A scenic photograph of a waterfall cascading over rocks in a lush forest. A large, gnarled log lies horizontally across the foreground, partially submerged in the water. The background is filled with dense green foliage and trees. The overall atmosphere is serene and natural.

Hybrid Fuel Cell Technology Overview

May 2001

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Overview

For the purpose of this CD and unless otherwise stated, hybrid fuel cell systems are power generation systems in which a high temperature fuel cell is combined with another power generating technology. The resulting system exhibits a synergism in which the combination performs with an efficiency far greater than can be provided by either system alone. Hybrid fuel cell designs under development include fuel cell with gas turbine, fuel cell with reciprocating (piston) engine, and designs that combine different fuel cell technologies. Hybrid systems have been extensively analyzed and studied over the past five years by the Department of Energy (DOE), industry, and others. These efforts have revealed that this combination is capable of providing remarkably high efficiencies. This attribute, combined with an inherent low level of pollutant emission, suggests that hybrid systems are likely to serve as the next generation of advanced power generation systems.

This potential, while impressive, will require substantial development effort if it is to be realized. Balance-of-plant issues, the required infrastructure, and the resolution of technical hurdles that are as yet ill defined, will all have to be addressed. Given what we presently know about the various hybrids, a variety of technical issues are likely to emerge in the near-term:

- High temperature heat exchangers and the associated materials will need to be developed.
- New gas turbine designs will be required with pressure ratios, firing temperatures, and operating conditions adjusted to optimize the system cycle efficiency and reduce costs.
- Gas turbine recuperation and intercooling developments will be needed.
- Systems reliability and operability will have to be established and optimized for selected cycles. This will include startup, load matching, and system upset.

Developing hybrid power cycles will be a complicated and difficult process. The power cycles consist of a series of modules or building blocks. Many of these blocks are available or will be available in the near future. Many such building blocks or subsystems are under development in programs that are partnered by the National Energy Technology Laboratory (NETL), and U.S. industry. These key subsystems include the products of the advanced turbine systems programs and various fuel cells also developed under NETL sponsorship. By carefully integrating these developed subsystems, with the results of future programs aimed at solving the remaining technical challenges, hybrid power cycles will become a reality. These early hybrid systems will provide efficiencies of 60 percent or higher and will meet the required emissions and cost criteria necessary to be successful in the marketplace.

Vision Statement

NETL will champion and lead the global development and commercialization of hybrid systems for power generation applications. Hybrid power systems will be the highest efficiency power plants ever built with a target manufacturing level of 400 MW/year by 2015.

Hybrid Fuel Cell Systems

Examples of hybrid fuel cell power generation cycles are the combine high-temperature fuel cells and gas turbines, reciprocating engines, or another fuel cell. These represent the hybrid power plants of the future. The conceptual systems have the potential to achieve efficiencies greater than 70 percent and be commercially ready by year 2010 or sooner. The hybrid fuel cell/turbine (FC/T) power plant will combine a high-temperature, conventional molten carbonate fuel cell (MCFC) or a solid oxide fuel cell (SOFC) with a low-pressure-ratio gas turbine, air compressor, combustor, and in some cases, a metallic heat exchanger (1). The synergistic effects of the hybrid fuel cell/turbine technology will also provide the benefits of reduced greenhouse gas emissions. Nitrous (NO_x) emissions will be an order of magnitude below those of non-fuel cell power plants and carbon monoxide emissions will be less than 2 parts per million (ppm) (2). There will also be a substantial reduction in the amount of carbon dioxide produced as compared to conventional gas power plants.

Fuel cells are inherently attractive for distributed power applications, but their use is limited by the need to package them in a system balance of plant (BOP) that allows them to function effectively. All fuel cells, especially high-temperature fuel cells, require spent fuel utilization/waste heat recovery subsystems. Low-temperature fuel cells also require fuel reforming subsystems. A common approach for providing these BOP functions has been to integrate the fuel cell with another generating technology.

There can be many different cycle configurations for the hybrid fuel cell/turbine plant. In the direct mode, the fuel cell serves as the combustor for the gas turbine while the gas turbine is the balance-of-plant for the fuel cell, with some generation. In the indirect mode, the fuel cell uses the gas turbine exhaust as air supply while the gas turbine is the balance of plant. In indirect systems, high-temperature heat exchangers are used (3).

For certain distributed power applications, fuel-cell/fuel-cell hybrids may be an effective approach to the BOP problem. Only certain combinations of fuel cells promise to synergize into a simplified BOP. Generally these team a high temperature internally reforming fuel cell with a low temperature fuel cell having a complementary electrolyte (anion-conducting versus cation-conducting electrolytes). The SOFC/PEFC hybrid (solid oxide fuel cell/polymer electrolyte fuel cell) is of particular interest. Both fuel cells are the focus of current R&D programs, and the following synergisms are inherent:

- *SOFC as Reformer for the PEFC* - fuel utilization in the SOFC is limited to a range necessary only to reform natural gas, but not completely oxidize it.
- *PEFC as Chemical Bottomer for the SOFC* - reformed fuel utilization is completed in the PEFC, where more favorable thermodynamics apply.
- *Inherent Sequestration Compliance* - bulk air separation or acid gas removal is not required for isolation of CO₂, if sequestration is ultimately practiced in some way.

- *Coproduction and Load Following Options* - coproduction of hydrogen, possibly in conjunction with a storage system, allows the SOFC to be base loaded, while cycling is more effectively done with the PEFC.

The hybrid system is key to the Department of Energy's Vision 21 plants. The Vision 21 program has set power plant goals of achieving efficiencies greater than 75 percent (LHV) for natural gas. The higher efficiencies play a key role in reducing emissions, another target in Vision 21 plants. As a comparison, conventional coal-burning power plants are typically 35 percent efficient and natural gas fired plants are now 40 to 50 percent efficient. Figure 1 shows the estimated efficiency ranges of current and future power generation systems.

Figure 1. Estimated performance of Power Generation Systems

The combination of the fuel cell and turbine operates by using the rejected thermal energy and residual fuel from a fuel cell to drive the gas turbine. The fuel cell exhaust gases are mixed and burned, raising the turbine inlet temperature while replacing the conventional combustor of the gas turbine. Use of a recuperator, a metallic gas-to-gas heat exchanger, transfers heat from the gas turbine exhaust to the fuel and air used in the fuel cell. Figure 2 illustrates an example of a proposed fuel cell/turbine system. One item to note in Figure 2 is that the combustor shown prior to the turbine is for startup purposes only.

Figure 2. Diagram of a proposed Siemens-Westinghouse hybrid system.
(Taken from DOE Project Fact Sheet - Fuel Cell/ATS Hybrid Systems.)

The hybrid plants are projected to cost below comparably sized fuel cells, (4) and be capable of producing electricity at costs of 10 to 20 percent below today's conventional turbine plants (1). Operation of the plant is almost totally automatic. Therefore, it can be monitored and managed remotely with the possibility of controlling hundreds of the power plants from a single location (2).

Initial systems will be less than 20 MW, with typical system sizes of 1-10 MW. Future systems, in the megawatt size, can boost efficiency even further by combining two solid oxide fuel cell modules with more advanced gas turbines and introducing sophisticated cooling and heating procedures. As mentioned above, another possibility of a hybrid power plant is to combine a solid oxide fuel cell with a proton exchange membrane (PEM) fuel cell. The SOFC would produce both electric power and hydrogen. This hydrogen would then be utilized by the PEM to generate more electric power (2).

Countries around the world are developing interest in the high-efficiency hybrid cycles. A 320 kW hybrid (SOFC and gas turbine) plant will enter service in Germany in 2001, operated by a consortium under the leadership of RWE Energie AG. This will be followed in 2002 by the first 1 MW plant, which will be operated by Energie Baden-Württemberg AG (EnBW), Electricite de France (EDF), Gaz de France, and Austria's TIWAG (2).

References

1. “Developing Power Systems for the 21st Century - Fuel Cell/ATS Hybrid Systems,” U.S. Department of Energy, National Energy Technology Laboratory & Office of Industrial Technologies, Project facts for Advanced Clean/ Efficient Power Systems, PS031.1099.
2. U. Eberl, “Fuel Cells and Gas Turbines: A Marriage of Efficiency,” Research and Innovation, January 2000.
3. R. Gemmen, et al., “Technical Development Issues and Dynamic Modeling of Gas Turbine and Fuel Cell Hybrid Systems,” *Proceedings of the 1999 Review Conference on Fuel Cell Technology*, August 1999.
4. “Fuel Cells - Opening New Frontiers in Power Generation,” U.S. Department of Energy, Office of Fossil Energy, National Energy Technology Laboratory, November 1999.

Summary of Projects

The U.S. Department of Energy is currently funding several hybrid projects. FuelCell Energy Inc. with Allison Engine Company and Capstone Turbine Corp. has a contract to develop key system components for high-temperature hybrid operation, and design a 40 MW ultra high efficiency power plant. Honeywell International Inc. was selected for a 3.5-year program to develop and demonstrate a proof-of-concept SOFC hybrid system for distributed power generation applications. Rolls Royce was awarded a contract to develop a small gas turbine with a potential for multiple applications both in stand alone and combined modes.

DOE used a PRDA to fund five projects to explore the feasibility of linking technologies in a way that would generate electricity from natural gas at high efficiencies. The scope of these reports included investigating the engineering and economic feasibility of combining technologies - such as fuel cells and gas turbines - to break through the 70-percent efficiency threshold. Such "hybrid" plants are an important option in reducing greenhouse gas emissions. A 70-percent efficient natural gas-fueled plant would produce nearly 30 percent less carbon dioxide than modern gas-burning power plants.

