.00	2445975.	2445975.	2445975.	2445975.	2445975.
CONTINGENCY COST, \$	5.00	1528735.	1528735.	1528735.	1528735.	1528735.
SUB TOTAL, \$	6.50	36273629.	36273629.	36273629.	36273629.	36273629.
ESCALATION COST, \$	10.00	2791657.	2906780.	3443447.	4317075.	5416185.
INTEREST DURING CONST, \$	18.00	2906780.	2906780.	2947888.	3002821.	3071683.
TOTAL CAPITALIZATION, \$	18.00	41971965.	41971965.	42664953.	43593524.	44761496.
TOTAL OF ELEC-CAPITAL	18.00	14.21279	15.22254	15.47338	15.81165	16.23422.
COST OF ELEC-FUEL	18.00	16.70352	16.70352	16.70352	16.70352	16.70352
COST OF ELEC-OP & MAIN	18.00	16.70352	16.70352	20.76495	20.76495	20.76495
TOTAL COST OF ELEC	18.00	52.44074	52.69101	52.94235	53.27912	55.70273

ACCOUNT	RATE, PERCENT	5.00	9.00	INT DURING CONST, PERCENT	10.00	15.00
TOTAL DIRECT COSTS, \$	51.00	30574693.	30574693.	30574693.	30574693.	30574693.
INDIRECT COST, \$	5.00	1724226.	1724226.	1724226.	1724226.	1724226.
PROF & OWNER COSTS, \$	8.00	2445975.	2445975.	2445975.	2445975.	2445975.
CONTINGENCY COST, \$	5.00	1528735.	1528735.	1528735.	1528735.	1528735.
SUB TOTAL, \$	6.50	36273629.	36273629.	36273629.	36273629.	36273629.
ESCALATION COST, \$	10.00	2791657.	2906780.	2906780.	2906780.	31144504.
INTEREST DURING CONST, \$	15.00	40203079.	41971965.	41971965.	41971965.	47278830.
TOTAL CAPITALIZATION, \$	15.00	14.21279	15.22254	15.22254	15.22254	17.14718
TOTAL OF ELEC-CAPITAL	15.00	16.70352	16.70352	16.70352	16.70352	17.70352
COST OF ELEC-FUEL	15.00	16.70352	16.70352	20.76495	20.76495	20.76495
COST OF ELEC-OP & MAIN	15.00	16.70352	16.70352	20.76495	20.76495	20.76495
TOTAL COST OF ELEC	15.00	52.26993	52.48057	52.69101	52.69101	53.21634

Table A 13.4.14 Continued

SOLID ELECTROLYTE FUEL CELL PARAMETRIC POINT NO. 2 COST OF ELECTRICITY, MILLS/KW-HR

ACCOUNT	RATE, PERCENT	IC, CC	FIXED CHARGE RATE, PCI	25, CC
TOTAL DIRECT COSTS, \$.0	30574693	14.4C	30574693
INDIRECT COST, \$	51.0	1724226	30574693	1724226
PROF & OWNER COSTS, \$	8.0	2445975	1724226	2445975
CONTINGENCY COST, \$	5.0	1528735	2445975	1528735
SUB TOTAL, \$.0	36273629	1528735	36273629
ESCALATION COST, \$	6.5	2791557	36273629	2791557
INTEREST DURING CONST, \$	10.0	2906780	2791557	2906780
TOTAL CAPITALIZATION, \$.0	41971965	2906780	41971965
COST OF ELEC-CAPITAL	25.0	18.45697	41971965	18.45697
COST OF ELEC-FUEL	.0	16.70352	15.22254	16.70352
COST OF ELEC-OP & MAIN	.0	20.76495	16.70352	20.76495
TOTAL COST OF ELEC	.0	45.92544	20.76495	45.92544
			52.69101	52.69101

ACCOUNT	RATE, PERCENT	FUEL COST, \$/10**6 BTU	3, 10	2, 4C
TOTAL DIRECT COSTS, \$.0	30574693	30574693	30574693
INDIRECT COST, \$	51.0	1724226	1724226	1724226
PROF & OWNER COSTS, \$	8.0	2445975	2445975	2445975
CONTINGENCY COST, \$	5.0	1528735	1528735	1528735
SUB TOTAL, \$.0	36273629	1528735	36273629
ESCALATION COST, \$	6.5	2791557	36273629	2791557
INTEREST DURING CONST, \$	10.0	2906780	2791557	2906780
TOTAL CAPITALIZATION, \$.0	41971965	2906780	41971965
COST OF ELEC-CAPITAL	18.0	15.22254	41971965	15.22254
COST OF ELEC-FUEL	.0	10.02211	15.22254	10.02211
COST OF ELEC-OP & MAIN	.0	20.76495	16.70352	20.76495
TOTAL COST OF ELEC	.0	45.00960	20.76495	45.00960
			51.87794	51.87794

ACCOUNT	RATE, PERCENT	CAPACITY FACTOR, PERCENT	50, CC	80, CC
TOTAL DIRECT COSTS, \$.0	30574693	30574693	30574693
INDIRECT COST, \$	51.0	1724226	1724226	1724226
PROF & OWNER COSTS, \$	8.0	2445975	2445975	2445975
CONTINGENCY COST, \$	5.0	1528735	1528735	1528735
SUB TOTAL, \$.0	36273629	1528735	36273629
ESCALATION COST, \$	6.5	2791557	36273629	2791557
INTEREST DURING CONST, \$	10.0	2906780	2791557	2906780
TOTAL CAPITALIZATION, \$.0	41971965	2906780	41971965
COST OF ELEC-CAPITAL	18.0	21.98812	41971965	21.98812
COST OF ELEC-FUEL	.0	16.70352	19.78930	16.70352
COST OF ELEC-OP & MAIN	.0	25.74849	16.70352	25.74849
TOTAL COST OF ELEC	.0	124.90744	21.70352	124.90744
			57.59628	57.59628

Table A 13.4.15

SOLID ELECTROLYTE FUEL CELL

ACCOUNT NO	AUX POWER, MWE	PERC PLANT POW	OPERATION COST	MAINTENANCE COST			
11	.00000	.00000	1680.00000	560000.00000			
18	.43800	100.00000	.00000	.00000			
TOTALS	.43800	.50251	1711.56564	560000.00000			
SOLID ELECTROLYTE FUEL CELL	87.6000		BASE CASE INPUT	87.1620			
NOMINAL POWER, MWE	8309.9998		NET POWER, MWE	8351.7584			
NOM HEAT RATE, BTU/KW-HR			NET HEAT RATE, BTU/KW-HR				
ST TURB HEAT RATE CHANGE							
CONDENSER DESIGN PRESS, IN HG A	2.0000		NUMBER OF SHELLS	.0000			
DESIGN PRESS, IN HG B	.0000		TUBE LENGTH, FT	.0000			
U, BTU/HR-F ² -F	.0000		TERMINAL TEMP DIFF, F	.0000			
HEAT REJECTION DESIGN TEMP, F	.0000		APPROACH, F	.0000			
RANGE, F	.0000		OFF DESIGN TEMP, F	.0000			
OFF DESIGN PRES, IN HG A	.0000		LP TURBINE BLADE LEN, IN	.0000			
1	87.600	2	.000	4	8310.000	5	2.000
6	.000	7	.000	9	.000	10	.000
11	.000	12	.000	14	1.000	15	.000
16	.000	17	2.250	19	4.000	20	.000
21	.000	22	4050.000	24	.000	25	.000
26	.000	27	5000.000	29	.000	30	.000
31	1.000	32	30.000	34	.000	35	1.000
36	150000.000	37	100000.000	39	1.000	40	280000.000
41	50000.000	42	100000.000	44	3000.000	45	50000.000
46	.000	47	.000	49	1.000	50	6.000
51	.000	52	5.350				
1	.000	2	.000	9	1.000	10	.000
6	.000	7	.000	14	.000	15	.000
11	.000	12	.000	19	.000	20	.000
16	.000	17	.000	24	268000.000	25	1.000
21	.000	22	.000	29	15800000.000	30	1000000.000
26	.000	27	.000	34	1910000.000	35	630000.000
31	247000.000	32	1.000	39	300000.000	40	1.000
36	.000	37	7000000.000	44	.000	45	.000
41	.000	42	.000	49	.000	50	.000
46	.000	47	.000	54	.000	55	.000
51	.200	52	.000	59	.600	60	560000.000

Table A 13.4.16

SOLID ELECTROLYTE FUEL CELL
PARAMETRIC POINT NO. 4
ACCOUNT LISTING

ACCOUNT NO. & NAME,	UNIT	AMOUNT	MAT \$/UNIT	INS \$/UNIT	MAT COST,\$	INS COST,\$
SITE DEVELOPMENT						
1. 1 LAND COST	ACRE	30.0	1000.00	.00	30000.00	.00
1. 2 CLEARING LAND	ACRE	10.0	.00	600.00	.00	5999.40
1. 3 GRADING LAND	ACRE	30.0	.00	3000.00	.00	90000.00
1. 4 ACCESS RAILROAD	MILE	5.0	115000.00	110000.00	575000.00	550000.00
1. 5 LOOP RAILROAD TRACK	MILE	.0	120000.00	70000.00	.00	.00
1. 6 SIDING R R TRACK	MILE	.5	125000.00	80000.00	62500.00	90000.00
1. 7 OTHER SITE COSTS	ACRE	0	.00	.00	72416.89	72416.89
PERCENT TOTAL DIRECT COST IN ACCOUNT		1 =	.490 ACCOUNT TOTAL,\$		739916.88	758416.28
EXCAVATION & PILING						
2. 1 COMMON EXCAVATION	YD3	86700.0	.00	3.00	.00	260100.00
2. 2 PILING	FT	0	6.50	8.50	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		2 =	.085 ACCOUNT TOTAL,\$.00	260100.00
PLANT ISLAND CONCRETE						
3. 1 PLANT IS. CONCRETE	YD3	28900.0	70.00	80.00	2023000.00	2312000.00
3. 2 SPECIAL STRUCTURES	YD3	0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		3 =	1.419 ACCOUNT TOTAL,\$		2023000.00	2312000.00
HEAT REJECTION SYSTEM						
4. 1 COOLING TOWERS	EACH	1.0	.00	.00	153500.00	76500.00
4. 2 CIRCULATING H2O SYS	EACH	1.0	.00	.00	637.12	854.29
4. 3 SURFACE CONDENSER	FT2	232.5	.00	.00	283709.77	162.74
PERCENT TOTAL DIRECT COST IN ACCOUNT		4 =	.169 ACCOUNT TOTAL,\$		437846.89	77517.03
STRUCTURAL FEATURES						
5. 1 STAT. STRUCTURAL	ST. TON	.0	650.00	175.00	.00	.00
5. 2 SILOS & BUNKERS	TPH	.0	1800.00	750.00	.00	.00
5. 3 CHIMNEY	FT	0	.00	.00	.00	.00
5. 4 STRUCTURAL FEATURES	EACH	1.0	942000.00	197000.00	942000.00	197000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		5 =	.373 ACCOUNT TOTAL,\$		942000.00	197000.00
BUILDINGS						
6. 1 STATION BUILDINGS	FT2	0	16	16	.00	.00
6. 2 ADMINISTRATION	FT2	10000.0	16.00	14.00	160000.00	140000.00
6. 3 WAREHOUSE & SHOP	FT2	15000.0	12.00	8.00	180000.00	120000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		6 =	.196 ACCOUNT TOTAL,\$		340000.00	260000.00
FUEL HANDLING & STORAGE						
7. 1 COAL HANDLING SYS	IPH	0	.00	.00	.00	.00
7. 2 DOLOMITE HAND. SYS	TPH	0	.00	.00	.00	.00
7. 3 FUEL OIL HAND. SYS	GAL	0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		7 =	.000 ACCOUNT TOTAL,\$.00	.00

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Table A 13.4.16 Continued

SOLID ELECTROLYTE FUEL CELL PARAMETRIC POINT NO. 4		ACCOUNT LISTING				
ACCOUNT NO. & NAME	UNIT	AMOUNT	MAT \$/UNIT	INS \$/UNIT	MAT COST,\$	INS COST,\$
FUEL PROCESSING						
9. 1	COAL GASIFIER	EACH	.00	.00	.00	.00
9. 2	DESULFURIZER	EACH	.00	.00	.00	.00
9. 3	VENTURI SCRUBBER	EACH	.00	.00	.00	.00
8. 4	COAL PRETREATMENT	EACH	.00	.00	.00	.00
8. 5	REFORMER	EACH	.00	.00	.00	.00
9. 6	SHIFT CONVERTER	EACH	.00	.00	.00	.00
3. 7	CO2 SCRUBBER	EACH	.00	.00	.00	.00
3. 8	BLOWER	EACH	1.00	2412000.00	1800000.00	1300000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		8 =	1.373	ACCOUNT TOTAL,\$	2412000.00	1800000.00
FIRING SYSTEM						
9. 1	PERCENT TOTAL DIRECT COST IN ACCOUNT	9 =	.000	ACCOUNT TOTAL,\$.00	.00
VAPOR GENERATOR (FIRED)						
10. 1	PERCENT TOTAL DIRECT COST IN ACCOUNT	10 =	.000	ACCOUNT TOTAL,\$.00	.00
ENERGY CONVERTER						
11. 1	FUEL CELL SYSTEM	EACH	1.0142000000.00	9000000.00	414200000.00	9000000.00
11. 2	STM TURBINE GENERATOR	EA	1.0011518000.00	876934.13	11518000.00	876934.13
11. 3	OXYGEN PLANT	EACH	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		11 =	53.485	ACCOUNT TOTAL,\$	153518000.00	9876934.12
COUPLING HEAT EXCHANGER						
12. 1	HEAT REC STEAM GEN	EACH	1.0015624000.00	6695000.00	15624000.00	6695000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		12 =	7.306	ACCOUNT TOTAL,\$	15624000.00	6695000.00
HEAT RECOVERY HEAT EXCH.						
13. 1	RECUPERATOR	EACH	1.0018400000.00	6700000.00	18400000.00	6700000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		13 =	8.216	ACCOUNT TOTAL,\$	18400000.00	6700000.00
WATER TREATMENT						
14. 1	DEMINERALIZER	GPM	59.5	2500.00	700.00	41864.00
14. 2	CONDENSATE POLISHING	KWE	372000.00	1.25	30	111600.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		14 =	.251	ACCOUNT TOTAL,\$	613799.99	153264.00
POWER CONDITIONING						
15. 1	POWER CONDITIONING	EG EA	1.0059000000.00	2700000.00	59000000.00	2700000.00
15. 2	STD TRANSFORMER	KVA	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		15 =	20.196	ACCOUNT TOTAL,\$	59000000.00	2700000.00
AUXILIARY MECH EQUIPMENT						
16. 1	BOILER FEED PUMP	&DR. KWE	365280.4	1.67	10	30528.04
16. 2	OTHER PUMPS	KWE	571487.1	.88	12	68578.45
16. 3	MISC SERVICE SYS	KWE	1142974.2	1.17	73	834371.15
16. 4	AUXILIARY BOILER	EHP	.00	4.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT		16 =	1.110	ACCOUNT TOTAL,\$	24502006.72	979477.63