In carrying out the feasibility studies, DOE asked companies to concentrate on power plant concepts that would be less than 20 megawatts in size and suitable for distributed power generation. The hybrid plants would also be capable of producing electricity at costs 10 to 20 percent below today's conventional plants, as well as reducing emissions of carbon dioxide, sulfur dioxide, and nitrogen oxides. The Department also specified that one of the components in the hybrid plant must be a high-temperature fuel cell. A brief summary of active and inactive hybrid projects is listed below:

1. In March of 1999, FuelCell Energy Inc. (formerly Energy Research Corporation), of Danbury, CT, with Allison Engine Company, Indianapolis, IN, and Capstone Turbine Corp., Woodland Hills, CA. was awarded a project under the Vision 21 program to create a fuel cell/turbine system that provides efficiencies and emissions targets that meet or exceed stringent Vision 21 goals. The 3-year program will include:
 - High fuel utilization fuel cell stack and system designs to increase the conversion efficiency of the fuel in the fuel cell.
 - Development of key system components including gas turbine, recuperators, and anode tail gas oxidizers for high-temperature operation.
 - Subscale tests of a commercial full-scale DFC[®] stack with a Microturbine in order to develop key design information needed for fuel cell/turbine integration.
 - Design of a 40 MW ultra high efficiency power plant including balance-of-plant design, plant layout, cost, and trade-off studies.

One important feature of DFC[®]/turbine system is that due to the uncoupled nature of fuel cell pressure and gas turbine pressure, the system works very efficiently over a wide range of air compression ratios (3 to 15). This important result implies that this hybrid system is suitable for a variety of applications and power plant sizes. Typically, the small-scale gas turbines (micro-turbines) use a low compression ratio, while the MW-size units are designed for high compression ratios (9 to 15). The DFC[®]/turbine system design is therefore suitable over a range of applications from sub-MW industrial to medium scale distributed generation to large distributed generation or central station plants. Further studies of the 20 MW hybrid systems incorporating existing single stage turbomachinery with a pressure ratio of 9 and 36 lb/sec of turbine exhaust flow have resulted in very high efficiencies ~ 65 percent LHV. While these systems do not meet the 75 percent LHV target efficiency of the Vision 21 long-term objectives, they may offer opportunities to field relatively efficient systems sooner than the 2010 time frame for Vision 21 with busbar cost-of-electric that are much lower than the 75 percent efficiency systems targeted for Vision 21.

2. In February 2001 Honeywell International Inc. was selected for award of a novel hybrid process based on a planar SOFC. Honeywell proposed a 3.5-year program to develop and demonstrate a proof-of-concept SOFC hybrid system for distributed power generation applications. The proposed system concept is based on Honeywell planar SOFC and turbogenerator technologies. The integration of the two technologies provides a hybrid system with many attractive features including low cost, small size, low emission, and high efficiency. The proposed program consists of three phases. Phase I will focus on system concept design and analysis and resolution of technical barriers concerning high fuel utilization and pressurized operation in SOFCs and high-temperature heat exchangers. Phase II will concentrate on defining and specifying the system, subsystems, components and performing a cost study. Phase III will involve developing, assembling, and testing a small SOFC hybrid system in the laboratory to demonstrate concept feasibility. The overall result of this proposed program will form the basis for further development of the technology toward commercialization.
3. Rolls Royce was awarded a contract to develop a small gas turbine with a potential for multiple applications both in stand alone and combined modes. The project will be carried to preliminary concept formulation stage. The contract has three tasks:
 - Task 1.0 -- Design of Core Technology Components and Specifications for Conceptual Turbine Unit: Configure a high efficiency, low cost, and very durable turbine. Design turbomachinery components to conduct preliminary structural dynamics, aerothermal evaluation, and approximate cost estimations.
 - Task 2.0 -- Design Validation of State-Of-The-Art Advancements: Validation will include identification of proposed configurations for specific markets.
 - Task 3.0 - Market Potential: Conduct a study analyzing the potential product market and market opportunities for the product(s).

4. World's first hybrid: Siemens-Westinghouse Power Corporation of Pittsburgh, PA developed and fabricated the first advanced power plant to combine a solid oxide fuel cell and a gas turbine. The microturbine generator was manufactured by Northern Research and Engineering Corporation of Woburn, MA. The factory acceptance test was completed in April 2000. Southern California Edison will operate the new hybrid plant at The National Fuel Cell Research Center at the University of California-Irvine. A year of testing in a commercial setting will be performed at this site. The system cycle is expected to generate electric power at 55 percent efficiency.

The pressurized system will generate 220 kilowatts of power and be operated at 3 atm of pressure. The fuel cell is made up of 1152 individual tubular ceramic cells and generates about 200 kilowatts of electricity. The microturbine generator will produce an additional 20 kilowatts of electricity at full power. There will be no sulfur dioxide pollutants released into the air. Nitrogen oxide emissions are likely to be less than 1 ppm.

A 320-kilowatt hybrid system is also in the planning stages. An initial commercial offering of a 1 MW fuel cell-microturbine power plant in late 2002 will be the end result of this Department of Energy/Siemens-Westinghouse partnership program.

For larger units, NETL has a cooperative agreement with Siemens-Westinghouse Power Corporation (SWPC) to develop a 60 percent efficient (LHV) fuel cell-gas turbine Hybrid power system. Under this agreement, SWPC has determined that, for a 1 MW plant, suitable gas turbines in the 220-kW size range will need to be developed for integration with the SOFC. These large systems are expected to be commercialized during the period 2003 to 2005.

PRDA Contracts

1. Siemens Westinghouse Power Corporation and Rolls-Royce Allison, “Pressurized Solid Oxide Fuel Cell/Gas Turbine Power System,” Contract Number DE-AC26-98FT40355. This study designed a near 20 MWe pressurized solid oxide fuel cell/gas turbine power system capable of 67% efficiency. This efficiency is ten points greater than that achievable with the best available large-plant conventional power generation technology, and twenty points above the efficiency achieved by a conventional 20 MWe-class gas turbine combined cycle power system. The specific power system concept was developed integrating HP and LP SOFC generators with an intercooled SOFC-reheated gas turbine. The system is planned to operate at a pressure of 7 atm and achieve 67.3% efficiency (net/LHV). NO_x levels are projected to be less than 1 ppm. The fuel cell will generate 16 MW of power and the gas turbine generate 4 MW of power.
2. Siemens Westinghouse Power Corporation and Caterpillar Inc., “A High Efficiency PSOFC/ATS-Gas Turbine Power System,” Contract Number DE-AC26-98FT40455. This system combines two 4.5 MWe Solar MercuryTM 50 gas turbine to generate 9 MW of power with the pressurized solid oxide fuel cell operated at 9 atm pressure to generate 11 MW of power. The system achieves an estimated power output of 20 MWe at an efficiency of 59.9% (net/LHV). This efficiency is twelve points higher than that achieved by the conventional gas turbine combined cycle power systems.
3. McDermott Technology, Inc. and Northern Research and Engineering Corporation (NREC), “Fuel Cell/Micro-Turbine Combined Cycle,” Contract Number DE-AC26-98FT40454. McDermott Technology, Inc., of Alliance, OH, developed a conceptual design that demonstrated the use of a micro-turbine, a non-pressurized fuel cell system and recuperator to achieve a cycle over 70% efficiency in a small size. The flow and heat requirements of components in the micro-turbine and Solid Oxide Fuel Cell Company CpnTM (Co-planar, n-stack) fuel cell module have been matched, resulting in a highly integrated package. The fuel cell is to produce nearly 660 kW of power, while the micro-turbine is to produce 70 kW. The gas turbine engine under development by NREC.
4. M-C Power Corporation, “High Efficiency Fossil Power Plant (HEFPP) Conceptualization Program,” Contract Number DE-AC26-98FT40356-02. M-C Power Corporation, of Burr Ridge IL, teamed with Bechtel, IGT, and Allison Engine Company, to determine the technical and economic feasibility of a 20 MW natural gas fueled MCFC combined with a gas turbine. The power split between the fuel cell and turbine was 85 percent/15 percent. The results of the study indicated the process would be about 70% efficient with NO_x levels less than 1 ppm. The COE is projected to be comparable to present-day alternatives.
5. Energy Research Corporation (ERC now FuelCell Energy) and Allison Engine Company, “High Efficiency Fossil Power Plants (HEFPP) Conceptualization Program,” Contract Number DE-RA26-97FT34164. ERC and Allison teamed to evaluate a carbonate fuel cell combined with a gas turbine and a steam turbine generator. The system is scheduled to operate at ambient pressure. The net power of the hybrid system is estimated at 20.6 MW and NO_x levels were estimated to be less than 1 ppm. The process is estimated to be 65 percent efficiency with off-the-shelf turbomachinery and 72 percent efficiency with cycle specific machinery. The COE is predicted to be comparable to present day alternatives.

Research and Development

Key Technical and Market Hurdles

The key technical and market hurdles were extracted from a November 1999 hybrid workshop (1). The key issues were discussed in breakout sessions and technical challenges were identified. The breakout sessions looked at nine topics. Listed below are the main topics and technical challenges for each breakout topic.