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Table A 13.4.16 Continued

SOLID ELECTROLYTE FUEL CELL
PARAMETRIC POINT NO. 4 ACCOUNT LISTING

ACCOUNT NO. & NAME	UNIT	AMOUNT	MAT \$/UNIT	INS \$/UNIT	MAT COST,\$	INS COST,\$
PIPE & FITTINGS						
17. 1 CONVENTIONAL PIPING	TON	845.0	3000.00	1800.00	2535000.00	1521000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 17 =		1.328	ACCOUNT TOTAL,\$		2535000.00	1521000.00
AUXILIARY ELEC EQUIPMENT						
18. 1 MISC MOTORS,ETC		571487.1	1.40	.17	800081.93	97152.81
18. 2 SWITCHGEAR & MCC PAN	KWE	1142974.2	1.95	.45	2228799.66	514338.38
18. 3 CONDUIT, CABLES, TRAYS	FT	2198000.0	1.32	1.36	2901359.97	2989279.97
18. 4 ISOLATED PHASE BUS	FT	300.0	510.00	450.00	153000.00	135000.00
18. 5 LIGHTING & COMMUN	KWE	1142974.2	.35	.43	400040.96	491478.90
PERCENT TOTAL DIRECT COST IN ACCOUNT 18 =		3.506	ACCOUNT TOTAL,\$		6483282.44	4227250.00
CONTROL, INSTRUMENTATION						
19. 1 COMPUTER	EACH	1.0	560000.00	15000.00	660000.00	15000.00
19. 2 OTHER CONTROLS	EACH	1.0	775500.00	50000.00	775500.00	50000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 19 =		.491	ACCOUNT TOTAL,\$		1435500.00	65000.00
PROCESS WASTE SYSTEMS						
20. 1 BOTTOM ASH	TPH	.0	.00	.00	.00	.00
20. 2 DRY ASH	TPH	.0	.00	.00	.00	.00
20. 3 WET SLURRY	TPH	.0	.00	.00	.00	.00
20. 4 ONSITE DISPOSAL	ACRE	.0	7676.49	11070.89	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 20 =		.000	ACCOUNT TOTAL,\$.00	.00
STACK GAS CLEANING						
21. 1 PRECIPITATOR	EACH	.0	.00	.00	.00	.00
21. 2 SCRUBBER	KWE	.0	.00	.00	.00	.00
21. 3 MISC STEEL & DUCYS		.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 21 =		.000	ACCOUNT TOTAL,\$.00	.00
TOTAL DIRECT COSTS,\$					266954548.00	38543958.50

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Table A 13.4.17
SOLID ELECTROLYTE FUEL CELL PARAMETRIC POINT NO. 4
COST OF ELECTRICITY, MILLS/KW·HR

ACCOUNT	RATE, PERCENT	LABOR RATE, \$/HR	CONST. PERCENT	CONTINGENCY, PERCENT	ESCALATION RATE, PERCENT	INT DURING CONST. PERCENT
TOTAL DIRECT COSTS, \$	100.0	15.00	100.0	100.0	100.0	100.0
INDIRECT COSTS, \$	1.0	32149.7884	3054.98504	3054.98504	3054.98504	3054.98504
PROF & OWNER COSTS, \$	8.0	27817.102	19657.418	19657.418	19657.418	19657.418
CONTINGENCY COST, \$	8.0	25719.830	2443.9880	2443.9880	2443.9880	2443.9880
SUB TOTAL, \$	16.0	40075.4640	37403.5680	37403.5680	37403.5680	37403.5680
ESCALATION COST, \$	6.0	8216.3584	7668.5605	7668.5605	7668.5605	7668.5605
TOTAL DURING CONST, \$	12.0	9464.2507	8833.2538	8833.2538	8833.2538	8833.2538
TOTAL CAPITALIZATION, \$	18.0	175.79703	14.74386	14.74386	14.74386	14.74386
COST OF ELEC-CAPITAL	18.0	11.34006	11.34006	11.34006	11.34006	11.34006
COST OF ELEC-FUEL	14.0	14.15883	14.15883	14.15883	14.15883	14.15883
COST OF ELEC-OP & MAIN	14.0	14.15883	14.15883	14.15883	14.15883	14.15883
TOTAL COST OF ELEC	50.0	41.29602	40.24281	40.24281	40.24281	40.24281
ACCOUNT						
TOTAL DIRECT COSTS, \$	100.0	5.00	100.0	100.0	100.0	100.0
INDIRECT COSTS, \$	51.0	3054.98504	3054.98504	3054.98504	3054.98504	3054.98504
PROF & OWNER COSTS, \$	20.0	19657.418	19657.418	19657.418	19657.418	19657.418
CONTINGENCY COST, \$	20.0	2443.9880	2443.9880	2443.9880	2443.9880	2443.9880
SUB TOTAL, \$	91.0	37403.5680	37403.5680	37403.5680	37403.5680	37403.5680
ESCALATION COST, \$	6.0	7668.5605	7668.5605	7668.5605	7668.5605	7668.5605
TOTAL DURING CONST, \$	10.0	8833.2538	8833.2538	8833.2538	8833.2538	8833.2538
TOTAL CAPITALIZATION, \$	10.0	14.74386	14.74386	14.74386	14.74386	14.74386
COST OF ELEC-CAPITAL	10.0	11.34006	11.34006	11.34006	11.34006	11.34006
COST OF ELEC-FUEL	10.0	14.15883	14.15883	14.15883	14.15883	14.15883
COST OF ELEC-OP & MAIN	10.0	14.15883	14.15883	14.15883	14.15883	14.15883
TOTAL COST OF ELEC	50.0	39.27943	40.24281	40.24281	40.24281	40.24281
ACCOUNT						
TOTAL DIRECT COSTS, \$	100.0	6.00	100.0	100.0	100.0	100.0
INDIRECT COSTS, \$	51.0	3054.98504	3054.98504	3054.98504	3054.98504	3054.98504
PROF & OWNER COSTS, \$	8.0	19657.418	19657.418	19657.418	19657.418	19657.418
CONTINGENCY COST, \$	8.0	2443.9880	2443.9880	2443.9880	2443.9880	2443.9880
SUB TOTAL, \$	67.0	37403.5680	37403.5680	37403.5680	37403.5680	37403.5680
ESCALATION COST, \$	12.0	7668.5605	7668.5605	7668.5605	7668.5605	7668.5605
TOTAL DURING CONST, \$	12.0	8833.2538	8833.2538	8833.2538	8833.2538	8833.2538
TOTAL CAPITALIZATION, \$	12.0	14.74386	14.74386	14.74386	14.74386	14.74386
COST OF ELEC-CAPITAL	12.0	11.34006	11.34006	11.34006	11.34006	11.34006
COST OF ELEC-FUEL	12.0	14.15883	14.15883	14.15883	14.15883	14.15883
COST OF ELEC-OP & MAIN	12.0	14.15883	14.15883	14.15883	14.15883	14.15883
TOTAL COST OF ELEC	50.0	39.27943	40.24281	40.24281	40.24281	40.24281
ACCOUNT						
TOTAL DIRECT COSTS, \$	100.0	6.00	100.0	100.0	100.0	100.0
INDIRECT COSTS, \$	51.0	3054.98504	3054.98504	3054.98504	3054.98504	3054.98504
PROF & OWNER COSTS, \$	8.0	19657.418	19657.418	19657.418	19657.418	19657.418
CONTINGENCY COST, \$	8.0	2443.9880	2443.9880	2443.9880	2443.9880	2443.9880
SUB TOTAL, \$	67.0	37403.5680	37403.5680	37403.5680	37403.5680	37403.5680
ESCALATION COST, \$	6.0	7668.5605	7668.5605	7668.5605	7668.5605	7668.5605
TOTAL DURING CONST, \$	12.0	8833.2538	8833.2538	8833.2538	8833.2538	8833.2538
TOTAL CAPITALIZATION, \$	12.0	14.74386	14.74386	14.74386	14.74386	14.74386
COST OF ELEC-CAPITAL	12.0	11.34006	11.34006	11.34006	11.34006	11.34006
COST OF ELEC-FUEL	12.0	14.15883	14.15883	14.15883	14.15883	14.15883
COST OF ELEC-OP & MAIN	12.0	14.15883	14.15883	14.15883	14.15883	14.15883
TOTAL COST OF ELEC	50.0	39.27943	40.24281	40.24281	40.24281	40.24281

Table A 13.4.17 Continued

SOLID ELECTROLYTE FUEL CELL PARAMETRIC POINT NO. 4 COST OF ELECTRICITY, MILLS/KW.HR

ACCOUNT	RATE, PERCENT	10.00	FIXED CHARGE RATE, PCT	25.00
TOTAL DIRECT COSTS, \$		305498504	305498504	305498504
INDIRECT COST, \$	51.0	19657418	19657418	19657418
PROF & OWNER COSTS, \$	8.0	24439880	24439880	24439880
CONTINGENCY COST, \$	8.0	24439880	24439880	24439880
SUB TOTAL, \$		374035680	374035680	374035680
ESCALATION COST, \$	6.5	76685605	76685605	76685605
TOTAL DURING CONST, \$		88332538	88332538	88332538
TOTAL CAPITALIZATION, \$	10.0	539053816	539053816	539053816
COST OF ELEC-CAPITAL	25.0	8	17.69263	20.47755
COST OF ELEC-FUEL	0.0	11.34006	11.34006	11.34006
COST OF ELEC-OP & MAIN	0.0	14.15888	14.15888	14.15888
TOTAL COST OF ELEC	0.0	33.68998	40.24281	45.97653

ACCOUNT	RATE, PERCENT	1.20	FUEL COST, \$/10*6 BTU	2.40
TOTAL DIRECT COSTS, \$		305498504	305498504	305498504
INDIRECT COST, \$	51.0	19657418	19657418	19657418
PROF & OWNER COSTS, \$	8.0	24439880	24439880	24439880
CONTINGENCY COST, \$	8.0	24439880	24439880	24439880
SUB TOTAL, \$		374035680	374035680	374035680
ESCALATION COST, \$	6.5	76685605	76685605	76685605
TOTAL DURING CONST, \$		88332538	88332538	88332538
TOTAL CAPITALIZATION, \$	10.0	539053816	539053816	539053816
COST OF ELEC-CAPITAL	18.0	14.74386	14.74386	14.74386
COST OF ELEC-FUEL	0.0	6.30404	11.34006	13.60807
COST OF ELEC-OP & MAIN	0.0	14.15888	14.15888	14.15888
TOTAL COST OF ELEC	0.0	35.70678	46.47984	52.51082

ACCOUNT	RATE, PERCENT	12.00	CAPACITY FACTOR, PERCENT	80.00
TOTAL DIRECT COSTS, \$		305498504	305498504	305498504
INDIRECT COST, \$	51.0	19657418	19657418	19657418
PROF & OWNER COSTS, \$	8.0	24439880	24439880	24439880
CONTINGENCY COST, \$	8.0	24439880	24439880	24439880
SUB TOTAL, \$		374035680	374035680	374035680
ESCALATION COST, \$	6.5	76685605	76685605	76685605
TOTAL DURING CONST, \$		88332538	88332538	88332538
TOTAL CAPITALIZATION, \$	10.0	539053816	539053816	539053816
COST OF ELEC-CAPITAL	18.0	79.86258	19.16702	11.97939
COST OF ELEC-FUEL	0.0	11.34006	11.34006	11.34006
COST OF ELEC-OP & MAIN	0.0	18.38695	14.44607	14.15888
TOTAL COST OF ELEC	0.0	109.58959	44.95315	37.29884