1. Optimize/Customize Turbine Subsystem

- Surge margin
- Hot section life
- Don't know optimal control strategy
- Interfacing requirements

2. Market Design Analyses

- Interaction of system with one another and grid
- Accurate models of equipment and integration
- Sharing of market information amongst competitive entities
- Determining system operational features and corresponding value

3. Integration of Fuel Cells and Engines

- Interface of major components
- Control
- Reconfiguration of major components
- Flow matching
- System operational features
- Thermal management
- System balance
- Safety

4. System Component Design and Optimization

Engine

- Enhance performance
- Increase turbine inlet temperature
- Structural integrity
- Speed control
- Bearings (never seize)

Heat Exchanger

- Higher temperature materials
- Thermal cycling
- Durability
- Corrosion

Combustor

- 0-100 % turn-down capacity
- Stability
- Vitiated air
- Withstand higher inlet air temperatures
- Low pressure drop
- Low emissions
- High efficiency

Power Electronics

- Cooling for various conditions
- Paralleling and islanding
- Modularity
- Control
- Standardization

Fuel Cells

- High pressure capacity
- Increased power density
- Materials (e.g. for high pressure)
- Internal reformation
- Flow and thermal management
- Controls

Fuel Processor

- Thermal management
- High pressure operation
- Fuel variability and flexibility
- Contaminant/odorant tolerance
- Contaminant removal
- Eliminate need for steam

5. Analysis tools for combined system

- Lack of empirical data (proprietary data)
- Accurate determination of time constants
- Concurrent dynamic and materials/stress analyses
- Computing power

6. Integration and Optimization of Fuel Cell/Turbine Combined Cycles

- To define proper configuration
- Matching transient response of fuel cell and turbine
- Fuel cell and turbine power output integration
- Stored energy, fuel cell, and turbine management/control
- Transients over entire duty cycle

7. Specific Functionality (Steady State and Dynamic) and Specification of System Components

- Lack of adequate input data
- Establish specifications within realistic bounds
- Machine tool development to match specifications
- Fuel compressor to meet 8700 hour requirement

8. System Component Design & Development

- Power density of fuel cell
- Hot section RAMD
- Sulfur tolerance
- Inverter cost, efficiency, and robustness

9. Simplify/Optimize System Configuration

- Low power density of fuel cell
- Unknown steady state and/or dynamic response
- Lack of RAMD numbers
- Packaging

Key Research and Development Issues

While hybrid fuel cell/turbine systems offer excellent efficiencies, there are a number of control and safety issues that will have to be resolved before hybrid systems can become commercially successful. Load loss and load following issues are given below.

Load Loss (2)

A sudden loss of load is difficult to handle in a hybrid fuel cell/gas turbine system. When a load loss occurs, the fuel cell immediately stops reacting fuel, which can then accumulate in the system. Depending upon system design, combustible mixtures could accumulate. Fuel will continue to enter the cell after the load loss because the response time of the fuel control system (fractions of a second) is much slower than that of the reaction rate drop (milliseconds). In addition there will be fuel in the lines and the reformer. Air will also continue to be pumped through the system by the gas turbine because the large thermal load of the fuel cell will maintain adequate turbine inlet temperature for some considerable time (minutes). These relatively large time differences make load loss a major safety issue that must be addressed in hybrid systems.

A methodology considered for solving the safety issue of load loss takes a two-pronged approach involving a rapid shutdown of the gas turbine and the flooding of the fuel cell with inert gases. For manufacturers of non-hybrid fuel cell systems operating at near ambient pressures, the flooding shutdown approach has been the one of choice. Typically, in the small fuel cell systems operated to date, a bank of high-pressure nitrogen bottles is located close to the cell and in case of emergencies is used to flood and purge the system with nitrogen. In large installations this method will probably be neither economically feasible nor practical.

In the hybrid system, the gas turbine section can be stopped rapidly by using a blow-off valve that vents the high-pressure air leaving the compressor to atmosphere before it can enter the fuel cell. This approach minimizes the entry of oxygen into the fuel cell and also stops the gas turbine. At the same time that the blow-off valve is being operated, the gas flow to the reformer can be diverted to an inert gas generator. This generator, which consists of a catalytic combustor and a cool-down heat exchanger, employs a blower to mix air with the fuel before it enters the catalytic combustor. In the catalytic combustor, the fuel burns completely to carbon dioxide and water. The products of combustion are then cooled by a combination of direct injection of water and by an air or water cool-down heat exchanger. The cooled products would then be used to purge the fuel cell. However, the various delay times inherent in this approach create cause for concern. If the blow-off valve acts sufficiently quickly to cut off the airflow to the fuel cell, the gas mixture inside the system will be sufficiently fuel rich that it probably will not burn or explode. Also, if the inert gas generator can be brought online quickly, perhaps augmented by some bottled nitrogen, the hydrogen and carbon monoxide can be diluted and purged from the system before a dangerous accumulation of fuel occurs. With both of these conditions met, a flammable mixture is avoided, and the hybrid system can be safely shutdown after a load loss.

Load-Following (2)

During load drops, less fuel is reacted in the fuel cell while during load increases more fuel is reacted. The two effects tend to balance out during small, or short lived transients so that there is little or no residual unburned hydrogen or carbon monoxide retained in the system, the primary safety concern for load-following applications. Thus, short-term load excursions that are no greater than 10 to 15% of the base-load condition will probably be handled by the fuel cell alone. The inertia of the rotating machinery will probably keep the gas turbine at a near constant output during these short-lived transients.

Short term decreases in current load that are in excess of 15% will probably cause the power output and speed of the gas turbine section to drop. This occurrence will lengthen the time needed to recover to the base-load condition. Thus large excursions in load will probably be dominated by the response time of the gas turbine.

If the load excursions are not short transients but are of long duration, a standard fuel control system may be able to respond sufficiently enough to reduce the fuel flows to the cell. This safety measure will prevent any large accumulation of hydrogen and carbon monoxide inside the fuel cell.

Long-lived load drops in excess of the base-load level will cause the temperature of the fuel cell to drop until a new equilibrium point is reached. The fuel controller would measure the power output and speed of the gas turbine and the power output of the fuel cell and the control system would decrease or increase fuel flow rate when a demand signal is received. The total power produced would be compared to the demand level and if a match were made within preset gas turbine speed limits the fuel flow would be maintained at the matching level.

Turndown of the system, however, may not be handled well by the simple fuel flow reductions of a standard fuel control system. Power reduction is hampered by the efficiency increase of the fuel cell as it is unloaded. To turndown successfully the system airflow may have to be reduced in conjunction with reductions in fuel flow. This process would probably involve bleeding air with a controllable blow-off valve from the compressor exit and dumping it overboard, probably into the system exhaust. Alternatively air could be bypassed around the recuperator, which would reduce the inlet temperature at the fuel cell and lower fuel cell efficiency. This response in combination with reduced fuel flows may be sufficient to predictably reduce the fuel cell power output. Combinations of all three above actions may be required to effectively turndown to a desired power level within a reasonable time period.

References:

- (1) Proceedings: 1999 National Fuel Cell Research Center Hybrid Workshop, Irvine CA., November 19, 1999.
- (2) S. Samuelsen, J. Brouwer, and D. White, "Gas-Turbine Fuel Cell Hybrids," Power Gen99, New Orleans, September 1, 1999.

NETL and Office of Energy Efficiency and Renewable Energy Activities

SECA

The Solid-State Energy Conversion Alliance (SECA), was recently formed with the goal of producing a core solid-state fuel cell module that can be produced at a cost of no more than \$400 per kilowatt. To achieve this goal, SECA encompasses virtually all the major workers in fuel cells, including numerous government agencies, commercial developers, universities, and national laboratories. NETL's Strategic Center for Natural Gas (SCNG), in partnership with the Pacific Northwest National Laboratory (PNNL), will coordinate SECA's effort to transition from the current segmented fuel cell R&D (where different fuel cell concepts are being pursued for various applications) to a coordinated national research initiative focused on today's most promising fuel cell technology. Our approach is to support two complementary and integrated efforts. Core technologies will be developed by "horizontal teams" led by researchers at DOE national labs and other leading fuel cell developers. "Vertical teams" led by major fuel cell manufacturers will integrate these core technologies into commercial fuel cell systems targeted for different market segments.

The basic building block of the proposed future fuel cell will be a compact and light-weight, 5-kilowatt solid state module that can be accurately mass-produced. The modularity of the system will remove a major contributor to current high fuel cell costs, the current need to separately design and custom-build fuel cell stacks for each particular application. In addition, the proposed solid state design will leverage numerous recent advances, such as production of thin-film solid electrolyte materials and precise, automated manufacturing technologies developed largely in the semi-conductor industry. Reduced manufacturing costs, when combined with the traditional high efficiency and outstanding environmental performance of the fuel cell, will make the SECA module the prime option for a wide range of electric power needs. Uses include stationary applications ranging from residences (1 or 2 module - 5 to 10 kW systems) to multi-megawatt central power stations. It will also be a prime option for powering tomorrow's electric vehicles. Ship propulsion and other specialized military applications will also be supported.

DOE/Office of Energy Efficiency and Renewable Energy

The Office of Energy Efficiency and Renewable Energy (EERE) develops and deploys efficient and clean energy technologies that meet our nation's energy needs, enhance our environment, and strengthen our national competitiveness. EERE defines hybrid power systems as one that uses any power system with more than one type of generator--usually a conventional generator powered by a diesel or gas engine, and a renewable energy source such as a photovoltaic (PV), wind, or hydroelectric power system. There are thousands of these systems in use today. They range in size from a few tens of watts to tens of kilowatts. The smaller systems are mostly on remote residences where homeowners add a few PV modules to their existing genset to reduce the noise and inconvenience of having the generator running all the time. Convenience may be more important than cost for these homeowner-maintained systems that are installed near their homes.

Conclusion

Hybrid power systems based on combining high-temperature fuel cells with another power generating technology provide a tremendous opportunity to reduce pollutant emissions, (including the emission of carbon dioxide), costs, and fuel use associated with energy conversion to heat and power. Successful hybrid plants, with less environmental impact than current technologies, will provide an option to meet the growing needs of electrical power, domestically and internationally. Initially, hybrid power plants less than 20 MW in size, suitable for distributed power generation, will be developed. Such hybrid power plants will produce electricity at costs 10 to 20 percent below today's conventional plants, as well as reducing emissions of both air pollutants and greenhouse gases.