Table A 13.4.18

SOLID ELECTROLYTE FUEL CELL

ACCOUNT NO	AUX POWER, MWE	PERC PLANT POW	OPERATION COST	MAINTENANCE COST
4	.16603	2.02027	.03101	.88247
11	.00000	.00000	15100.00000	.00000
14	.00000	.00000	6.43781	.00000
18	8.05200	97.97973	.00000	.00000
TOTAL	8.21803	.71104	15258.15442	6300000.87500
SOLID ELECTROLYTE FUEL CELL				
NOMINAL POWER, MWE	1154.0000		BASE CASE INPUT	1155.7820
NOM HEAT RATE, BTU/KW-HR	5629.9998		NET POWER, MWE	5670.0311
ST TURB HEAT RATE CHANGE	.9942		NET HEAT RATE, BTU/KW-HR	
CONDENSER				
DESIGN PRESSURE, IN HG A	2.0000		NUMBER OF SHELLS	2.0000
NUMBER OF TUBES/SHELL	5.9017		TUBE LENGTH, FT	75.2339
U, BTU/HR-FI2-F	608.8535		TERMINAL TEMP DIFF, F	5.0000
HEAT REJECTION				
DESIGN TEMP, F	51.4000		APPROACH, F	21.6744
OFF DESIGN PRES, IN HG A	23.0000		OFF DESIGN TEMP, F	77.0000
	1.2126		LP TURBINE BLADE LEN, IN	25.0000
1	1164.000	2	.000	5
6	372.000	7	2.000	10
11	1.000	12	1899195.000	2.000
16	2.000	13	1.000	3.000
21	.500	18	3.000	5.000
26	.000	23	.000	.000
31	1.000	28	15000.000	.500
36	2198000.000	33	1.000	1.000
41	197000.000	38	15000.000	942000.000
46	.000	43	1.000	50000.000
51	.000	48	1.000	6.000
56	.400	53	.000	.000
6	.000	58	1.000	.000
11	.000	13	.000	.000
16	.000	18	.000	.000
21	.000	23	2412000.000	1.000
26	1.000	28	142000000.000	1800000.000
31	.000	33	18400000.000	9000000.000
36	6696000.000	38	1.000	1.000
41	.000	43	2700000.000	6700000.000
46	.000	48	.000	.000
51	.000	53	.000	.000
56	.400	58	.000	.000
				15100.000
				6300000.000

Table A 13.4.19

SOLID ELECTROLYTE FUEL CELL PARAMETRIC POINT NO.1E		ACCOUNT LISTING				
ACCOUNT NO. & NAME,	UNIT	AMOUNT	MAT \$/UNIT	INS \$/UNIT	MAT COST,\$	INS COST,\$
SITE DEVELOPMENT						
1. 1 LAND COST	ACRE	42.0	1000.00		42000.00	.00
1. 2 CLEARING LAND	ACRE	14.0	.00	600.00	.00	8399.15
1. 3 GRADING LAND	ACRE	42.0	.00	3000.00	.00	126000.00
1. 4 ACCESS RAILROAD	MILE	5.0	115000.00		575000.00	550000.00
1. 5 LOOP RAILROAD TRACK	MILE	1.0	120000.00	170000.00		.00
1. 6 SIDING R R TRACK	MILE	1.5	125000.00	80000.00		120000.00
1. 7 OTHER SITE COSTS	ACRE	0	.00	.00	106130.62	106130.62
PERCENT TOTAL DIRECT COST	IN ACCOUNT	1 = 1.376	ACCOUNT TOTAL,\$		904630.62	904630.62
EXCAVATION & PILING						
2. 1 COMMON EXCAVATION	YD3	36450.0	.00	3.00	.00	109350.00
2. 2 PILING	FT	0	6.50	8.50	.00	.00
PERCENT TOTAL DIRECT COST	IN ACCOUNT	2 = .083	ACCOUNT TOTAL,\$.00	109350.00
PLANT ISLAND CONCRETE						
3. 1 PLANT IS. CONCRETE	YD3	12150.0	70.00	80.00	850500.00	972000.00
3. 2 SPECIAL STRUCTURES	YD3	0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST	IN ACCOUNT	3 = 1.386	ACCOUNT TOTAL,\$		850500.00	972000.00
HEAT REJECTION SYSTEM						
4. 1 COOLING TOWERS	EACH	0	.00	.00	.00	.00
4. 2 CIRCULATING H2O SYS	EACH	1.0	.00	.00	.00	.00
4. 3 SURFACE CONDENSER	FT2	0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST	IN ACCOUNT	4 = .000	ACCOUNT TOTAL,\$.00	.00
STRUCTURAL FEATURES						
5. 1 STAT. STRUCTURAL ST.	TON	0	650.00	175.00	.00	.00
5. 2 SILOS & BUNKERS	TPH	64.7	1800.00	750.00	116381.81	43492.42
5. 3 CHIMNEY	FT	0	330000.00	70000.00	330000.00	70000.00
5. 4 STRUCTURAL FEATURES	EACH	1.0	.430	.00	446381.81	118492.42
PERCENT TOTAL DIRECT COST	IN ACCOUNT	5 =	ACCOUNT TOTAL,\$		446381.81	118492.42
BUILDINGS						
6. 1 STATION BUILDINGS	FT3	0	.16	.16	.00	.00
6. 2 ADMINISTRATION	FT2	10000.0	16.00	14.00	160000.00	140000.00
6. 3 WAREHOUSE & SHOP	FT2	10000.0	12.00	8.00	120000.00	80000.00
PERCENT TOTAL DIRECT COST	IN ACCOUNT	6 = .380	ACCOUNT TOTAL,\$		280000.00	220000.00
FUEL HANDLING & STORAGE						
7. 1 COAL HANDLING SYS	TPH	54.7	.00	.00	1083565.48	586629.56
7. 2 DOLOMITE HAND. SYS	TPH	34.2	.00	.00	610490.90	358623.95
7. 3 FUEL OIL HAND. SYS	GAL	25000.0	.00	.00	6727.20	5526.17
PERCENT TOTAL DIRECT COST	IN ACCOUNT	7 = 2.017	ACCOUNT TOTAL,\$		1700783.56	950779.67

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Table A 13.4.19 Continued

ACCOUNT LISTING		ACCOUNT LISTING					
SOLID ELECTROLYTE FUEL CELL PARAMETRIC POINT NO. 18		ACCOUNT LISTING					
ACCOUNT NO.	NAME	UNIT	AMOUNT	MAT \$/UNIT	INS \$/UNIT	MAT COST,\$	INS COST,\$
FUEL PROCESSING							
1.	COAL GASIFIER	EACH	1.0	42000000.00	15400000.00	42000000.00	15400000.00
2.	DESULFURIZER	EACH	1.0	5000000.00	1250000.00	5000000.00	1250000.00
3.	VENTURURI SCRUBBER	EACH	1.0	1910000.00	480000.00	1910000.00	480000.00
4.	COAL PRETREATMENT	EACH	1.0	1110000.00	280000.00	1110000.00	280000.00
5.	REFORMER	EACH	.0	.00	.00	.00	.00
6.	SHIFT CONVERTER	EACH	.0	.00	.00	.00	.00
7.	CO2 SCRUBBER	EACH	1.0	2340000.00	1000000.00	2340000.00	1000000.00
8.	BLOWER	EACH	3	52360000.00	18410000.00	52360000.00	18410000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			3	=52.839	ACCOUNT TOTAL,\$		
FIRING SYSTEM							
9.	1		.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			9	=	.000	ACCOUNT TOTAL,\$	
VAPOR GENERATOR (FIRED)							
10.	1		.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			10	=	.000	ACCOUNT TOTAL,\$	
ENERGY CONVERTER							
11.	1	FUEL CELL SYSTEM EACH	1.0	19000000.00	2500000.00	13800000.00	2500000.00
11.	2	STM TURBINE GENERATOR EA	.0	.00	.00	.00	.00
11.	3	OXYGEN PLANT EACH	.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			11	=16.204	ACCOUNT TOTAL,\$	10800000.00	2500000.00
COUPLING HEAT EXCHANGER							
12.	1	HEAT REC STEAM GEN EACH	1.0	2110000.00	90000.00	2110000.00	90000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			12	=	.229	ACCOUNT TOTAL,\$	
HEAT RECOVERY HEAT EXCH.							
13.	1	RECUPERATOR EACH	1.0	420000.00	153000.00	420000.00	153000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			13	=	4.359	ACCOUNT TOTAL,\$	
WATER TREATMENT							
14.	1	DEMINERALIZER GPM	.0	2500.00	700.00	.00	.00
14.	2	CONDENSATE POLISHING KWE	.0	1.25	.30	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			14	=	.000	ACCOUNT TOTAL,\$	
POWER CONDITIONING							
15.	1	POWER CONDITIONING EG EA	1.0	17500000.00	750000.00	17500000.00	750000.00
15.	2	STD TRANSFORMER KVA	.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			15	=13.884	ACCOUNT TOTAL,\$	17500000.00	750000.00
AUXILIARY MECH EQUIPMENT							
16.	1	BOILER FEED PUMP &DR KWE	.0	.55	.04	.00	.00
16.	2	OTHER PUMPS KWE	.0	.88	.72	.00	.00
16.	3	MISC SERVICE SYS KWE	279323.9	1.17	326812.52	203908.67	203908.67
16.	4	AUXILIARY BOILER FRH	.0	4.00	.80	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			16	=	.404	ACCOUNT TOTAL,\$	203908.67

Table A 13.4.19 Continued

SOLID ELECTROLYTE FUEL CELL
PARAMETRIC POINT NO.18

ACCOUNT LISTING

ACCOUNT NO. & NAME	UNIT	AMOUNT	MAT \$/UNIT	INS \$/UNIT	MAT COST.\$	INS COST.\$
PIPE & FITTINGS						
17. 1 CONVENTIONAL PIPING	ION	60.0	3000.00	1800.00	180000.00	108000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 17 =			.219	ACCOUNT TOTAL.\$	180000.00	108000.00
AUXILIARY ELEC EQUIPMENT						
18. 1 MISC MOTORS,ETC		83798.1	1.40	.17	117317.32	14285.67
18. 2 SWITCHGEAR & MCC PAN	KWE	279326.9	1.95	.45	544687.54	125697.12
18. 3 CONDUIT, CABLES, TRAYS	FT	400000.0	1.32	1.36	527999.99	543999.99
18. 4 ISOLATED PHASE BUS	FT		510.00	450.00		
18. 5 LIGHTING & COMMUN	KWE	279326.9	.35	.43	97764.43	120110.59
PERCENT TOTAL DIRECT COST IN ACCOUNT 18 =			1.591	ACCOUNT TOTAL.\$	1287769.27	804053.37
CONTROL, INSTRUMENTATION						
19. 1 COMPUTER	EACH	1.0	200000.00	6000.00	200000.00	6000.00
19. 2 OTHER CONTROLS	EACH	1.0	166000.00	100000.00	166000.00	100000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 19 =			.359	ACCOUNT TOTAL.\$	366000.00	106000.00
PROCESS WASTE SYSTEMS						
20. 1 BOTTOM ASH	TPH	.0	.00	.00	.00	.00
20. 2 DRY ASH	TPH	.0	.00	.00	.00	.00
20. 3 WET SLURRY	ACRE	34.2	1036072.06	259018.02	1036072.06	259018.02
20. 4 ONSITE DISPOSAL	ACRE	174.4	875.64	10105.84	1198916.50	1762170.44
PERCENT TOTAL DIRECT COST IN ACCOUNT 20 =			3.238	ACCOUNT TOTAL.\$	2234988.56	2021188.45
STACK GAS CLEANING						
21. 1 PRECIPITATOR	EACH	.0	1782899.77	1158884.84		
21. 2 SCRUBBER	KWE	.0	.00	.00	.00	.00
21. 3 MISC STEEL & DUCTS		.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 21 =			.000	ACCOUNT TOTAL.\$.00	.00
TOTAL DIRECT COSTS.\$					101648854.00	29798301.75

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Table A 13.4.20
SOLID ELECTROLYTE FUEL CELL PARAMETRIC POINT NO.18
COST OF ELECTRICITY, MILLS/KWH-HR

ACCOUNT	RATE, PERCENT	5.00	6.00	LABOR RATE, \$/HR	15.00	21.50
TOTAL DIRECT COSTS, \$	11.00	113515227	125543729	143819270	153822814	163824774
INDIRECT COSTS, \$	1.00	9602151	101862181	215053377	20324770	20324770
PROF & OWNER COSTS, \$	3.00	3481236	10147498	11505331	12397105	12397105
CONTINGENCY COST, \$	6.00	7110950	1752624	17636830	8725729	8725729
SUB TOTAL, \$	6.5	143710192	155306228	165046893	215505220	215505220
ESCALATION COST, \$	10.00	16941742	18270879	19437089	26417168	26417168
INTRST DURING CONST, \$	13.00	18244356	19716511	20953113	27371899	27371899
TOTAL CAPITALIZATION, \$	13.00	178238298	192331516	205457098	269394856	269394856
COST OF ELEC-CAPITAL	0.0	28.47306	28.47306	33.68505	33.68505	33.68505
COST OF ELEC-FUEL	0.0	5.47306	5.47306	5.47306	5.47306	5.47306
COST OF ELEC-OP & MAIN	0.0	12.24773	12.24773	12.24773	12.24773	12.24773
TOTAL COST OF ELEC	0.0	43.82342	45.92965	47.69888	51.40585	56.62205

ACCOUNT	RATE, PERCENT	5.00	6.00	CONTINGENCY, PERCENT	5.00	20.00
TOTAL DIRECT COSTS, \$	11.00	131447165	131447165	131447165	131447165	131447165
INDIRECT COST, \$	1.00	15197133	15197133	15197133	15197133	15197133
PROF & OWNER COSTS, \$	3.00	10515773	10515773	10515773	10515773	10515773
CONTINGENCY COST, \$	6.00	6572359	7838830	6572359	6572359	6572359
SUB TOTAL, \$	6.5	157160070	165046898	165046898	165046898	165046898
ESCALATION COST, \$	10.00	18527225	19437089	19437089	21626580	21626580
INTRST DURING CONST, \$	13.00	19117484	20953113	20953113	22289370	22289370
TOTAL CAPITALIZATION, \$	13.00	187437716	205457098	205457098	228365412	228365412
COST OF ELEC-CAPITAL	0.0	27.35181	28.47306	29.97809	33.32062	33.32062
COST OF ELEC-FUEL	0.0	5.47306	5.47306	5.47306	5.47306	5.47306
COST OF ELEC-OP & MAIN	0.0	12.24773	12.24773	12.24773	12.24773	12.24773
TOTAL COST OF ELEC	0.0	45.07261	46.26637	47.69888	51.04142	51.04142