Hybrid Conferences and Links

Conferences:

1. First International Conference on Hybrid Power Systems jointly sponsored by: U.S. Department of Energy, Office of Fossil Energy; United Nations, Economic Commission for Europe; National Energy Technology Laboratory. May 21-25, 2001, The Sutton Place Hotel, Newport Beach, CA.
2. 2nd Annual Intl. Colloquium & Exhibit on Environmentally Preferred Advanced Energy Generation. May 21-23, 2001, U of California, Irvine, CA.

Links:

Hybrids and Distributive Generation

- Hybrid Vehicles - bimonthly. Published by Energy Futures, Inc. Boulder, CO 80306. Online subscription available at www.energy-futures.com/
- Distributive Generation - <http://www.dginfosource.com>

NETL

- NETL Hybrid Program - <http://www.netl.doe.gov/scng/index.html>
- NETL Vision 21 Program - <http://www.netl.doe.gov/products/power1/vision21frameset.htm>

Other DOE

- Energy Efficiency and Renewable Energy Network-
http://www.eren.doe.gov/EE/power_distributed_generation.html

Utility Associations and Institutes

- AGA (American Gas Association)
- APGA (American Public Gas Association)
- EPRI (Electric Power Research Institute)
- GTI (Gas Technology Institute) - IGT; GRI

Fuel Cell Associations

- NFCRC (National Fuel Cell Research Center)
- USFCC (US Fuel Cell Council)
- WFCC (World Fuel Cell Council)

Gas Turbine Associations

- GTA (Gas Technology Association)
- IGTI (International Gas Turbine Association)

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Fuelcell Energy, Inc. and Allison Engine Company — *High Efficiency Fossil Power Plants (HEFPP) Conceptualization Program (Volume II)*

M-C Power Corporation — *High Efficiency Fossil Power Plants (HEFPP) Conceptualization Program (Final Report)*

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Siemens Westinghouse/Caterpillar, Inc. — *A High Efficiency PSOFC/ATS-Gas Turbine Power System*

Brochures and Factsheets

1. Distributed Generation: <http://www.netl.doe.gov/publications/brochures/pdfs/distgen.pdf>
2. Development Status of Hybrids: http://www.netl.doe.gov/products/power/ats/ref_shelf/hybrid.pdf
3. Rocket Engine Design Adapted for Turbine Power: <http://www.netl.doe.gov/publications/factsheets/project/proj007.pdf>

***High Efficiency Fossil Power Plants (HEFPP)
Conceptualization Program***

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**Final Report
Volume I - Technical Report**

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EXECUTIVE SUMMARY

This is the Final Report covering Energy Research Corporation's High Efficiency Fossil Power Plants (HEFPP) Conceptualization Program. The project was a cost-shared effort supported by the US Department of Energy through the Federal Energy Technology Center, Morgantown WV, under PRDA DE-RA26-97FT34164.

Under this project, Energy Research Corporation (ERC) has developed a fuel cell / turbine system that integrates the highly efficient baseline atmospheric pressure direct fuel cell with an unfired Brayton cycle running on the fuel cell waste heat. System analyses have been done using performance projections based on the baseline commercial product assumptions, which are reviewed annually in DOE's due diligence studies. Based on these performance assumptions, an efficiency of about 71% has been projected, with 85% of the power coming from the fuel cell section and the balance from the turbine section. System calculations have also been done with more aggressive fuel cell performance assumptions which indicate that LHV efficiencies close to 80% are possible in the longer term.

ERC has been evaluating plant designs which combine the carbonate direct fuel cell (DFC) with combustion and/or steam turbine generation equipment since early studies were done under a CRADA with FETC. The baseline hybrid system developed in the earlier studies involved the transfer of waste heat into a fired Brayton cycle through a very high temperature (2000 °F) heat exchanger. Under the HEFPP program, an alternate system has been developed which incorporates a number of improvements:

- The heat exchange temperature has been reduced to less than 1600°F, which is in the range of commonly available high temperature heat exchange materials.
- The Brayton cycle is an unfired system, indirectly heated with fuel cell waste heat. This yields the highest efficiency, since all primary fuel consumption is done in the fuel cell - which is the more efficient portion of the system.
- Cathode gas and anode gas recycle blowers have been eliminated. All gas supply is accomplished with the normal natural gas pressure and the air compressor associated with the turbine equipment.
- High efficiency is obtained without the use of an additional steam bottoming cycle, eliminating the need for high pressure boilers (a concern with respect to unattended operation).

In addition to these improvements over our previous hybrid system, the system also retains these features of the DFC hybrid:

- Two key features of our baseline simple cycle DFC product are retained: Ambient pressure operation and internal reforming. This is the most efficient and reliable

- The system allows the turbine pressure ratio to be set independently of fuel cell pressure considerations.
- Although in normal hybrid operation the turbine will be unfired, a burner could be provided in the turbine to allow independent operation of the turbine, replacing the existing fuel cell startup burner. Conceptually, the turbine section could be producing power during the fuel cell startup.
- The turbine section can be used to load follow - utilizing stored kinetic energy - while the fuel cell is efficiently baseloaded at constant power. This is only possible because of the decoupled nature of the fuel cell and turbine sections of the plant.

Figure ES-1 shows a simplified system diagram for the hybrid system developed in the project. The system concept is as follows. Fuel and water are sent to the system Heat Recovery Steam Generation (HRSG) unit, where steam is produced and mixed with heated fuel for use as the DFC fuel gas feed. The DFC - which produces 17 MW - operates at 78% fuel utilization, and residual fuel from the anode exit is consumed in an anode exhaust oxidizer. Air is compressed in an intercooled compressor to about 16 atm, heated with system exhaust in the HRSG, heated further with exhaust from the anode exhaust oxidizer, and expanded in a turbine. The turbine produces about 3.4MW net power output. The expanded, low pressure air leaving the turbine is used

as the oxidant in the anode exhaust oxidizer. Flue gas leaving the oxidizer is first cooled by the turbine air, and then sent as the cathode feed gas to the fuel cells. The cathode exhaust gas is sent through the HRSG to provide the required pre-heat and water vaporization.

The following operating characteristics are projected for the hybrid system:

Fuel Cell:

DC Power: 17.62 MW
AC Power, Gross: 17.00 MW

Turbine

Compressor Power: 5.84 MW
Expander Output: 9.28 MW
Net Output: 3.44 MW

Parasitic 0.03 MW

Total Plant Output: 20.41 MW

Overall LHV Efficiency: 71.08%

In addition to development and optimization of the process design, the project also included the development of an optimization tool, consisting of an EXCEL^(TM) based spreadsheet model of the system. The basic system design of the selected hybrid plant was developed with ChemCAD^(TM) process modeling software, which was augmented with an ERC-developed fuel cell add-in model. This approach was used to select the optimum flowsheet. Once this was done, a model of the selected system was built in Microsoft EXCEL^(TM). This model is being submitted along with the final report. It allows the user to vary key system parameters to assess the impact on system performance, and provides a more in-depth understanding of the design.

Once a system design was selected from among the process options, work was done to specify the power plant equipment items. This was done using both the ChemCAD^(TM) model and the EXCEL^(TM) spreadsheets, and both tools were found to agree very closely. Once equipment specifications and sizes were established, a preliminary layout of the powerplant was developed. This is shown in Figure ES-2.

Figure ES-2
Preliminary Layout of 20MW Hybrid Powerplant

A cost estimate was developed for the powerplant, using assumptions specified by FETC in the September 24, 1998 "Reporting Requirements". An installed equipment cost of \$17,520,242, or \$858/kW is projected. Cost of energy calculations were done, also using economic assumptions specified in the FETC Reporting Requirements. As shown in Figure ES-3, the system was found to be cost competitive with conventional combined cycle equipment for cases where natural gas cost was \$2/MMBtu or greater.

The cost estimate was also evaluated in a Risk Assessment task, in which probabilistic curves were assigned to individual cost estimate line items, and the impact on projected system cost was assessed. Crystal Ball ^(TM) software was used for this analysis. This software is a Microsoft EXCEL ^(TM) add-in which runs over the spreadsheet-based cost estimate and uses a Monte-Carlo simulation to evaluate the overall probability of meeting or exceeding the projected cost. The simulation projected an 80.8% probability that the system cost would be equal to or lower than the overall estimated cost.

Figure ES-3: Cost of Energy Comparison with Gas Turbine Cycle

Figure ES-4: Risk Analysis Probability Distribution for Powerplant Cost

ERC was assisted in the program by Allison Engine Company, who provided valuable insight into the turbine design issues. Along with providing design and cost details to support the selected system design (which utilizes a specialized turbine with intercooled compressors and a reheat expansion cycle), Allison personnel also noted that the design concept could be used with more conventional turbomachinery. Near term plants could be configured with available machinery from their existing product lines and achieve LHV efficiencies in the mid 60 percent range.

In summary, the HEFPP project has identified an attractive DFC/Turbine hybrid design which offers the promise of very high efficiency in a relatively simple system. Additional work which was out of the scope of the study and which needs to be done on the design includes:

- Analysis of performance issues at partload
- Startup, shutdown, and control issues
- Evaluation of possible independent operation of turbine, e.g. during fuel cell heatup
- Evaluation of near-term options for proof-of-concept, field trials, and early market entry, including:
 - Smaller systems. Hybrids based on ERC's market entry sub-MW, 1.5 MW and 3 MW power plants could offer very high efficiencies at their respective size levels
 - Field test powerplants, consisting of hybrids with ERC's 200kW, 1MW, or 2MW near term field test systems
- Detailed design of system-specific components. An advantage of the selected system is that there are no long term developmental issues in the balance of plant (like the high temperature heat exchanger from the earlier system). Certain components are unique to the hybrid system, and will require additional design specification. Examples include the anode exhaust oxidizer (which operates at a higher temperature than the one in the baseline simple cycle system) and the multi-stage turbomachinery.