ACCOUNT	RATE, PERCENT	5.00	6.00	ESCALATION RATE, PERCENT	10.00	12.50
TOTAL DIRECT COSTS, \$	11.00	131447165	131447165	131447165	131447165	131447165
INDIRECT COST, \$	1.00	15197133	15197133	15197133	15197133	15197133
PROF & OWNER COSTS, \$	3.00	10515773	10515773	10515773	10515773	10515773
CONTINGENCY COST, \$	6.00	7886830	7886830	7886830	7886830	7886830
SUB TOTAL, \$	6.5	165046898	165046898	165046898	165046898	165046898
ESCALATION COST, \$	10.00	19437089	19437089	19437089	22766274	22766274
INTRST DURING CONST, \$	13.00	20353113	21339370	21339370	2202194	2202194
TOTAL CAPITALIZATION, \$	13.00	205457098	210563516	210563516	226238522	226238522
COST OF ELEC-CAPITAL	0.0	29.97809	30.72502	31.73175	33.01760	33.01760
COST OF ELEC-FUEL	0.0	5.47306	5.47306	5.47306	5.47306	5.47306
COST OF ELEC-OP & MAIN	0.0	12.24773	12.24773	12.24773	12.24773	12.24773
TOTAL COST OF ELEC	0.0	46.96385	46.44382	49.05254	50.77839	50.77839

ACCOUNT	RATE, PERCENT	8.00	10.00	INT DURING CONST, PERCENT	12.50	15.00
TOTAL DIRECT COSTS, \$	11.00	131447165	131447165	131447165	131447165	131447165
INDIRECT COST, \$	1.00	15197133	15197133	15197133	15197133	15197133
PROF & OWNER COSTS, \$	3.00	10515773	10515773	10515773	10515773	10515773
CONTINGENCY COST, \$	6.00	7886830	7886830	7886830	7886830	7886830
SUB TOTAL, \$	6.5	165046898	165046898	165046898	165046898	165046898
ESCALATION COST, \$	10.00	19437089	19437089	19437089	19437089	19437089
INTRST DURING CONST, \$	13.00	16695512	205457098	205457098	21622419	21622419
TOTAL CAPITALIZATION, \$	13.00	195975045	201199498	201199498	21622419	21622419
COST OF ELEC-CAPITAL	0.0	26.47306	26.47306	26.47306	26.47306	26.47306
COST OF ELEC-FUEL	0.0	5.47306	5.47306	5.47306	5.47306	5.47306
COST OF ELEC-OP & MAIN	0.0	12.24773	12.24773	12.24773	12.24773	12.24773
TOTAL COST OF ELEC	0.0	46.46134	47.07766	47.69888	49.49231	49.49231

Table A 13.4.20 Continued
 SOLID ELECTROLYTE FUEL CELL PARAMETRIC POINT NO.18
 COST OF ELECTRICITY MILLS/KW.HR

ACCOUNT	PERCENT	1C.CC	FIXED CHARGE RATE, PCT	21.60	25.00
TOTAL DIRECT COSTS,\$.0	131447165.	18.00	131447165.	131447165.
INDIRECT COST,\$	51.0	15197133.	131447165.	15197133.	15197133.
PROF & OWNER COSTS,\$	8.0	10515773.	10515773.	10515773.	10515773.
CONTINGENCY COST,\$	6.0	7886830.	7886830.	7886830.	7886830.
SUB TOTAL,\$.0	165046898.	165046898.	165046898.	165046898.
ESCALATION COST,\$	6.5	19457089.	19457089.	19457089.	19457089.
INTRST DURING CONST,\$	10.0	20953113.	20953113.	20953113.	20953113.
TOTAL CAPITALIZATION,\$.0	205457098.	205457098.	205457098.	205457098.
COST OF ELEC-FUEL	25.0	16.65449	29.97809	35.97371	41.67323
COST OF ELEC-OP & MAIN	.0	5.47306	5.47306	5.47306	5.47306
TOTAL COST OF ELEC	.0	12.24773	12.24773	12.24773	12.24773
	.0	34.37529	47.69888	53.69450	59.35703

ACCOUNT	RATE, PERCENT	FUEL COST, \$/10**6 BTU	1.50	2.50	1.02
TOTAL DIRECT COSTS,\$.0	131447165.	131447165.	131447165.	131447165.
INDIRECT COST,\$	51.0	15197133.	15197133.	15197133.	15197133.
PROF & OWNER COSTS,\$	8.0	10515773.	10515773.	10515773.	10515773.
CONTINGENCY COST,\$	6.0	7886830.	7886830.	7886830.	7886830.
SUB TOTAL,\$.0	165046898.	165046898.	165046898.	165046898.
ESCALATION COST,\$	6.5	19457089.	19457089.	19457089.	19457089.
INTRST DURING CONST,\$	10.0	20953113.	20953113.	20953113.	20953113.
TOTAL CAPITALIZATION,\$.0	205457098.	205457098.	205457098.	205457098.
COST OF ELEC-FUEL	18.0	29.97809	29.97809	29.97809	29.97809
COST OF ELEC-OP & MAIN	.0	3.21945	9.65834	16.09723	6.56767
TOTAL COST OF ELEC	.0	12.24773	12.24773	12.24773	12.24773
	.0	45.44527	51.88416	58.32306	48.79349

ACCOUNT	RATE, PERCENT	CAPACITY FACTOR, PERCENT	45.00	65.00	80.00
TOTAL DIRECT COSTS,\$.0	131447165.	131447165.	131447165.	131447165.
INDIRECT COST,\$	51.0	15197133.	15197133.	15197133.	15197133.
PROF & OWNER COSTS,\$	8.0	10515773.	10515773.	10515773.	10515773.
CONTINGENCY COST,\$	6.0	7886830.	7886830.	7886830.	7886830.
SUB TOTAL,\$.0	165046898.	165046898.	165046898.	165046898.
ESCALATION COST,\$	6.5	19457089.	19457089.	19457089.	19457089.
INTRST DURING CONST,\$	10.0	20953113.	20953113.	20953113.	20953113.
TOTAL CAPITALIZATION,\$.0	205457098.	205457098.	205457098.	205457098.
COST OF ELEC-CAPITAL	18.0	162.38131	38.97151	29.97809	24.35720
COST OF ELEC-FUEL	.0	5.47306	5.47306	5.47306	5.47306
COST OF ELEC-OP & MAIN	.0	16.69476	12.54928	12.24773	12.05927
TOTAL COST OF ELEC	.0	184.54152	61.46921	47.69888	41.88952

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Table A 13.4.21

SOLID ELECTROLYTE FUEL CELL

ACCOUNT NO	AUX POWER, MWE	PERC PLANT POW	OPERATION COST	MAINTENANCE COST
7	.94372	40.27445	183.02236	.00000
8	.08405	3.58707	.00000	.00000
11	.09500	46.73031	2150.00000	1240000.00000
18	1.22046	9.40817	1.13303	.00000
20	2.34323	1.08154	2435.78137	1240000.00000
TOTALS				
SOLID ELECTROLYTE FUEL CELL				
NOMINAL POWER, MWE 219.0000				
NOM HEAT RATE, BTU/KW-HR 6369.9998				
ST TURB HEAT RATE CHANGE .0000				
CONDENSER				
DESIGN PRESSURE, IN HG A 2.0000				
NUMBER OF TUBES/SHELL 608.8535				
U, BTU/HR-FI2-F				
HEAT REJECTION				
DESIGN TEMP, F				
RANGE, F				
OFF DESIGN PRES, IN HG A				
1	219.000	2.000	.000	3
6	.000	7	.000	8
11	.000	12	.000	13
16	.000	17	.000	18
21	1.500	22	12150.000	23
26	.000	27	10000.000	28
31	1.000	32	60.000	33
36	.000	37	.000	38
41	70000.000	42	200000.000	43
46	.000	47	.000	48
51	5.350	52	5.350	53
1	1.000	2	1.000	3
6	.000	7	.000	8
11	5000000.000	12	1250000.000	13
16	2800000.000	17	.000	18
21	.000	22	.000	23
26	.000	27	.000	28
31	.000	32	.000	33
36	30000.000	37	1.000	38
41	.000	42	.000	43
46	.000	47	.000	48
51	.000	52	.000	53
1	1.000	2	1.000	3
6	.000	7	.000	8
11	5000000.000	12	1250000.000	13
16	2800000.000	17	.000	18
21	.000	22	.000	23
26	.000	27	.000	28
31	.000	32	.000	33
36	30000.000	37	1.000	38
41	.000	42	.000	43
46	.000	47	.000	48
51	.000	52	.000	53
1	1.000	2	1.000	3
6	.000	7	.000	8
11	5000000.000	12	1250000.000	13
16	2800000.000	17	.000	18
21	.000	22	.000	23
26	.000	27	.000	28
31	.000	32	.000	33
36	30000.000	37	1.000	38
41	.000	42	.000	43
46	.000	47	.000	48
51	.000	52	.000	53
1	1.000	2	1.000	3
6	.000	7	.000	8
11	5000000.000	12	1250000.000	13
16	2800000.000	17	.000	18
21	.000	22	.000	23
26	.000	27	.000	28
31	.000	32	.000	33
36	30000.000	37	1.000	38
41	.000	42	.000	43
46	.000	47	.000	48
51	.000	52	.000	53
1	1.000	2	1.000	3
6	.000	7	.000	8
11	5000000.000	12	1250000.000	13
16	2800000.000	17	.000	18
21	.000	22	.000	23
26	.000	27	.000	28
31	.000	32	.000	33
36	30000.000	37	1.000	38
41	.000	42	.000	43
46	.000	47	.000	48
51	.000	52	.000	53
1	1.000	2	1.000	3
6	.000	7	.000	8
11	5000000.000	12	1250000.000	13
16	2800000.000	17	.000	18
21	.000	22	.000	23
26	.000	27	.000	28
31	.000	32	.000	33
36	30000.000	37	1.000	38
41	.000	42	.000	43
46	.000	47	.000	48
51	.000	52	.000	53
1	1.000	2	1.000	3
6	.000	7	.000	8
11	5000000.000	12	1250000.000	13
16	2800000.000	17	.000	18
21	.000	22	.000	23
26	.000	27	.000	28
31	.000	32	.000	33
36	30000.000	37	1.000	38
41	.000	42	.000	43
46	.000	47	.000	48
51	.000	52	.000	53
1	1.000	2	1.000	3
6	.000	7	.000	8
11	5000000.000	12	1250000.000	13
16	2800000.000	17	.000	18
21	.000	22	.000	23
26	.000	27	.000	28
31	.000	32	.000	33
36	30000.000	37	1.000	38
41	.000	42	.000	43
46	.000	47	.000	48
51	.000	52	.000	53
1	1.000	2	1.000	3
6	.000	7	.000	8
11	5000000.000	12	1250000.000	13
16	2800000.000	17	.000	18
21	.000	22	.000	23
26	.000	27	.000	28
31	.000	32	.000	33
36	30000.000	37	1.000	38
41	.000	42	.000	43
46	.000	47	.000	48
51	.000	52	.000	53
1	1.000	2	1.000	3
6	.000	7	.000	8
11	5000000.000	12	1250000.000	13
16	2800000.000	17	.000	18
21	.000	22	.000	23
26	.000	27	.000	28
31	.000	32	.000	33
36	30000.000	37	1.000	38
41	.000	42	.000	43
46	.000	47	.000	48
51	.000	52	.000	53
1	1.000	2	1.000	3
6	.000	7	.000	8
11	5000000.000	12	1250000.000	13
16	2800000.000	17	.000	18
21	.000	22	.000	23
26	.000	27	.000	28
31	.000	32	.000	33
36	30000.000	37	1.000	38
41	.000	42	.000	43
46	.000	47	.000	48
51	.000	52	.000	53
1	1.000	2	1.000	3
6	.000	7	.000	8
11	5000000.000	12	1250000.000	13
16	2800000.000	17	.000	18
21	.000	22	.000	23
26	.000	27	.000	28
31	.000	32	.000	33
36	30000.000	37	1.000	38
41	.000	42	.000	43
46	.000	47	.000	48
51	.000	52	.000	53
1	1.000	2	1.000	3
6	.000	7	.000	8
11	5000000.000	12	1250000.000	13
16	2800000.000	17	.000	18
21	.000	22	.000	23
26	.000	27	.000	28
31	.000	32	.000	33
36	30000.000	37	1.000	38
41	.000	42	.000	43
46	.000	47	.000	48
51	.000	52	.000	53
1	1.000	2	1.000	3
6	.000	7	.000	8
11	5000000.000	12	1250000.000	13
16	2800000.000	17	.000	18
21	.000	22	.000	23
26	.000	27	.000	28
31	.000	32	.000	33
36	30000.000	37	1.000	38
41	.000	42	.000	43
46	.000	47	.000	48
51	.000	52	.000	53
1	1.000	2	1.000	3
6	.000	7	.000	8
11	5000000.000	12	1250000.000	13
16	2800000.000	17	.000	18
21	.000	22	.000	23
26	.000	27	.000	28
31	.000	32	.000	33
36	30000.000	37	1.000	38
41	.000	42	.000	43
46	.000	47	.000	48
51	.000	52	.000	53
1	1.000	2	1.000	3
6	.000	7	.000	8
11	5000000.000	12	1250000.000	13
16	2800000.000	17	.000	18
21	.000	22	.000	23
26	.000	27	.000	28
31	.000	32	.000	33
36	30000.000	37	1.000	38
41	.000	42	.000	43
46	.000	47	.000	48
51	.000	52	.000	53
1	1.000	2	1.000	3
6	.000	7	.000	8
11	5000000.000	12	1250000.000	13
16	2800000.000	17	.000	18
21	.000	22	.000	23
26	.000	27	.000	28
31	.000	32	.000	33
36	30000.000	37	1.000	38
41	.000	42	.000	43
46	.000	47	.000	48
51	.000	52	.000	53
1	1.000	2	1.000	3
6	.000	7	.000	8
11	5000000.000	12	1250000.000	13
16	2800000.000	17	.000	18
21	.000	22	.000	23
26	.000	27	.000	28
31	.000	32	.000	33
36	30000.000	37	1.000	38
41	.000	42	.000	43
46	.000	47	.000	48
51	.000	52	.000	53
1	1.000	2	1.000	3
6	.000	7	.000	8
11	5000000.000	12	1250000.000	13
16	2800000.000	17	.000	18
21	.000	22	.000	23
26	.000	27	.000	28
31	.000	32	.000	33
36	30000.000	37	1.000	38
41	.000	42	.000	43
46	.000	47	.000	48
51	.000	52	.000	53
1	1.000	2	1.000	3
6	.000	7	.000	8
11	5000000.000			

Table A 13.4.4.22

MOLTEN CARBONATE FUEL CELL SYSTEM
PARAMETRIC POINT NO. 4 ACCOUNT LISTING

ACCOUNT NO. & NAME,	UNIT	AMOUNT	MAT \$/UNIT	INS \$/UNIT	MAT COST,\$	INS COST,\$
SITE DEVELOPMENT						
1. 1 LAND COST	ACRE	50.0	1000.00	.00	50000.00	.00
1. 2 CLEARING LAND	ACRE	16.7	.00	.00	.00	9999.00
1. 3 GRADING LAND	ACRE	50.0	.00	3000.00	.00	150000.00
1. 4 ACCESS RAILROAD	MILE	5.0	115000.00	110000.00	575000.00	550000.00
1. 5 LOOP RAILROAD TRACK	MILE	.5	120000.00	70000.00	.00	.00
1. 6 SIDING R R TRACK	MILE	.5	125000.00	80000.00	62500.00	40000.00
1. 7 OTHER SITE COSTS	ACRE	.0	.00	.00	118248.90	118248.90
PERCENT TOTAL DIRECT COST IN ACCOUNT 1 =			.514	ACCOUNT TOTAL,\$	805748.89	958247.88
EXCAVATION & PILING						
2. 1 COMMON EXCAVATION	YD3	185100.0	.00	3.00	.00	555300.00
2. 2 PILING	FT	.0	6.50	8.50	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 2 =			.171	ACCOUNT TOTAL,\$.00	555300.00
PLANT ISLAND CONCRETE						
3. 1 PLANT IS. CONCRETE	YD3	51700.0	70.00	80.00	4319000.00	4936000.00
3. 2 SPECIAL STRUCTURES	YD3	.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 3 =			2.843	ACCOUNT TOTAL,\$	4319000.00	4936000.00
HEAT REJECTION SYSTEM						
4. 1 COOLING TOWERS	EACH	1.0	.00	.00	153500.00	76500.00
4. 2 CIRCULATING H2O SYS	EACH	1.0	.00	.00	640.55	858.90
4. 3 SURFACE CONDENSER	FT2	233.7	.00	.00	283714.13	153.62
PERCENT TOTAL DIRECT COST IN ACCOUNT 4 =			.158	ACCOUNT TOTAL,\$	437854.68	77522.51
STRUCTURAL FEATURES						
5. 1 STAY STRUCTURAL ST. ION	TON	.0	650.00	175.00	.00	.00
5. 2 SILOS & BUNKERS	TPH	.0	1800.00	750.00	.00	.00
5. 3 CHIMNEY	FT	.0	.00	.00	.00	.00
5. 4 STRUCTURAL FEATURES EACH	EACH	1.0	982000.00	217000.00	982000.00	217000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 5 =			.368	ACCOUNT TOTAL,\$	982000.00	217000.00
BUILDINGS						
6. 1 STATION BUILDINGS	FT2	.0	.16	.16	.00	.00
6. 2 ADMINISTRATION	FT2	10000.0	16.00	14.00	160000.00	140000.00
6. 3 WAREHOUSE & SHOP	FT2	15000.0	12.00	8.00	190000.00	120000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 6 =			.184	ACCOUNT TOTAL,\$	340000.00	260000.00
FUEL HANDLING & STORAGE						
7. 1 COAL HANDLING SYS	IPH	.0	.00	.00	.00	.00
7. 2 DOLDRITE HAND. SYS	TPH	.0	.00	.00	.00	.00
7. 3 FUEL OIL HAND. SYS	GAL	.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 7 =			.000	ACCOUNT TOTAL,\$.00	.00

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Table A 13.4.22 Continued

MOLIER CARBONATE FUEL CELL SYSTEM PARAMETRIC POINT NO. 4		ACCOUNT LISTING				
ACCOUNT NO.	NAME	UNIT	AMOUNT	INS \$/UNIT	MAT COST,\$	INS COST,\$
FUEL PROCESSING						
1	COAL GASIFIER	EACH	1.00		.00	.00
2	SULFURIZER	EACH	1.00		.00	.00
3	VENTURI SCRAMBLER	EACH	1.00		.00	.00
4	COAL PRETREATMENT	EACH	1.00		.00	.00
5	REFORMER	EACH	1.00		.00	.00
6	SHIFT CONVERTER	EACH	1.00		.00	.00
7	CO2 SCRUBBER	EACH	1.00		.00	.00
8	SLOWER	EACH	1.00		.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			8		230000.00	230000.00
PERCENT TOTAL ACCOUNT TOTAL			8		230000.00	230000.00
FIRING SYSTEM						
9	1				.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			9		.00	.00
PERCENT TOTAL ACCOUNT TOTAL			9		.00	.00
VAPOUR GENERATOR (FIRE)						
10	1				.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			10		.00	.00
PERCENT TOTAL ACCOUNT TOTAL			10		.00	.00
ENERGY CONVERTER						
11	1	FUEL CELL SYSTEM	1.00		720000.00	720000.00
11.1	1	FUEL CELL SYSTEM	1.00		117000.00	117000.00
11.2	2	SPM TURBINE GENERATOR	2.00		117000.00	117000.00
11.3	3	OXYGEN PLANT	3.00		117000.00	117000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			11		1170000.00	1170000.00
PERCENT TOTAL ACCOUNT TOTAL			11		1170000.00	1170000.00
COUPLING HEAT EXCHANGER						
12	1	HEAT REC STEAM GEN	1.00		404000.00	404000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			12		404000.00	404000.00
PERCENT TOTAL ACCOUNT TOTAL			12		404000.00	404000.00
HEAT RECOVERY HEAT EXCH.						
13	1	RECUPERATOR	1.00		628000.00	628000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			13		628000.00	628000.00
PERCENT TOTAL ACCOUNT TOTAL			13		628000.00	628000.00
WATER TREATMENT						
14	1	DEMINERALIZER	1.00		191000.00	191000.00
14.1	1	DEMINERALIZER	1.00		191000.00	191000.00
14.2	2	CONDENSATE POLISHING	2.00		191000.00	191000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			14		382000.00	382000.00
PERCENT TOTAL ACCOUNT TOTAL			14		382000.00	382000.00
POWER CONDITIONING						
15	1	STD TRAN FORMER	1.00		270000.00	270000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT			15		270000.00	270000.00
PERCENT TOTAL ACCOUNT TOTAL			15		270000.00	270000.00
AUXILIARY MECH EQUIPMENT						
16	1	BOILER FEED PUMP	1.00		655425.05	655425.05
16.1	1	BOILER FEED PUMP	1.00		655425.05	655425.05
16.2	3	OTHER PUMPS	3.00		1373201.00	1373201.00
16.3	3	MISC SERVICE SYS	3.00		1373201.00	1373201.00
16.4	4	AUXILIARY BOILER	4.00		2577043.59	2577043.59
PERCENT TOTAL DIRECT COST IN ACCOUNT			16		5554870.64	5554870.64
PERCENT TOTAL ACCOUNT TOTAL			16		5554870.64	5554870.64

Table A 13.4.22 Continued

MOLTEN CARBONATE FUEL CELL SYSTEM PARAMETRIC POINT NO. 4		ACCOUNT LISTING				
ACCOUNT NO. & NAME,	UNIT	AMOUNT	MAT \$/UNIT	INS \$/UNIT	MAT COST.\$	INS COST.\$
PIPE & FITTINGS						
17. 1 CONVENTIONAL PIPING	TON	770.0	3000.00	1300.00	2310000.00	1386000.00
PERCENT TOTAL DIRECT COST	IN ACCOUNT 17 =	1.135	ACCOUNT TOTAL \$		2310000.00	1386000.00
AUXILIARY ELEC EQUIPMENT						
18. 1 MISC MOTOR/ELECT		586838.1	1.40	.17	821573.29	99762.47
18. 2 SWITCHGEAR & MCC PAN	KWE	1173676.1	1.95	.45	2288668.44	528154.25
18. 3 CONDUIT, CABLES, TRAYS	FT	2358000.0	1.32	1.35	3112559.97	3206879.97
18. 4 ISOLATED PHASE BUS	3000.0		510.00	450.00	1530000.00	1350000.00
18. 5 LIGHTING & COMMUN	KWE	1173676.1	.35	.43	410786.64	504680.73
PERCENT TOTAL DIRECT COST	IN ACCOUNT 18 =	3.459	ACCOUNT TOTAL \$		6786588.31	4474477.37
CONTROL, INSTRUMENTATION						
19. 1 COMPUTER	EACH		560000.00	15000.00	560000.00	15000.00
19. 2 OTHER CONTROLS	EACH		722000.00	435000.00	722000.00	435000.00
PERCENT TOTAL DIRECT COST	IN ACCOUNT 19 =	1.0	.563	ACCOUNT TOTAL \$	1382000.00	450000.00
PROCESS WASTE SYSTEMS						
20. 1 BOTTOM ASH	TPH		.00	.00	.00	.00
20. 2 DRY ASH	TPH		.00	.00	.00	.00
20. 3 WET SLURRY	TPH		.00	.00	.00	.00
20. 4 ONSITE DISPOSAL	ACRE		7676.49	11070.89	.00	.00
PERCENT TOTAL DIRECT COST	IN ACCOUNT 20 =		.000	ACCOUNT TOTAL \$.00	.00
STACK GAS CLEANING						
21. 1 PRECIPITATOR	EACH		.00	.00	.00	.00
21. 2 SCRUBBER	KWE		.00	.00	.00	.00
21. 3 MISC STEEL & DUCTS			.00	.00	.00	.00
PERCENT TOTAL DIRECT COST	IN ACCOUNT 21 =		.000	ACCOUNT TOTAL \$.00	.00
TOTAL DIRECT COSTS.\$					290771956.00	24017045.00

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Table A 13.4.23
POLYMER CRYSTALLINE CELL VOLTAGE PARALLEL POINT NO. 4 COST OF ELECTRICITY, MILLS/KW·HR

ACCOUNT	RATE, PERCENT	LABOR RATE, \$/HR	CONTINGENCY, PERCENT	ESCALATION RATE, PERCENT	INSTRUMENT DURING CONST, PERCENT
TOTAL DIRECT COSTS, \$	100.00	15.00	10.00	10.00	10.00
INDIRECT COSTS, \$	11.97	17.75	17.75	17.75	17.75
PROF OWNER COSTS, \$	1.00	1.00	1.00	1.00	1.00
CONTINGENCY COST, \$	24.97	24.97	24.97	24.97	24.97
SUB TOTAL, \$	141.94	141.94	141.94	141.94	141.94
INSTRUMENT DURING CONST, \$	11.97	11.97	11.97	11.97	11.97
TOTAL CAPITALIZATION, \$	153.91	153.91	153.91	153.91	153.91
COST OF ELEC-CAP, \$	11.97	11.97	11.97	11.97	11.97
COST OF ELEC-OP & MAINT, \$	11.97	11.97	11.97	11.97	11.97
TOTAL COST OF ELEC, \$	23.94	23.94	23.94	23.94	23.94
TOTAL COST OF ELEC, \$	167.85	167.85	167.85	167.85	167.85
TOTAL DIRECT COSTS, \$	100.00	15.00	10.00	10.00	10.00
INDIRECT COSTS, \$	11.97	17.75	17.75	17.75	17.75
PROF OWNER COSTS, \$	1.00	1.00	1.00	1.00	1.00
CONTINGENCY COST, \$	24.97	24.97	24.97	24.97	24.97
SUB TOTAL, \$	141.94	141.94	141.94	141.94	141.94
INSTRUMENT DURING CONST, \$	11.97	11.97	11.97	11.97	11.97
TOTAL CAPITALIZATION, \$	153.91	153.91	153.91	153.91	153.91
COST OF ELEC-CAP, \$	11.97	11.97	11.97	11.97	11.97
COST OF ELEC-OP & MAINT, \$	11.97	11.97	11.97	11.97	11.97
TOTAL COST OF ELEC, \$	23.94	23.94	23.94	23.94	23.94
TOTAL COST OF ELEC, \$	167.85	167.85	167.85	167.85	167.85
TOTAL DIRECT COSTS, \$	100.00	15.00	10.00	10.00	10.00
INDIRECT COSTS, \$	11.97	17.75	17.75	17.75	17.75
PROF OWNER COSTS, \$	1.00	1.00	1.00	1.00	1.00
CONTINGENCY COST, \$	24.97	24.97	24.97	24.97	24.97
SUB TOTAL, \$	141.94	141.94	141.94	141.94	141.94
INSTRUMENT DURING CONST, \$	11.97	11.97	11.97	11.97	11.97
TOTAL CAPITALIZATION, \$	153.91	153.91	153.91	153.91	153.91
COST OF ELEC-CAP, \$	11.97	11.97	11.97	11.97	11.97
COST OF ELEC-OP & MAINT, \$	11.97	11.97	11.97	11.97	11.97
TOTAL COST OF ELEC, \$	23.94	23.94	23.94	23.94	23.94
TOTAL COST OF ELEC, \$	167.85	167.85	167.85	167.85	167.85