With these activities as a next logical step, a path forward is possible which can lead to the availability of very high efficiency power plants in the near future.

1.0 INTRODUCTION

1.1 Technology Background and Earlier Hybrid Work

This is the Final Report covering Energy Research Corporation's High Efficiency Fossil Power Plants (HEFPP) Conceptualization Program. The project was a cost-shared effort supported by the US Department of Energy through the Federal Energy Technology Center, Morgantown WV, under PRDA DE-RA26-97FT34164.

Energy Research Corporation (ERC) began working on the carbonate fuel cell in the mid-seventies, and is currently recognized as the technology leader. An advanced version of the carbonate fuel cell system has been developed at ERC in which the necessary reforming reactions occur within the fuel cell stacks, eliminating the need for a reforming subsystem in the power plant. This version of the carbonate technology - the Direct Fuel Cell (DFC) - is characterized by extreme simplicity, which contributes to the system's inherently low cost and high reliability. In an extension of this philosophy of system simplification, ERC's system operates at ambient pressure, and is configured with only one prime mover, a cold air blower. The system concept is shown in Figure 1.1.

Figure 1.1
Baseline Simple Cycle System Design

The process involves treating natural gas to remove impurities, after which it is mixed with steam and sent to the fuel cell stacks. The fuel/steam mixture is reformed in the stacks, providing the hydrogen which is consumed in the fuel cell anodes (at a hydrogen utilization of up to 80%). The anode reaction also produces CO_2 , which is required in the fuel cell cathodes. The cathode feed is produced by taking the anode exhaust, catalytically reacting any residual fuel with air, and sending the flue gas with excess air to the cathodes. Cathode exit gases are sent to a packaged heat recovery unit to supply heat for steam generation, fuel pre-heat, and final heating of the steam/fuel anode feed gas.

This DFC power plant system has several attractive features. The system operates at ambient pressure, which has been shown in studies conducted by ERC and Fluor Daniel, Inc. (FDI) to have a beneficial impact on cost of energy. Compared to operation at higher pressures, the atmospheric pressure design utilizes lower cost components and exhibits longer fuel cell stack life. Potential customers of these power plants have indicated that the system will be acceptable in a wider range of applications if it is not pressurized. Also, operation of the system at atmospheric pressure limits the pressure differentials which can be imposed on the fuel cell stacks in the event of a plant upset. Another aspect of atmospheric pressure operation is that it facilitates the use of internal reforming, which is inhibited at higher operating pressures.

The internal reforming feature of ERC's Direct Fuel Cell Power plant has many advantages. The obvious cost savings associated with the elimination of a large fired external reformer is one of the principle advantages. Internal reforming also enhances system efficiency, since the reforming reaction helps to remove stack waste heat, eliminating the need to increase the cathode gas flow (with a substantial increase in parasitic power) to remove heat. Heat removal via cathode gas alone usually also requires a cathode gas recycle system, which adds to system cost and complexity. An additional efficiency gain occurs because internal reforming can be done at low steam to carbon ratios (since consumption of reactant in the anodes drives the reforming equilibrium to completion), which improves the performance of the fuel cell anodes.

The system concept has been tested at the multi-megawatt level in ERC's 2MW Santa Clara Demonstration Project (SCDP), a cost-shared DOE Cooperative Agreement program which demonstrated the advantages of the system concept. The strengths of the atmospheric pressure, internal reforming, Simplified Design were demonstrated by the robust performance of the balance of plant (BOP). Over the 4000 grid-connected hours of operation, the BOP exhibited an availability of 98% (99% if operator procedural errors are discounted).

Given the attractive features of the atmospheric pressure, internally reformed, DFC concept, both in principle and as demonstrated at the 2MW level in Santa Clara, the use of the system as the basis for an ultra high efficiency power plant is a logical step. In addition to the baseline simple cycle design, this project had as another starting point a baseline hybrid system, which had been developed earlier by ERC under a ERC/FETC Crada which evaluated hybrid systems ranging from 20 to 200MW in size. The system design used in the 200MW system (scaled down to 20MW), was selected as a baseline for the current program, since it exhibited efficiency above 70%.

A flow diagram of this first generation hybrid power plant is shown in Figure 1.2. In this system, about 95% of the fuel was sent to the fuel cell stacks, where the fuel was reformed and consumed at 80% utilization. Five percent of the fuel was compressed and sent to the gas turbine combustor. The gas turbine compressor delivered air at 360 psia to a high temperature heat exchanger which heated the air to 982°C (1800°F). The heated air then flowed to the gas turbine combustor where it was heated further to 1094°C (2000°F) before flowing through the turbine. Turbine exhaust was split, with 50% of the stream flowing to the fuel cell cathodes and 50% flowing to the anode exhaust burner. Exhaust from the anode exhaust burner heated the compressed air in the high temperature heat exchanger before joining the stream to the fuel cell cathodes. Exhaust from the cathodes at 677°C (1250°F) went to a steam bottoming system. Exhaust from the cathodes at 677°C (1250°F) went to a steam bottoming system.

MO1537A

Figure 1.2
First Generation Hybrid Design System

The projected performance of this system at the 20 MW level is summarized in Table 1.1.

Table 1.1
Power Output Summary from First Generation 20 MW hybrid system

Direct Fuel Cell Power	15.7 MW
Gas Turbine Power	2.8 MW
Steam Bottoming Cycle	2.9 MW

Total Power Output, Gross DC	21.4 MW
Parasitic Power	0.8 MW

Net Power Output, Net AC	20.6 MW

This system has a high projected efficiency (72.5% LHV), preserves the basic features of the DFC design (atmospheric pressure, internal reforming), and provides a degree of independence between the fuel cell and turbine systems. However the system is substantially more complex than the baseline simple cycle system. Anode and cathode recycle streams have been added, and the system requires a steam turbine bottoming cycle to achieve >70% efficiency. In addition, it requires a heat exchanger which operates at 2000°F. Addressing these concerns was a major objective of the current HEFPP program.

1.2 HEFPP Project Description

The overall objective of the program was the development of a conceptual design of a cost effective, environmentally friendly, power plant which operates with an LHV fuel efficiency on natural gas of 70% or better. It was important that the high efficiency system be configured with technologies which will exhibit the high reliability and availability demanded by power applications, and which can be commercialized in the first decade of the next century. An additional objective of the program was the identification of the required path forward for development and commercialization of the system.

To provide a baseline for optimization efforts and cost analyses, FETC prepared a "Reporting Requirements" document which was reviewed by HEFPP contractors and finalized in a September 24, 1998 issue. The design and cost assumptions established in the Reporting Requirements document are summarized in Table 1.2

Table 1.2
Summary of FETC Reporting Requirements Assumptions

<ul style="list-style-type: none"> Systems should be sufficiently developed in time for demonstration in the year 2010 or sooner Application is Commercial distributed-power generation utility - grid connected. 	
Fuel:	Natural gas, Composition: CH ₄ , .96; N ₂ , .02; CO ₂ , .02; S, 4 ppmv. mercaptans
Fuel supply pressure	15 psig
Fuel supply temperature	59°F
Air composition:	Oxygen, .207; Nitrogen, .78; CO ₂ , .003; Water, .01 (60% relative humidity)
Ambient air temperature	59°F
Ambient air pressure	14.7 psia
Ambient Air Parameter Ranges	
Pressure	Sea level to 5000 ft.
Temperature	-20°F to 120°F
Relative Humidity	0 to 100%
Performance Requirements	
Design-Point Capacity	20 MW net AC, +/-2 MW
Design-Point Efficiency	> 70% (net AC/LHV)
Power System Dispatch Mode	Base loaded
Turndown Requirement	None
Overpower Requirement	None
Output Power Conditions	60 Hz
	Utility-grid quality
Noise Control	Consistent with typical gas turbine power plant practice
Power System Physical Design Objectives	
Installation	Outdoors
Transportation Options	Truck, sea, air, rail
Transportation Basis	500 miles
Plant Design Lifetime	Conventional power plant equipment - 25 years
Utilities	All utilities needed to support the plant will be available
Power Plant Operation	
Annual Operating Time	Fifty weeks
Annual Planned Shutdown	Two weeks
Instrumentation and Control	The I&C system design concept and cost will be based on the projected needs of mature fuel cell/gas turbine technologies and commercial power plant operation.
Economic Evaluation Parameters	
Power System Cost Estimation Basis	Mature technologies & integration techniques, not first-of-a kind.
Cost Basis	Mid-1998 US dollars
Fuel Cost	\$3.00/MMBtu (HHV)
Capital Charge Rate	15%
Capacity Factor	92%
COE Evaluation Method	Constant dollars
Power Plant Optimization Basis:	Consistent with >70% (LHV) efficiency, achieve a design-point COE that is 10-20% below the COE achieved by today's conventional-plant technology.
Basis for COE Comparison:	Comparably sized gas turbine cycle

In addition to the cost analysis basis positions established in the Reporting Requirements document, ERC's HEFPP project SOW called for additional specific costs sensitivity analyses and risk analysis, along with the development of the system design, cost estimate, and spreadsheet model. These project results for these activities are described in the following sections.

2.0 PROJECT RESULTS

The tasks in the program were as follows:

Task 1: Conceptual Electronic Systems Model, which involved the development of the system design, and the generation of a spreadsheet based model of the system.

Task 2: Equipment Sizing and Specification, based on the chosen system.

Task 3: Cost Estimating, based on the equipment sizing.

Task 4: Financial Analysis and Risk Analysis, which included cost of energy estimation, and evaluation of the sensitivity of the COE to fuel cost.

The project was initiated in June, 1998 and ran through March, 1999. Results of the task activities are described below.

2.1 Task 1: Conceptual Electronic Systems Model

2.1.1 Process Development

As noted above, at the start of the project, a hybrid baseline was in place based on the earlier CRADA work. The process development work in the HEFPP project had the objective of identifying a more simple hybrid design. Areas which were examined included advanced "stretch-goal" fuel cell concepts as well as improved hybrid system ideas. Initial studies were conducted in the following areas:

- Advanced Fuel Cell Concepts, projecting substantially higher fuel utilization or operating voltages than are currently obtainable.
- Advanced Fuel Cell Configurations, for example fuel cell networks, in which fuel cell stacks are bottomed by other fuel cell stacks. In this approach the residual fuel from one set of stacks can be sent to a second string of stacks for additional utilization.
- Improved hybrid systems, with the objective of identifying simpler designs, with fewer developmental components than the baseline.

When the FETC Reporting Requirements document was issued in September it stated that "system performance parameters used in these calculations should either be currently obtainable or any extrapolations should be justified." With this direction, it was determined that the appropriate fuel cell performance assumptions used in the project

should be the projections used in our baseline commercial plant design. This design is reviewed annually in DOE's due diligence effort. The fuel cell performance assumed in the design has been demonstrated in small cells and the technology effort required to achieve it in full size stacks is in place in ERC's DOE-supported Product Development and Improvement (PDI) program. ERC has other ongoing projects which are developing cell designs with higher performance levels, but these are longer term activities.

Given this direction, it was decided to pursue the advanced stack and stack configuration concepts only for comparison purposes. The advanced stack concept clearly looks to long term performance levels beyond the scope defined in the Reporting Requirements. To some extent the advanced stack configuration approach also needs longer term stretch goals to be feasible. The concept is equivalent to developing a stack technology which can operate at very high fuel utilizations. One drawback is that the capital cost and replacement cost of the additional stacks can become prohibitive unless stack life can be extended, or unless stacks can be developed which can operate efficiently on very dilute fuel. Both approaches should be considered in longer term studies, but were out of the scope of the current work.

The work therefore focused on evaluation of alternate hybrid systems which eliminated some of the complexity and developmental components from the baseline design. The first system evaluated was an alternate hybrid system in which residual fuel in the anode exhaust was utilized directly by combustion in the Brayton cycle. This eliminated the high temperature heat exchanger in the baseline hybrid system. Various system options were implemented and simulation runs were performed. While the system did eliminate the high temperature heat exchanger, it had issues which led to its rejection. The principle problem with the system was additional system complexity associated with cooling and drying the anode exhaust before compression.

The hybrid system which was evaluated next involved a return to the approach of indirectly transferring sensible heat to the turbine cycle. By refining the heat exchange approach used for this transfer, the temperature of the heat exchange equipment was reduced from the 2000°F in the baseline design to a more practical 1600°F. It was also decided to eliminate supplemental firing of fuel in the turbine cycle, the principle being that since the fuel cell is the most efficient conversion device in the system, all fuel should be sent to it. The development of this system then boiled down to extracting heat as efficiently as possible from our baseline fuel cell system.

The system design which resulted from these efforts is shown schematically in Figure 2.1. The system concept is as follows. Fuel and water are sent to the system Heat Recovery Steam Generation (HRSG) unit, where steam is produced and mixed with heated fuel for use as the fuel cell gas feed. The fuel cells - which produce 17 MW - operate at 78% fuel utilization, and residual fuel from the anode exit is consumed in an

anode exhaust oxidizer. Air is compressed in an intercooled compressor to about 16 atm, heated with system exhaust in the HRSG, heated further with exhaust from the anode exhaust oxidizer, and expanded in a turbine. The turbine, which incorporates a reheat cycle, produces about 3.4MW net power output. The expanded, low pressure air leaving the turbine is used as the oxidant in the anode exhaust oxidizer. Flue gas leaving the oxidizer is first cooled by the turbine air, and then sent as the cathode feed gas to the fuel cells. The cathode exhaust gas is sent through the HRSG to provide the required pre-heat and water vaporization.

Figure 2.1
Simplified System Schematic of Selected Hybrid Design

Attractive features of this system include the following:

- The system meets the project goal of LHV efficiency above 70%.
- The heat exchange temperature has been reduced to less than 1600°F, which is in the range of commonly available high temperature heat exchange materials.
- The Brayton cycle is an unfired system, indirectly heated with fuel cell waste heat. This yields the highest efficiency, since all primary fuel consumption is done in the fuel cell - which is the more efficient portion of the system.
- Cathode gas and anode gas recycle blowers have been eliminated. All fuel and oxidant supply is accomplished with the 15 psig natural gas pressure and the air compressor associated with the turbine equipment.

- High efficiency is obtained without the use of an additional steam bottoming cycle, eliminating the need for high pressure boilers (a concern with respect to unattended operation).
- Two key features of our baseline simple cycle DFC product are retained: Ambient pressure operation and internal reforming. This is the most efficient and reliable configuration of the fuel cell section of the power plant. It also avoids the need for a high pressure boiler to provide fuel cell steam.
- The system allows the turbine pressure ratio to be set independently of fuel cell pressure considerations.
- Although in normal hybrid operation the turbine will be unfired, a burner could be provided in the turbine to allow independent operation of the turbine, replacing the existing fuel cell startup burner. Conceptually, the turbine section could be producing power during the fuel cell startup.
- In principle, the turbine section could be used to load follow - utilizing stored kinetic energy - while the fuel cell is efficiently operated at constant power. This is only possible because of the decoupled nature of the fuel cell and turbine sections of the plant.

Fully exploiting the last two items in the list will involve additional study, and was beyond the scope of this project, which was limited to baseload cases without turndown or overpower requirements.

During the development of this system, Allison Engine Company was consulted to provide overall review of the flowsheet concepts, and to provide detailed turbine design information. Allison provided the turbine specifications and costs which supported the Tasks 2 and 3 equipment sizing and costing efforts. They also helped refine the modeling effort in Task 1 by providing turbine and compressor efficiency data based on our preliminary model runs. They also offered the observation that while the selected system is attractive and meets the project goals, it requires turbomachinery with intercooled compressors and expander reheat which are not off-the-shelf items. No technological issues stand in the way, but the normal costs of new product development and retooling would have to be incurred. It was noted that if a system could be configured without compressor intercooling and expander reheat, it could be a good match with existing Allison equipment.

This was pursued to the point where system configurations were identified which did not meet the 70% efficiency objective, but which had efficiencies in the mid-60's and which could use existing turbomachinery from the 501K family. While the 70% system remained the project baseline, these lower efficiency alternates are worth investigating in future studies.

A summary of the results of the process development activities described above is presented in Table 2.1.

Table 2.1
Summary of System Designs Evaluated
The Bold Column is the Selected System Design

	1	2	3	4	5	6	7
	Simple Cycles			Hybrids with Intercooling & Reheat		Hybrids with Single Stage Compression & Expansion	
Fuel Cell Performance Assumptions							
Cell Voltage, mV	846	826	817	826	817	817	817
Fuel Utilization, %	94	94	78	94	78	78	78
Fuel Cell Section							
DC Power Out, MW	22.27	22.27	22.03	22.27	17.62	17.62	17.62
AC Power Out, Gross	21.75	21.75	21.52	21.49	17.00	17.00	17.00
Brayton Cycle							
Compressor Power				3.84	5.84	4.95	3.79
Expander				5.58	9.35	6.60	5.90
Power Net				1.73	3.51	1.65	2.11
Net AC Out				1.70	3.44	1.62	2.07
Parasitic	0.26	0.26	0.26	0.03	0.03	0.03	0.03
Net Power Output	21.49	21.49	21.26	23.16	20.41	18.59	19.04
Overall LHV Efficiency	74%	72%	58%	78%	71%	65%	66%

The first three columns are simple cycle (fuel cell only) systems, which represent various levels of fuel cell performance projection. The first two columns are aggressive projections, incorporating long term fuel cell technology advances or fuel cell network concepts. These studies indicated that within the bounds of what is thermodynamically possible, simple cycle or networked DFC's could achieve efficiencies in the mid-70's in the very long term.

Columns four and five are for the selected hybrid system. The first hybrid incorporates stretch-goal fuel cell projections, and achieves an efficiency close to 80%. It can be

seen that the turbomachinery does not contribute much additional power to the system, since there is so little waste heat left over from the very efficient fuel cell. The second hybrid incorporates our standard commercial plant projections. The system is projected to operate at 71.08% LHV efficiency at the rated power output point of 20.41 MW. This is the system which was carried forward into the EXCEL^(TM) model development and costing tasks described below.

The last two columns, 6 and 7, are for hybrid systems incorporating existing single stage turbomachinery at two pressure ratios. The column 6 case (with a pressure ratio of 9, and 36 lb/sec of turbine exhaust flow) is a good match for Allison's 501KB7 in unboosted configuration. While these systems do not meet the 70% LHV target efficiency, they may offer opportunities to field relatively efficient systems sooner than the 2010 time frame.

2.1.2 EXCEL^(TM) Model Development

The HEFPP project statement of work called for the development of a model of the system for electronic reporting purposes and enhanced sharing with the technical community. It was requested that commonly available computer spreadsheet software such as Microsoft EXCEL^(TM) be used for this purpose. Once the process was defined by the work with ChemCAD^(TM) modeling, work began on preparing the deliverable spreadsheet model.