Table A 13.4.23 Continued
MOLTEN CARBONATE FUEL CELL SYSTEM COST OF ELECTRICITY, MILLS/KW·HR
PARAMETRIC POINT NO. 4

ACCOUNT	RATE, PERCENT	10.00	14.40	18.00	25.00
TOTAL DIRECT COSTS, \$.0	325589000.	325589000.	325589000.	325589000.
INDIRECT COST, \$	51.0	17756692.	17756692.	17756692.	17756692.
PROF & OWNER COSTS, \$	8.0	26047120.	26047120.	26047120.	26047120.
CONTINGENCY COST, \$	8.0	26047120.	26047120.	26047120.	26047120.
SUB TOTAL, \$.0	395439932.	395439932.	395439932.	395439932.
ESCALATION COST, \$	6.5	81073951.	81073951.	81073951.	81073951.
TOTAL CAPITALIZATION, \$	10.0	93387382.	93387382.	93387382.	93387382.
COST OF ELEC-CAPITAL	.0	569901256.	569901256.	569901256.	569901256.
COST OF ELEC-FUEL	25.0	12.19904	15.24880	18.29856	21.17889
COST OF ELEC-OP & MAIN	.0	12.55007	12.55007	12.55007	12.55007
TOTAL COST OF ELEC	.0	16.05295	16.05295	16.05295	16.05295
	.0	37.07457	40.80205	43.85181	49.78190

ACCOUNT	RATE, PERCENT	1.20	2.00	3.10	4.40
TOTAL DIRECT COSTS, \$.0	325589000.	325589000.	325589000.	325589000.
INDIRECT COST, \$	51.0	17756692.	17756692.	17756692.	17756692.
PROF & OWNER COSTS, \$	8.0	26047120.	26047120.	26047120.	26047120.
CONTINGENCY COST, \$	8.0	26047120.	26047120.	26047120.	26047120.
SUB TOTAL, \$.0	395439932.	395439932.	395439932.	395439932.
ESCALATION COST, \$	6.5	81073951.	81073951.	81073951.	81073951.
TOTAL CAPITALIZATION, \$	10.0	93387382.	93387382.	93387382.	93387382.
COST OF ELEC-CAPITAL	.0	569901256.	569901256.	569901256.	569901256.
COST OF ELEC-FUEL	18.0	15.24880	15.24880	15.24880	15.24880
COST OF ELEC-OP & MAIN	.0	12.55007	12.55007	10.04005	15.09008
TOTAL COST OF ELEC	.0	16.05295	16.05295	16.05295	16.05295
	.0	38.83179	43.85181	41.34180	46.36182

ACCOUNT	RATE, PERCENT	45.00	50.00	55.00	60.00
TOTAL DIRECT COSTS, \$.0	325589000.	325589000.	325589000.	325589000.
INDIRECT COST, \$	51.0	17756692.	17756692.	17756692.	17756692.
PROF & OWNER COSTS, \$	8.0	26047120.	26047120.	26047120.	26047120.
CONTINGENCY COST, \$	8.0	26047120.	26047120.	26047120.	26047120.
SUB TOTAL, \$.0	395439932.	395439932.	395439932.	395439932.
ESCALATION COST, \$	6.5	81073951.	81073951.	81073951.	81073951.
TOTAL CAPITALIZATION, \$	10.0	93387382.	93387382.	93387382.	93387382.
COST OF ELEC-CAPITAL	.0	569901256.	569901256.	569901256.	569901256.
COST OF ELEC-FUEL	18.0	22.02604	19.82344	15.24880	12.39965
COST OF ELEC-OP & MAIN	.0	12.55007	12.55007	12.55007	12.55007
TOTAL COST OF ELEC	.0	16.42418	16.30353	16.05295	15.89634
	.0	51.00028	48.67703	43.85181	40.83605

Table A 13.4.25 PHOSPHORIC ACID FUEL CELL SYSTEM SUMMARY PLANT RESULTS

PARAMETRIC POINT	1	2	3	4	5	6	7	8
THERMODYNAMIC EFF	.000	.000	.000	.000	.000	.000	.000	.000
POWER PLANT EFF	.355	.351	.351	.348	.296	.355	.355	.355
OVERALL ENERGY EFF	.238	.236	.236	.293	.198	.238	.238	.238
CAP COST MILLION \$	10.436	39.516	37.614	357.206	15.355	10.436	10.436	10.436
CAP COST MILLION \$	448.243	428.885	423.778	440.395	833.006	448.243	448.243	448.243
COE CAPITAL	14.170	13.558	13.397	13.922	26.333	14.170	14.170	14.170
COE FUEL	25.012	25.268	25.268	19.590	29.946	25.012	25.012	25.012
COE OP & MAIN	10.893	10.910	10.793	10.464	9.919	4.580	3.291	2.346
COST OF ELECTRIC	50.076	48.736	49.458	43.976	66.198	43.762	42.473	41.528
EST TIME OF CONST	1.500	2.000	2.500	4.000	1.500	1.500	1.500	1.500

PARAMETRIC POINT	9	10	11	12	13	14	15	16
THERMODYNAMIC EFF	.000	.000	.000	.000	.000	.000	.000	.000
POWER PLANT EFF	.355	.355	.355	.355	.355	.358	.360	.299
OVERALL ENERGY EFF	.238	.239	.239	.239	.239	.249	.242	.201
CAP COST MILLION \$	10.788	8.771	7.900	8.642	8.064	10.904	10.236	116.549
CAP COST MILLION \$	463.336	376.725	335.306	371.160	346.364	468.332	439.640	554.590
CAPITAL COST \$/KWE	14.647	11.909	10.726	11.733	10.949	14.805	13.898	18.796
COE CAPITAL	25.012	25.012	25.012	25.012	25.012	25.752	24.658	29.692
COE FUEL	2.437	7.973	6.469	9.863	9.562	10.991	10.833	9.244
COE OP & MAIN	22.096	44.894	42.208	46.608	45.523	51.548	49.390	57.732
COST OF ELECTRIC	1.500	1.500	1.500	1.500	1.500	1.500	1.500	2.500
EST TIME OF CONST	1.500	1.500	1.500	1.500	1.500	1.500	1.500	1.500

PARAMETRIC POINT	17	18	19	20	21	22	23	24
THERMODYNAMIC EFF	.000	.000	.000	.000	.000	.000	.000	.000
POWER PLANT EFF	.000	.000	.000	.000	.000	.000	.000	.000
OVERALL ENERGY EFF	.000	.000	.000	.000	.000	.000	.000	.000
CAP COST MILLION \$.000	.000	.000	.000	.000	.000	.000	.000
CAPITAL COST \$/KWE	.000	.000	.000	.000	.000	.000	.000	.000
COE CAPITAL	.000	.000	.000	.000	.000	.000	.000	.000
COE FUEL	.000	.000	.000	.000	.000	.000	.000	.000
COE OP & MAIN	.000	.000	.000	.000	.000	.000	.000	.000
COST OF ELECTRIC	.000	.000	.000	.000	.000	.000	.000	.000
EST TIME OF CONST	.000	.000	.000	.000	.000	.000	.000	.000

Table A 13.4.27 PHOSPHORIC ACID FUEL CELL SYSTEM NATURAL RESOURCE REQUIREMENTS

	1	2	3	4	5	6	7	8	9
PARAMETRIC POINT	1.32491	1.33847	1.33847	1.08042	1.58623	1.32491	1.32491	1.32491	1.32491
COAL, LB/KW-HR	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SORBANT OR SEED, LB/KW-HR	.000	.000	.000	.000	.000	.000	.000	.000	.000
TOTAL WATER, GAL/KW-HR	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
COOLING WATER	.000	.000	.000	.000	.000	.000	.000	.000	.000
GASIFIER PROCESS H2O	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
WASTE HANDLING SLURRY	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SCRUBBER WASTE WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
NOX SUPPRESSION	5.44	4.88	3.91	3.60	13.05	6.44	6.44	6.44	6.44
TOTAL LAND ACRES/100MWE	6.44	4.88	3.91	3.60	13.05	6.44	6.44	6.44	6.44
MAIN PLANT	.00	.00	.00	.00	.00	.00	.00	.00	.00
DISPOSAL LAND	.00	.00	.00	.00	.00	.00	.00	.00	.00
LAND FOR ACCESS RR	.00	.00	.00	.00	.00	.00	.00	.00	.00

	10	11	12	13	14	15	16
PARAMETRIC POINT	1.32491	1.32491	1.32491	1.32491	1.26844	1.30622	1.57281
COAL, LB/KW-HR	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SORBANT OR SEED, LB/KW-HR	.000	.000	.000	.000	.000	.000	.000
TOTAL WATER, GAL/KW-HR	.00000	.00000	.00000	.00000	.00000	.00000	.00000
COOLING WATER	.000	.000	.000	.000	.000	.000	.000
GASIFIER PROCESS H2O	.00000	.00000	.00000	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP	.00000	.00000	.00000	.00000	.00000	.00000	.00000
WASTE HANDLING SLURRY	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SCRUBBER WASTE WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000
NOX SUPPRESSION	6.44	4.29	6.44	6.44	10.74	6.44	7.85
TOTAL LAND ACRES/100MWE	6.44	4.29	6.44	6.44	10.74	6.44	7.85
MAIN PLANT	.00	.00	.00	.00	.00	.00	.00
DISPOSAL LAND	.00	.00	.00	.00	.00	.00	.00
LAND FOR ACCESS RR	.00	.00	.00	.00	.00	.00	.00

	17	18	19	20	21	22	23	24
PARAMETRIC POINT	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000
COAL, LB/KW-HR	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SORBANT OR SEED, LB/KW-HR	.000	.000	.000	.000	.000	.000	.000	.000
TOTAL WATER, GAL/KW-HR	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
COOLING WATER	.000	.000	.000	.000	.000	.000	.000	.000
GASIFIER PROCESS H2O	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
WASTE HANDLING SLURRY	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SCRUBBER WASTE WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
NOX SUPPRESSION	.00	.00	.00	.00	.00	.00	.00	.00
TOTAL LAND ACRES/100MWE	.00	.00	.00	.00	.00	.00	.00	.00
MAIN PLANT	.00	.00	.00	.00	.00	.00	.00	.00
DISPOSAL LAND	.00	.00	.00	.00	.00	.00	.00	.00
LAND FOR ACCESS RR	.00	.00	.00	.00	.00	.00	.00	.00

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SUMMARY PLANT RESULTS

Table A 13.4.28ALKALINE FUEL CELL SYSTEM

PARAMETRIC POINT	1	.000	2	.000	3	.000	4	.000	5	.000	6	.000	7	.000	8	.000
THERMODYNAMIC EFF		.381		.377		.377		.366		.318		.381		.381		.381
POWER PLANT EFF		.256		.254		.254		.307		.214		.256		.256		.256
OVERALL ENERGY EFF		13.577		51.872		129.826		536.268		16.649		13.577		13.577		13.577
CAP COST MILLION \$		620.243		599.230		598.525		699.952		890.038		620.243		620.243		620.243
CAPITAL COST \$/KWE		19.607		18.943		18.921		22.127		28.136		19.607		19.607		19.607
COE CAPITAL		23.311		23.552		23.552		18.665		27.871		23.311		23.311		23.311
COE FUEL		17.837		17.936		17.939		18.077		11.179		7.193		5.086		3.987
COE OP & MAIN		60.756		60.430		60.412		58.869		67.186		50.111		47.964		46.365
COST OF ELECTRIC		1.500		2.000		2.500		4.000		1.500		1.500		1.500		1.500
EST TIME OF CONST																

PARAMETRIC POINT	9	.000	10	.000	11	.000	12	.000	13	.000	14	.000	15	.000	16	.000
THERMODYNAMIC EFF		.381		.381		.381		.381		.381		.381		.381		.381
POWER PLANT EFF		.256		.256		.256		.256		.256		.256		.256		.256
OVERALL ENERGY EFF		13.577		10.941		9.812		11.569		11.318		13.577		13.577		13.577
CAP COST MILLION \$		636.297		499.840		448.239		528.507		517.640		636.297		636.297		636.297
CAPITAL COST \$/KWE		20.115		15.801		14.170		16.707		16.345		20.115		20.115		20.115
COE CAPITAL		23.311		23.311		23.311		23.311		23.311		23.311		23.311		23.311
COE FUEL		3.559		11.442		8.701		16.467		16.467		17.837		17.380		17.167
COE OP & MAIN		46.985		50.554		46.182		56.485		56.123		60.756		60.118		60.612
COST OF ELECTRIC		1.500		1.500		1.500		1.500		1.500		1.500		1.500		1.500
EST TIME OF CONST																

PARAMETRIC POINT	17	.000	18	.000	19	.000	20	.000	21	.000	22	.000	23	.000	24	.000
THERMODYNAMIC EFF		.000		.000		.000		.000		.000		.000		.000		.000
POWER PLANT EFF		.000		.000		.000		.000		.000		.000		.000		.000
OVERALL ENERGY EFF		.000		.000		.000		.000		.000		.000		.000		.000
CAPITAL COST MILLION \$.000		.000		.000		.000		.000		.000		.000		.000
CAPITAL COST \$/KWE		.000		.000		.000		.000		.000		.000		.000		.000
COE CAPITAL		.000		.000		.000		.000		.000		.000		.000		.000
COE FUEL		.000		.000		.000		.000		.000		.000		.000		.000
COE OP & MAIN		.000		.000		.000		.000		.000		.000		.000		.000
COST OF ELECTRIC		.000		.000		.000		.000		.000		.000		.000		.000
EST TIME OF CONST																

NATURAL RESOURCE REQUIREMENTS

Table A 13.4.30 ALKALINE FUEL CELL SYSTEM

	1	2	3	4	5	6	7	8
PARAMETRIC POINT	1.23480	1.24753	1.24753	1.02943	1.47633	1.23480	1.23480	1.23480
COAL, LB/KW-HR	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SORBANT OR SEED, LB/KW-HR	.000	.000	.000	.000	.000	.000	.000	.000
TOTAL WATER, GAL/KW-HR	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
COOLING WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
GASIFIER PROCESS H2O	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
WASTE HANDLING SLURRY	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SCRUBBER WASTE WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
NOX SUPPRESSION	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
TOTAL LAND ACRES/100MWE	6.85	5.20	4.15	3.92	13.36	6.85	6.85	6.85
MAIN PLANT	6.85	5.20	4.15	3.92	13.36	6.85	6.85	6.85
DISPOSAL LAND	.00	.00	.00	.00	.00	.00	.00	.00
LAND FOR ACCESS RR	.00	.00	.00	.00	.00	.00	.00	.00