The model establishes power output, efficiency, key process performance parameters, process flows, gas compositions, temperatures, pressures, and mass and energy balances throughout the power plant system. The model is programmed with certain power plant design assumptions which are presented as reference values. The sensitivity to each of the key reference design assumptions can be evaluated by an input parameter deviation. The model iterates the flows and energies in the system based on the parameter deviation until there is energy, atom and mass balance convergence throughout the system. The resulting impact on the power plant performance and conditions throughout the system is displayed.

A description of the model and a complete "user's manual" is included as Volume II of this Final Report. The model analysis at the Rated Power Output condition of 20.41 MW is summarized in Figure 2.2 and Table 2.2. Figure 2.2 shows the detailed system schematic, with heat and mass balance node numbers on relevant streams. The 3-page Table 2.2 summarizes the heat and mass balance for the plant, referring to node numbers on Figure 2.2.

HG:hcf3a.node

Figure 2.2
Detailed System Schematic with Node Numbers

Table 2.2: High Efficiency Power Plant Heat and Mass Balance Summary

INPUT PARAMETERS		FIXED REFERENCE		+/- DEVIATION	OUTPUT CHARACTERISTICS	
CH4 INPUT, MOLES/HR	=	296.00	0		NET AC POWER, KW	= 20,411
PROCESS AIR FLOW, MOLES/HR	=	4570	0		NET AC LHV EFFICIENCY, %	= 71.08
AMBIENT AIR TEMP F	=	59	0		PARASITIC LOSSES, KW	= 31
WATER VAPOR PRESS, PSIA	=	0.147	0		GROSS AC POWER	= 20442
FUEL LOWER HEATING VALUE, BTU/SCF	=	872.5	N/A		PLANT EXHAUST TEMP, F	= 208.6
CELL STACK					FUEL CELLS	
CELL VOLTAGE	=	0.817	0		FUEL CELL AC OUTPUT, KW	= 17002
CURRENT DENSITY, AMPS/CM ²	=	0.16	0		FUEL CELL DC POWER, KW	= 17619
NUMBER STACKS		48	N/A		FUEL UTILIZATION, %	= 78.0
NUMBER CELLS/STACK	=	360	N/A		CATHODE EXIT TEMP, DEG F	= 1173.6
CELL AREA, M ²	=	0.78	N/A		CATHODE INLET TEMP, DEG F	= 1035.8
ANODE EXIT TEMP DEG C	=	1170	0		FUEL CELL OXYGEN UTILIZATION, %	= 54.0
METHANE CONVERSION, %	=	100	N/A		FUEL CELL INLET O2 FRACTION	= 0.1236
STACK HEAT LOSS, MBTU/HR	=	700,000	0		OUTLET O2 FRACTION	= 0.0711
INVERTER EFFICIENCY, %	=	96.5	0		FUEL CELL INLET CO2 FRACTION	= 0.1793
PARASITIC PWR, CONTROLS&MISC, KW	=	10	0		OUTLET CO2 FRACTION	= 0.0572
FUEL PROCESSING	=				GAS TURBINE	
PREREFORM EXIT PRES, PSIA	=	17	0		TURBINE GEN NETPOWER AC, KW	= 3440
PREREFORM INLET TEMP, C	=	752	E-3 EXIT		HI TURBINE INLET TEMPERATURE, F	= 1526
STEAM/CH4 RATIO	=	1.5	0		LOW TURBINE INLET TEMPERATURE, F	= 1220
RU TEMP, F	=	1100	0		HIGH TURBINE POWER, KW	= 8035
GAS TURBINE			0		LOW TURBINE POWER, KW	= 1315
GAS TURBINE COMP EFF, %	=	85	0		1ST STAGE COMPRESSOR WORK, KW	= 1827
GAS TURBINE TURB EFF, %	=	90	0		2ND STAGE COMPRESSOR WORK, KW	= 2002
GT COMPRESSOR PRESS RATIOS	=	2.63	0		3RD STAGE COMPRESSOR WORK, KW	= 2011
EACH INTERCOOLER PRESS LOSS, PSI	=	1	0			
COMPRESSOR DUCT AND INLET LOSSES, %	=	3	0			
PRESSURE LOSS, COMP OUT TO TURB IN, PSI	=	4	0		PROCESS WATER TOTAL, GAL/MIN	= 16
PRESSURE LOSS, E-9, PSI	=	1	0			
1ST TURBINE STAGE OUTLET PRESSURE, PSIA	=	23	0			
TURBINE EXIT DIFFUSER LOSSES, %, 1ST/2ND	=	7/5	0			
GENERATOR EFFICIENCY, %	=	98	0			
TURBINE STAGE HEAT LOSS, BTU/HR	=	10000	0			
HEAT EXCHANGER SUMMARY		REFERENCE	+/- DEVIATION		REQUIRED HEAT EXCHANGER EFFECTIVENESS	
E-1 SECOND STAGE AIR HEATER	COLD SIDE EXIT	1058	0		E-1	= 0.68
E-2 PROCESSED FUEL REHEATER	COLD SIDE EXIT	932	0		E-2	= 0.86
E-3 FUEL /STEAM HEATER	COLD SIDE EXIT	752	0		E-3	= 0.75
E-4 FIRST STAGE AIR HEATER	COLD SIDE EXIT	816	0		E-4	= 0.96
E-5 BOILER	COLD SIDE EXIT	239	0		E-5	= 0.81
E-6 NATURAL GAS HEATER	COLD SIDE EXIT	212	0		E-6	= 0.81
E-7 FEEDWATER HEATER	COLD SIDE EXIT	203	0		E-7	= 0.80
E-8 TURBINE INLET HEATER	COLD SIDE EXIT	1526	0		E-8	= 0.89
E-9 TURBINE INLET REHEATER	COLD SIDE EXIT	1220	0		E-9	= 0.84
E-10 INTERCOOLER	HOT SIDE EXIT	122	0		E-10	= 0.79
E-11 INTERCOOLER	HOT SIDE EXIT	122	0		E-11	= 0.86
E-12 COOLING TOWER	COLD SIDE EXIT	86	0		E-12	= 0.81

Table 2.2: High Efficiency Power Plant Heat and Mass Balance Summary, continued

STATION		TEMP DEG F	PRESS PSIA	ENTHALPY BTU/HR	COMPOSITION, VOLUME							FLOW MOLES/HR	FLOW LBS/HR
NO	DESCRIPTION				FRACTION								
					H2	H2O	CH4	CO	CO2	O2	N2		
1	NATURAL GAS IN	59.0	29.70	-7,949,462	0.0000	0.0000	0.9600	0.0000	0.0200	0.0000	0.0200	296.00	4,985
2	DESULFUR NAT GAS	15.0	24.70	-7,949,462	0.0000	0.0000	0.9600	0.0000	0.0200	0.0000	0.0200	296.00	4,985
3	HEATED NAT GAS	212.0	24.60	-7,551,780	0.0000	0.0000	0.9600	0.0000	0.0200	0.0000	0.0200	296.00	4,985
4	FUEL/STEAM/RECYCLE	233.7	19.20	-51,454,583	0.0012	0.5991	0.3835	0.0000	0.0083	0.0000	0.0080	752.02	13,187
5	PREREFORMER INLET	752.0	19.00	-47,683,213	0.0012	0.5991	0.3835	0.0000	0.0083	0.0000	0.0080	752.02	13,187
6	PREREFORMER EXIT	611.3	17.00	-47,682,947	0.0748	0.5448	0.3539	0.0001	0.0264	0.0000	0.0078	774.58	13,187
7	FUEL/STEAM TO E-2	611.3	16.71	-46,948,630	0.0742	0.5406	0.3511	0.0001	0.0262	0.0000	0.0077	768.57	12,984
8	FUEL/STEAM FROM E-2	932.0	16.61	-44,362,077	0.0742	0.5406	0.3511	0.0001	0.0262	0.0000	0.0077	768.57	12,984
9	FUEL TO STACK Rus	932.0	16.60	-44,362,077	0.0746	0.5357	0.3555	0.0001	0.0264	0.0000	0.0078	768.57	12,984
10	FUEL FROM Rus	1100.0	15.60	-30,524,258	0.5111	0.2058	0.1256	0.0734	0.0784	0.0000	0.0057	1,045.61	12,984
11	FUEL TO ANODES	1100.0	15.60	-30,524,258	0.5111	0.2058	0.1256	0.0734	0.0784	0.0000	0.0057	1,045.61	12,984
12	FUEL FROM ANODES	1170.0	15.52	-243,537,416	0.0715	0.3897	0.0000	0.0423	0.4938	0.0000	0.0027	2,195.09	66,199
13	FUEL TO BURNER	1170.0	15.50	-243,537,416	0.0715	0.3897	0.0000	0.0423	0.4938	0.0000	0.0027	2,195.09	66,199
14	BURNER PREMIX	1124.5	15.45	-199,470,000	0.0232	0.1332	0.0000	0.0137	0.1623	0.1398	0.5278	6,765.09	197,763
15	BURNER EXHAUST	1585.3	15.35	-199,469,960	0.0000	0.1593	0.0000	0.0000	0.1793	0.1236	0.5377	6,640.15	197,763
16	EXHAUST FROM E-8	1309.4	15.27	-216,413,113	0.0000	0.1593	0.0000	0.0000	0.1793	0.1236	0.5377	6,640.15	197,763
17	CATHODE INLET	1035.8	15.19	-232,676,546	0.0000	0.1593	0.0000	0.0000	0.1793	0.1236	0.5377	6,640.15	197,763
18	CATHODE EXHAUST	1173.6	15.11	-94,334,827	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
19	EXHAUST TO E-1	1173.6	15.15	-94,334,827	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
20	EXHAUST TO E-2	982.1	15.05	-102,758,864	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
21	EXHAUST TO E-3	922.3	15.00	-105,345,417	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
22	EXHAUST TO E-4	834.1	14.95	-109,116,787	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
23	EXHAUST TO E-5	446.8	14.85	-125,129,423	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
24	EXHAUST TO E-6	248.1	14.80	-133,024,387	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
25	EXHAUST TO E-7	238.0	14.75	-133,422,069	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
26	EXHAUST FROM SYSTEM	208.6	14.70	-134,573,876	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
27	PROCESS AIR IN	59.0	14.70	9,478,849	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
28	IST STAGE COMP IN	59.0	14.45	9,478,849	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
29	IST STAGE COMP OUT	253.7	38.01	15,714,514	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
30	2ND STAGE COMP IN	122.0	35.87	11,490,654	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
31	2ND STAGE COMP OUT	334.3	94.35	18,322,460	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
32	3RD STAGE COMP IN	122.0	90.52	11,490,654	0.0000	0.0583	0.0000	0.0000	0.0091	0.6273	2.3636	4,570.00	131,564
33	3RD STAGE COMP OUT	335.4	238.06	18,355,459	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
34	PREHEATER E-4 OUT	816.0	236.73	34,368,095	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
35	PREHEATER E-1 OUT	1058.0	235.39	42,792,133	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
36	IST TURBINE IN	1526.0	231.68	59,735,286	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
37	IST TURBINE OUT	755.6	25.81	32,302,844	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
38	2ND TURBINE IN	1220.0	23.00	48,566,276	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
39	2ND TURBINE OUT	1094.0	15.50	44,067,163	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564