	9	10	11	12	13	14	15	16
PARAMETRIC POINT	1.23480	1.23480	1.23480	1.23480	1.23480	1.23480	1.23480	1.22718
COAL, LB/KW-HR	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SORBANT OR SEED, LB/KW-HR	.000	.000	.000	.000	.000	.000	.000	.000
TOTAL WATER, GAL/KW-HR	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
COOLING WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
GASIFIER PROCESS H2O	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
WASTE HANDLING SLURRY	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SCRUBBER WASTE WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
NOX SUPPRESSION	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
TOTAL LAND ACRES/100MWE	6.85	6.85	6.85	6.85	6.85	6.85	6.85	6.85
MAIN PLANT	6.85	6.85	6.85	6.85	6.85	6.85	6.85	6.85
DISPOSAL LAND	.00	.00	.00	.00	.00	.00	.00	.00
LAND FOR ACCESS RR	.00	.00	.00	.00	.00	.00	.00	.00

	17	18	19	20	21	22	23	24
PARAMETRIC POINT	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
COAL, LB/KW-HR	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SORBANT OR SEED, LB/KW-HR	.000	.000	.000	.000	.000	.000	.000	.000
TOTAL WATER, GAL/KW-HR	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
COOLING WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
GASIFIER PROCESS H2O	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
WASTE HANDLING SLURRY	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SCRUBBER WASTE WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
NOX SUPPRESSION	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
TOTAL LAND ACRES/100MWE	.00	.00	.00	.00	.00	.00	.00	.00
MAIN PLANT	.00	.00	.00	.00	.00	.00	.00	.00
DISPOSAL LAND	.00	.00	.00	.00	.00	.00	.00	.00
LAND FOR ACCESS RR	.00	.00	.00	.00	.00	.00	.00	.00

Table A 13.4.31

MOLTEN CARBONATE FUEL CELL SYSTEM SUMMARY PLANT RESULTS

PARAMETRIC POINT	1	2	3	4	5	6	7	9
THERMODYNAMIC EFF	.000	.000	.000	.000	.000	.000	.000	.000
POWER PLANT EFF	.488	.480	.480	.544	.488	.488	.488	.488
OVERALL ENERGY EFF	.329	.323	.323	.457	.329	.329	.329	.329
CAP COST MILLION \$	11.516	44.954	113.868	559.901	19.871	11.516	11.516	11.516
CAPITAL COST \$/KWE	514.413	504.804	511.350	482.370	887.615	514.413	514.413	514.413
COE CAPITAL	16.262	15.958	16.165	15.249	28.059	16.262	16.262	16.262
COE FUEL	18.169	18.469	18.469	12.550	18.169	18.169	18.169	18.169
COE OP & MAIN	23.378	23.512	23.610	16.053	20.801	8.638	5.690	3.501
COST OF ELECTRIC	57.809	57.939	58.244	43.852	67.028	43.068	40.120	37.931
EST TIME OF CONST	1.500	2.000	2.500	5.000	1.500	1.500	1.500	1.500

PARAMETRIC POINT	9	10	11	12	13	14	15	16
THERMODYNAMIC EFF	.000	.000	.000	.000	.000	.000	.000	.000
POWER PLANT EFF	.488	.488	.488	.488	.488	.364	.448	.488
OVERALL ENERGY EFF	.329	.329	.329	.329	.329	.306	.312	.329
CAP COST MILLION \$	11.860	14.624	10.673	11.486	11.450	11.876	11.778	11.423
CAPITAL COST \$/KWE	529.773	653.233	476.727	513.067	511.447	540.085	528.435	510.244
COE CAPITAL	16.747	20.650	15.070	16.219	16.168	17.073	16.705	16.130
COE FUEL	18.169	18.163	18.169	18.169	18.169	18.750	20.588	18.169
COE OP & MAIN	3.611	30.748	18.956	23.289	23.110	23.889	23.528	23.065
COST OF ELECTRIC	38.527	69.567	52.195	57.677	57.447	59.712	60.821	57.364
EST TIME OF CONST	1.500	1.500	1.500	1.500	1.500	1.500	1.500	1.500

PARAMETRIC POINT	17	18	19	20	21	22	23	24
THERMODYNAMIC EFF	.000	.000	.000	.000	.000	.000	.000	.000
POWER PLANT EFF	.480	.000	.000	.000	.000	.000	.000	.000
OVERALL ENERGY EFF	.323	.000	.000	.000	.000	.000	.000	.000
CAP COST MILLION \$	149.979	.000	.000	.000	.000	.000	.000	.000
CAPITAL COST \$/KWE	673.514	.000	.000	.000	.000	.000	.000	.000
COE CAPITAL	21.291	.000	.000	.000	.000	.000	.000	.000
COE FUEL	18.465	.000	.000	.000	.000	.000	.000	.000
COE OP & MAIN	18.933	.000	.000	.000	.000	.000	.000	.000
COST OF ELECTRIC	58.694	.000	.000	.000	.000	.000	.000	.000
EST TIME OF CONST	2.500	.000	.000	.000	.000	.000	.000	.000

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Table A 13.4.32
MOLYBEN CARBONATE FUEL CELL SYSTEM
SUMMARY PLANT RESULTS

PARAMETRIC POINT	1	2	3	4	5	6	7	8	9
TOTAL CAPITAL COST	11.52	44.95	113.37	565.33	15.87	11.52	11.52	11.52	11.52
FUEL PROCESSING EQUIPMENT	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
FUEL CELL SYSTEM	4.75	11.52	47.50	171.50	3.10	4.75	4.75	4.75	4.75
STEAM TURBINE GENERATOR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
OXYGEN PLANT	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HEAT REC. STEAM GENERATOR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
RECUOPERATOR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
POWER CONDITIONING	1.33	7.00	17.50	59.83	1.33	1.33	1.33	1.33	1.33
ICI MAJOR COMPONENT COST	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TOT MAJOR COMPONENT COST	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SALAMAC OF PLANT COST	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TOTAL DIRECT COST	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
INDIF. LABOR COSTS	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
PROF. & OWNER COST	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CONSTRUCTION COST	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ESCALATION COST	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TOTAL CAPITAL COST	11.52	44.95	113.37	565.33	15.87	11.52	11.52	11.52	11.52
COST OF ELEC-FUEL	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
COST OF ELEC-OR&MAIN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TOTAL COST OF ELEC	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
COE D. 0.5 CAP. FACTOR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
COE 1.0 CAP. COST	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
COE 1.2 CAP. COST	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
COE (CONVENTIONAL)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
COE (ESCALATION)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00

Table A 13.4.32 Continued

MOLTEN CARBONATE FUEL CELL SYSTEM

SUMMARY PLANT RESULTS

PARAMETRIC POINT	17	18	19	20	21	22	23	24
TOTAL CAPITAL COST	149.98	.00	.00	.00	.00	.00	.00	.00
FUEL PROCESSING EQUIPMENT	.290	.000	.000	.000	.000	.000	.000	.000
FUEL CELL SYSTEM	31.700	.000	.000	.000	.000	.000	.000	.000
STEAM TURBINE GENERATOR	.000	.000	.000	.000	.000	.000	.000	.000
OXYGEN PLANT	27.660	.000	.000	.000	.000	.000	.000	.000
HEAT REC STEAM GENERATOR	2.950	.000	.000	.000	.000	.000	.000	.000
RECUPERATOR	1.344	.000	.000	.000	.000	.000	.000	.000
POWER CONDITIONING	17.500	.000	.000	.000	.000	.000	.000	.000
TOT MAJOR COMPONENT COST	81.444	.000	.000	.000	.000	.000	.000	.000
TOT MAJOR COMPONENT COST	365.743	.000	.000	.000	.000	.000	.000	.000
BALANCE OF PLANT COST	27.144	.000	.000	.000	.000	.000	.000	.000
SITE LABOR	70.069	.000	.000	.000	.000	.000	.000	.000
TOTAL DIRECT COST	462.957	.000	.000	.000	.000	.000	.000	.000
INDIRECT COSTS	35.735	.000	.000	.000	.000	.000	.000	.000
PROF & OWNER COSTS	37.037	.000	.000	.000	.000	.000	.000	.000
CONTINGENCY COST	25.483	.000	.000	.000	.000	.000	.000	.000
ESCALATION COST	54.554	.000	.000	.000	.000	.000	.000	.000
INT DURING CONSTRUCTION	57.769	.000	.000	.000	.000	.000	.000	.000
TOTAL CAPITALIZATION	673.514	.000	.000	.000	.000	.000	.000	.000
COST OF ELEC-CAPITAL	21.291	.000	.000	.000	.000	.000	.000	.000
COST OF ELEC-FUEL	18.469	.000	.000	.000	.000	.000	.000	.000
COST OF ELEC-OP&MAINT	18.933	.000	.000	.000	.000	.000	.000	.000
TOTAL COST OF ELEC	58.694	.000	.000	.000	.000	.000	.000	.000
COE 0.5 CAP. FACTOR	66.176	.000	.000	.000	.000	.000	.000	.000
COE 0.8 CAP. FACTOR	54.018	.000	.000	.000	.000	.000	.000	.000
COE 1.2XCAP. COST	62.952	.000	.000	.000	.000	.000	.000	.000
COE 1.2XFUEL COST	57.388	.000	.000	.000	.000	.000	.000	.000
COE (CONTINGENCY=0)	57.728	.000	.000	.000	.000	.000	.000	.000
COE (ESCALATION=0)	60.478	.000	.000	.000	.000	.000	.000	.000

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Table A 13.4.33

MOLTEN CARBONATE FUEL CELL SYSTEM

NATURAL RESOURCE REQUIREMENTS

	1	2	3	4	5	6	7	8
PARAMETRIC POINT	96240	97832	97832	69716	96240	96240	96240	96240
COAL, LB/KW-HR	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SORBANT OR SEED, LB/KW-HR	.000	.000	.000	.197	.000	.000	.000	.000
TOTAL WATER, GAL/KW-HR	.000	.000	.000	.0000	.0000	.0000	.0000	.0000
COOLING WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
GASIFIER PROCESS H2O	.00000	.00000	.00000	.19704	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
WASTE HANDLING SLURRY	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SCRUBBER WASTE WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
NOX SUPPRESSION	.00000	.00000	.00000	5.77	11.17	.00000	.00000	.00000
TOTAL LAND ACRES/100MWE	6.70	5.05	4.04	4.23	11.17	6.70	6.70	6.70
MAIN PLANT	6.70	5.05	4.04	4.23	11.17	6.70	6.70	6.70
DISPOSAL LAND	.00	.00	.00	.00	.00	.00	.00	.00
LAND FOR ACCESS RR	.00	.00	.00	1.54	.00	.00	.00	.00

	9	10	11	12	13	14	15	16
PARAMETRIC POINT	96240	96240	96240	96240	96240	1.03409	1.01409	96240
COAL, LB/KW-HR	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SORBANT OR SEED, LB/KW-HR	.000	.000	.000	.000	.000	.000	.000	.000
TOTAL WATER, GAL/KW-HR	.000	.000	.000	.000	.000	.000	.000	.000
COOLING WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
GASIFIER PROCESS H2O	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
WASTE HANDLING SLURRY	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SCRUBBER WASTE WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
NOX SUPPRESSION	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
TOTAL LAND ACRES/100MWE	6.70	6.70	6.70	6.70	6.70	6.82	11.22	6.70
MAIN PLANT	6.70	6.70	6.70	6.70	6.70	6.82	11.22	6.70
DISPOSAL LAND	.00	.00	.00	.00	.00	.00	.00	.00
LAND FOR ACCESS RR	.00	.00	.00	.00	.00	.00	.00	.00

	17	18	19	20	21	22	23	24
PARAMETRIC POINT	97832	97832	97832	97832	97832	97832	97832	97832
COAL, LB/KW-HR	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SORBANT OR SEED, LB/KW-HR	.000	.000	.000	.000	.000	.000	.000	.000
TOTAL WATER, GAL/KW-HR	.000	.000	.000	.000	.000	.000	.000	.000
COOLING WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
GASIFIER PROCESS H2O	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
WASTE HANDLING SLURRY	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SCRUBBER WASTE WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
NOX SUPPRESSION	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
TOTAL LAND ACRES/100MWE	7.19	7.19	7.19	7.19	7.19	7.19	7.19	7.19
MAIN PLANT	7.19	7.19	7.19	7.19	7.19	7.19	7.19	7.19
DISPOSAL LAND	.00	.00	.00	.00	.00	.00	.00	.00
LAND FOR ACCESS RR	.00	.00	.00	.00	.00	.00	.00	.00

Table A 13.4.34

SOLID ELECTROLYTE FUEL CELL

SUMMARY PLANT RESULTS

PARAMETRIC POINT	1	2	3	4	5	6	7	8
THERMODYNAMIC EFF	.000	.000	.000	.000	.000	.000	.000	.000
POWER PLANT EFF	.697	.409	.409	.602	.397	.416	.416	.416
OVERALL ENERGY EFF	.469	.343	.343	.506	.333	.350	.350	.350
CAP COST MILLION \$	9.670	105.249	105.249	579.054	17.699	10.666	10.666	10.666
CAPITAL COST \$/KWE	424.390	481.540	487.595	466.397	972.033	480.714	480.714	480.714
COE CAPITAL	13.416	15.223	15.414	14.744	30.728	15.196	15.196	15.196
COE FUEL	12.726	16.704	16.704	11.340	17.206	16.402	16.402	16.402
COE OP & MAIN	16.002	20.765	21.123	14.159	24.222	7.881	5.312	3.419
COST OF ELECTRIC	42.143	52.691	53.240	40.243	72.156	39.479	36.910	35.017
EST TIME OF CONST	1.500	2.000	2.500	5.000	1.500	1.500	1.500	1.500

PARAMETRIC POINT	9	10	11	12	13	14	15	16
THERMODYNAMIC EFF	.000	.000	.000	.000	.000	.000	.000	.000
POWER PLANT EFF	.416	.371	.320	.423	.416	.403	.416	.347
OVERALL ENERGY EFF	.350	.312	.269	.355	.350	.339	.350	.292
CAP COST MILLION \$	11.013	9.633	9.276	10.511	10.666	10.722	9.035	11.889
CAPITAL COST \$/KWE	496.325	434.166	418.057	473.716	480.714	483.276	407.181	535.834
COE CAPITAL	15.690	13.725	13.216	14.975	15.196	15.276	12.872	16.939
COE FUEL	16.402	18.392	21.307	16.141	16.402	16.945	16.402	19.658
COE OP & MAIN	3.530	16.176	14.460	20.176	20.635	21.101	20.635	24.304
COST OF ELECTRIC	35.622	48.292	48.982	51.292	52.233	53.322	49.909	60.901
EST TIME OF CONST	1.500	1.500	1.500	1.500	1.500	1.500	1.500	1.500

PARAMETRIC POINT	17	18	19	20	21	22	23	24
THERMODYNAMIC EFF	.000	.000	.000	.000	.000	.000	.000	.000
POWER PLANT EFF	.485	.530	.457	.567	.000	.000	.000	.000
OVERALL ENERGY EFF	.408	.537	.478	.382	.000	.000	.000	.000
CAP COST MILLION \$	9.799	205.457	893.285	7.668	.000	.000	.000	.000
CAPITAL COST \$/KWE	441.604	948.307	858.698	339.517	.000	.000	.000	.000
COE CAPITAL	13.960	29.978	27.145	16.733	.000	.000	.000	.000
COE FUEL	14.070	2.473	6.346	15.652	.000	.000	.000	.000
COE OP & MAIN	17.899	12.248	18.398	10.911	.000	.000	.000	.000
COST OF ELECTRIC	45.930	47.699	51.890	37.297	.000	.000	.000	.000
EST TIME OF CONST	1.500	3.000	5.500	1.500	.000	.000	.000	.000

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Table A 13.4.35
SOLID ELECTROLYTE FUEL CELL
SUMMARY PLANT RESULTS

Table A 13.4.35
SOLID ELECTROLYTE FUEL CELL
SUMMARY PLANT RESULTS

PARAMETRIC POINT	1	2	3	4	5	6	7	8
TOTAL CAPITAL COST	2.67	41.97	156.25	539.05	17.70	10.67	10.57	10.67
FUEL PROCESSING COST	0.07	2.68	40.00	142.412	3.400	4.000	4.000	4.000
FUEL CELL SYSTEM	0.00	0.00	0.00	11.518	3.900	0.000	0.000	0.000
STEAM TURBINE GENERATOR	0.00	0.00	0.00	0.000	0.000	0.000	0.000	0.000
OXYGEN PLANT	0.00	0.00	0.00	0.000	0.000	0.000	0.000	0.000
HEAT REC STEAM GENERATOR	0.34	1.918	17.500	59.600	1.380	1.380	1.380	1.380
RECUPERATOR	0.34	1.918	17.500	59.600	1.380	1.380	1.380	1.380
POWER CONDITIONING	1.330	1.330	1.330	1.330	1.330	1.330	1.330	1.330
TOT MAJOR COMPONENT COST	4.325	15.556	54.808	240.954	3.984	6.110	6.110	6.110
BALANCE OF PLANT COST	18.146	265.650	237.536	215.774	493.786	275.768	275.768	275.768
SITE LABOR	18.146	18.146	18.146	18.146	18.146	18.146	18.146	18.146
INDIRECT COSTS	315.107	350.788	343.086	284.322	705.811	481.171	481.171	481.171
PROF & OWNER COST	26.208	19.062	16.483	17.008	58.413	24.582	24.582	24.582
CONINGENCY COST	14.178	17.539	18.495	21.146	71.733	16.253	16.253	16.253
ESCALATION COST	21.728	17.027	19.435	66.427	43.559	24.261	24.261	24.261
TOTAL CAPITAL COST	427.500	491.540	487.535	466.744	972.033	480.714	480.714	480.714
TOTAL COST OF ELECT-FUEL	12.720	16.704	15.704	14.740	36.728	15.196	15.196	15.196
TOTAL COST OF ELECT-OP	142.143	170.691	171.120	140.663	372.474	16.402	16.402	16.402
COE OF ELECT	46.403	50.566	50.210	40.663	82.474	39.479	39.479	39.479
COE OF CAP. FACTOR	33.477	49.625	50.132	37.239	55.707	32.448	32.448	32.448
COE OF XCAP. COST	44.682	56.032	56.591	43.151	73.597	39.949	39.949	39.949
COE OF FUEL COST	41.643	52.703	52.532	29.473	71.075	36.706	36.706	36.706
COE (ESCALATION=0)	42.806	53.703	54.635	42.532	73.075	37.230	37.230	37.230

PARAMETRIC POINT	9	10	11	12	13	14	15	15
TOTAL CAPITAL COST	11.01	9.63	9.28	10.51	10.67	10.72	10.72	10.72
FUEL PROCESSING COST	0.07	0.00	0.00	0.00	0.00	0.00	0.00	0.00
FUEL CELL SYSTEM	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
STEAM TURBINE GENERATOR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
OXYGEN PLANT	0.153	0.153	0.153	0.153	0.153	0.153	0.153	0.153
HEAT REC STEAM GENERATOR	0.570	0.570	0.570	0.570	0.570	0.570	0.570	0.570
RECUPERATOR	1.518	1.518	1.518	1.518	1.518	1.518	1.518	1.518
POWER CONDITIONING	1.518	1.518	1.518	1.518	1.518	1.518	1.518	1.518
TOT MAJOR COMPONENT COST	6.386	5.202	4.949	5.995	6.110	6.140	6.140	6.140
BALANCE OF PLANT COST	237.807	234.446	223.043	270.135	275.368	276.720	276.720	276.720
SITE LABOR	48.199	48.199	48.199	48.199	48.199	48.199	48.199	48.199
INDIRECT COSTS	73.692	73.692	73.692	73.692	73.692	73.692	73.692	73.692
PROF & OWNER COST	25.885	25.885	25.885	25.885	25.885	25.885	25.885	25.885
CONINGENCY COST	25.412	25.412	25.412	25.412	25.412	25.412	25.412	25.412
ESCALATION COST	49.632	49.632	49.632	49.632	49.632	49.632	49.632	49.632
TOTAL CAPITAL COST	496.325	496.325	496.325	496.325	496.325	496.325	496.325	496.325
TOTAL COST OF ELECT-FUEL	15.620	15.620	15.620	15.620	15.620	15.620	15.620	15.620
TOTAL COST OF ELECT-OP	16.307	16.307	16.307	16.307	16.307	16.307	16.307	16.307
COE OF ELECT	35.622	35.622	35.622	35.622	35.622	35.622	35.622	35.622
COE OF CAP. FACTOR	33.876	33.876	33.876	33.876	33.876	33.876	33.876	33.876
COE OF XCAP. COST	38.700	38.700	38.700	38.700	38.700	38.700	38.700	38.700
COE OF FUEL COST	38.009	38.009	38.009	38.009	38.009	38.009	38.009	38.009
COE (ESCALATION=0)	36.037	36.037	36.037	36.037	36.037	36.037	36.037	36.037

Table A 13.4.35 Continued

SOLID ELECTROLYTE FUEL CELL

SUMMARY PLANT RESULTS

PARAMETRIC POINT	17	18	19	20	21	22	23	24
TOTAL CAPITAL COST	9.80	205.46	893.28	7.67	.000	.000	.000	.000
FUEL PROCESSING EQUIPMENT	.067	52.360	75.372	.067	.000	.000	.000	.000
FUEL CELL SYSTEM	3.400	18.800	167.000	1.900	.000	.000	.000	.000
STEAM TURBINE GENERATOR	.000	.000	.000	.000	.000	.000	.000	.000
OXYGEN PLANT	.000	.000	.000	.000	.000	.000	.000	.000
HEAT REC STEAM GENERATOR	.105	.211	11.800	.084	.000	.000	.000	.000
RECUPERATOR	.500	4.200	28.000	.440	.000	.000	.000	.000
POWER CONDITIONING	1.380	17.500	59.000	1.380	.000	.000	.000	.000
R TOT MAJOR COMPONENT COST	5.452	93.071	341.174	3.871	.000	.000	.000	.000
E TOT MAJOR COMPONENT COST	245.713	429.578	327.964	171.386	.000	.000	.000	.000
S BALANCE OF PLANT COST	37.604	39.592	36.982	37.390	.000	.000	.000	.000
U SITE LABOR	47.163	137.537	89.083	42.484	.000	.000	.000	.000
L TOTAL DIRECT COST	330.480	606.707	454.030	251.260	.000	.000	.000	.000
T INDIRECT COSTS	24.053	70.144	45.432	21.667	.000	.000	.000	.000
P PROF & OWNER COSTS	26.438	48.537	36.322	20.101	.000	.000	.000	.000
B CONTINGENCY COST	14.872	35.402	38.593	11.307	.000	.000	.000	.000
R ESCALATION COST	22.610	39.806	130.336	17.383	.000	.000	.000	.000
F INT DURING CONSTRUCTION	23.151	96.711	153.384	17.799	.000	.000	.000	.000
A TOTAL CAPITALIZATION	441.604	948.307	858.698	339.517	.000	.000	.000	.000
K COST OF ELEC-CAPITAL	13.960	29.978	27.145	10.733	.000	.000	.000	.000
D COST OF ELEC-FUEL	14.070	5.473	6.346	15.652	.000	.000	.000	.000
O COST OF ELEC-OP&MAINT	17.899	12.248	18.398	10.911	.000	.000	.000	.000
W TOTAL COST OF ELEC	45.930	47.699	51.890	37.297	.000	.000	.000	.000
N COE 0.5 CAP. FACTOR	50.403	56.994	60.337	40.785	.000	.000	.000	.000
COE 0.8 CAP. FACTOR	43.124	41.890	45.510	35.116	.000	.000	.000	.000
COE 1.2XCAP. COST	48.722	53.694	57.319	39.443	.000	.000	.000	.000
COE 1.2XFUEL COST	48.744	48.793	53.159	40.427	.000	.000	.000	.000
COE (CONTINGENCY=0)	45.405	46.266	50.066	36.898	.000	.000	.000	.000
COE (ESCALATION=0)	46.620	50.738	57.154	37.827	.000	.000	.000	.000

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Table A 13.4.36

SOLID ELECTROLYTE FUEL CELL

NATURAL RESOURCE REQUIREMENTS

	1	2	3	4	5	6	7	8
PARAMETRIC POINT	67408	92123	92123	62543	94895	90460	90460	90460
COAL, LB/KW-HR	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SORBANT OR SEED, LB/KW-HR	.000	.000	.000	.186	.000	.000	.000	.000
TOTAL WATER, GAL/KW-HR	.000	.000	.000	.000	.000	.000	.000	.000
COOLING WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
GASIFIER PROCESS H2O	.00000	.00000	.00000	.18547	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
WASTE HANDLING SLURRY	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SCRUBBER WASTE WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
NOX SUPPRESSION	6.58	2.58	2.07	4.17	8.24	3.38	3.38	3.38
TOTAL LAND ACRES/100MWE	6.58	2.58	2.07	2.60	8.24	3.38	3.38	3.38
MAIN PLANT	.00	.00	.00	.00	.00	.00	.00	.00
DISPOSAL LAND	.00	.00	.00	1.57	.00	.00	.00	.00
LAND FOR ACCESS RR	.00	.00	.00	.00	.00	.00	.00	.00

	9	10	11	12	13	14	15	16
PARAMETRIC POINT	90460	1.01435	1.17510	.89019	90460	93453	90460	90460
COAL, LB/KW-HR	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SORBANT OR SEED, LB/KW-HR	.000	.000	.000	.000	.000	.000	.000	.000
TOTAL WATER, GAL/KW-HR	.000	.000	.000	.000	.000	.000	.000	.000
COOLING WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
GASIFIER PROCESS H2O	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
WASTE HANDLING SLURRY	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SCRUBBER WASTE WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
NOX SUPPRESSION	3.38	6.76	6.76	3.38	3.38	3.38	3.38	3.38
TOTAL LAND ACRES/100MWE	3.38	6.76	6.76	3.38	3.38	3.38	3.38	3.38
MAIN PLANT	.00	.00	.00	.00	.00	.00	.00	.00
DISPOSAL LAND	.00	.00	.00	.00	.00	.00	.00	.00
LAND FOR ACCESS RR	.00	.00	.00	.00	.00	.00	.00	.00

	17	18	19	20	21	22	23	24
PARAMETRIC POINT	77601	59686	1.03604	82910	.00000	.00000	.00000	.00000
COAL, LB/KW-HR	.00000	31580	0.9882	.00000	.00000	.00000	.00000	.00000
SORBANT OR SEED, LB/KW-HR	.000	.101	.268	.000	.000	.000	.000	.000
TOTAL WATER, GAL/KW-HR	.000	.000	.000	.000	.000	.000	.000	.000
COOLING WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
GASIFIER PROCESS H2O	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
WASTE HANDLING SLURRY	.00000	.0654	.0338	.00000	.00000	.00000	.00000	.00000
SCRUBBER WASTE WATER	.00000	.03581	.03937	.00000	.00000	.00000	.00000	.00000
NOX SUPPRESSION	3.38	99.87	50.72	6.64	.00000	.00000	.00000	.00000
TOTAL LAND ACRES/100MWE	3.38	99.87	50.72	6.64	.00000	.00000	.00000	.00000
MAIN PLANT	.00	19.39	11.82	6.64	.00	.00	.00	.00
DISPOSAL LAND	.00	80.48	37.15	.00	.00	.00	.00	.00
LAND FOR ACCESS RR	.00	.00	1.75	.00	.00	.00	.00	.00