Table 2.2: High Efficiency Power Plant Heat and Mass Balance Summary, continued

STATION		TEMP DEG F	PRESS PSIA	ENTHALPY BTU/HR	COMPOSITION, VOLUME							FLOW	
NO	DESCRIPTION				FRACTION							MOLES/HR	LBS/HR
					H2	H2O	CH4	CO	CO2	O2	N2		
40	PROCESS WATER IN	59.0	30	-52,214,531	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	444.00	7,999
41	TREATED WATER	59.0	24	-52,214,531	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	444.00	7,999
42	FEEDWATER HTR OUT	203.0	23	-51,062,724	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	444.00	7,999
43	STEAM FROM BOILER	239.0	22	-43,167,937	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	444.00	7,999
44	STEAM TO EJECTOR	239.0	21.5	-43,167,937	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	444.00	7,999
45	FUEL RECYCLE	611.3	17	-734,257	0.0742	0.5407	0.3512	0.0000	0.0262	0.0000	0.0077	12.02	203
46	STEAM/ RECYCLE MIX	250.1	19.5	-43,902,319	0.0020	0.9879	0.0093	0.0000	0.0007	0.0000	0.0002	456.02	8,202
47	COOLING PMP INLET	86.0	29	-608,992,883	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	5,200.00	93,683
48	COOLING PMP EXIT	86.0	36	-608,992,883	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	5,200.00	93,683
49	COOLANT TO E-10	86.0	36	-234,228,032	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	2,000.00	36,032
50	COOLANT FROM E-10	203.2	34	-230,004,171	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	2,000.00	36,032
51	COOLANT TO E-11	86.0	36	-374,764,851	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	3,200.00	57,651
52	COOLANT FROM E-11	204.5	34	-367,933,045	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	3,200.00	57,651
53	COOLANT TO E-12	204.0	34	-597,937,610	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	5,200.00	93,683
54	AIR TO TOWER	59.0	14.7	248,897,563	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	119,999.88	3,454,636
55	AIR FROM TOWER	72.2	14.7	259,952,836	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	119,999.88	3,454,636

HEAT & MASS BALANCE AROUND THE ENTIRE PLANT

SYSTEM ENERGY AND MOLE BALANCE

STATION		ENERGY	MASS, LB/HR	MOLES				
		BTU/HR		H2	O2	C	N2	
IN	FUEL IN	1	-7,949,462	4,985	568.3	5.9	290.1	5.9
	PROCESS AIR IN	27	9,478,849	131,564	45.7	982.5	13.7	3564.6
	WATER IN	40	-52,214,531	7,999	444.0	222.0	0.0	0.0
	COOLING AIR IN	54	248,897,563	3,454,636	1200.0	25799.8	360.0	93599.9
	TOTAL		198,212,419	3,599,184	2258.0	27010.3	663.8	97170.4
OUT	FUEL CELL DC POWER		60,133,647					
	EXHAUST	26	-134,573,876	144,548	1058.0	1210.5	303.8	3570.5
	COOLING AIR OUT	55	259,952,836	3,454,636	1200.0	25799.8	360.0	93599.9
	GT POWER		11,978,915		2258.0	27010.3	663.8	97170.5
	GT HEAT LOSS		20,000					
	STACK HEAT LOSS		700,000					
	TOTAL		198,211,523	3,599,184				
	ERROR (IN-OUT)		896	0	0.0	0.0	0.0	0.0
	(IN-OUT)/ IN , %		0.0005					

2.2 Task 2: Equipment Sizing and Specification

The objective of this task was to develop specifications for the system and individual major equipment items adequate for general disclosure and to the detail necessary to make judgments about system feasibility within 2010 time frame. Specifications were also required to a sufficient level of detail to allow the generation of an accurate cost estimate in Task 3.

The equipment sizing work had its basis in the process modeling activity of Task 1. Once the system was selected and refined, and a complete process model was developed, the equipment specifications could be derived from the modeled predictions of equipment duties. In the case of this project two models were available - ChemCAD^(TM) and EXCEL^(TM) - which allowed for cross-checking of the system parameters.

With respect to feasibility within the next decade, there is no question that all of the specified equipment can be available in this time frame. As noted above, the new hybrid system eliminated the developmental items which existed in the earlier system. The equipment in the system falls into the following categories:

- Developmental Fuel Cells, which are projected to commercially available in the next two years, and performing at the projected level in the next four years. The DC power section includes 12 4-stack fuel cell modules with exactly the same design as planned for the baseline simple cycle MW scale powerplants.
- Multi-stage turbomachinery, which is not commercially available in exactly these sizes now, but which embodies conventional technology. This equipment could certainly be made available whenever market conditions or policy direction trigger the design and tooling investment.
- Conventional process equipment, such as heat exchangers, etc.

Perhaps the most developmental non-fuel cell item is the anode exhaust oxidizer. This is a catalytic reactor which reacts anode tail gas with air. Similar units have been built for the 2MW Santa Clara plant and ERC's full height stack test facility. The only difference here is that the flue gas temperature is hotter, 1600°F vs 1300°F. This can be handled with a staged approach, in which some of the fuel is reacted with air to provide a 1300 °F flue gas (in a reactor similar to those used at SCDP and ERC), and additional fuel can then be added in non-catalytic sections of the oxidizer to provide the higher temperatures from auto-ignition.

The equipment specifications are summarized in Table 2.3, which lists the equipment, capacities, materials of construction, and design characteristics for the major equipment items. The configuration of equipment in the plant has been evaluated and is illustrated in Figure 2.3, which shows the preliminary power plant layout.

Table 2.3: Major Equipment List

ITEM #	MAJOR EQUIPMENT ITEM	UNIT CAPACITY	MATERIAL OF CONSTRUCTION	DESIGN CHARACTERISTICS
	AREA 1 - FUEL CELL SUB-SYSTEM			
1	FUEL CELL MODULES	12 MODULES, 4 STACKS/MODULE		380 CELLS/STACK, CELL AREA 0.78M ²
2	DESULFURIZER	1872 SCFM Natural Gas	CARBON STEELS	ACTIVATED CHARCOAL
3	ANODE EXHAUST OXIDIZER	27.6 MMBTU/HR	STAINLESS STEELS	CATALYTIC INLET SECTION
4	HEAT RECOVERY UNIT(2 REQUIRED)			
4.1	E-1 SECOND STAGE AIR HEATER	4.5 MMBTU/HR, EFFECTIVENESS=0.68	STAINLESS STEEL	FINNED TUBES, MULTI-PASS, CROSS COUNTER FLOW,
4.2	E-2 PROCESSED FUEL HEATER	1.25 MMBTU/HR, EFFECTIVENESS=0.86	STAINLESS STEEL	FINNED TUBES, MULTI-PASS, CROSS COUNTER FLOW,
4.3	E-3 FUEL/STEAM HEATER	1.9 MMBTU/HR, EFFECTIVENESS=0.75	STAINLESS STEEL	FINNED TUBES, MULTI-PASS, CROSS FLOW
4.4	E-4 FIRST STAGE AIR HEATER	8.85 MMBTU/HR, EFFECTIVENESS = 0.96	STAINLESS STEEL	FINNED TUBES, MULTI-PASS, CROSS COUNTER FLOW
4.5	E-5 BOILER	3.85 MMBTU/HR, EFFECTIVENESS = 0.81	STAINLESS STEEL	FINNED TUBES, 1 PASS, CROSS FLOW
4.6	E-6 NATURAL GAS HEATER	0.2 MMBTU/HR, EFFECTIVENESS=0.81	STAINLESS STEEL	FINNED TUBES, 1 PASS, CROSS FLOW,
4.7	E-7 FEEDWATER HEATER	0.6 MMBTU/HR, EFFECTIVENESS =0.80	STAINLESS STEEL	FINNED TUBES, 1 PASS, CROSS FLOW
5	WATER TREATMENT SYSTEM	5 MHP, 4 GPM R.O. FEED PUMP, (2) 7 GPM, .75 MHP TRANSFER PUMPS, 2500 GAL WATER STORAGE TANK	MISC	PACKAGED ASSEMBLY
6	STEAM JET EJECTOR	DELTA P = 2 PSI		
7	PRECONVERTER	1851 SCFM Natural Gas		
8	PIPING & PIPE INSULATION		STAINLESS STEEL	

Table 2.3: Major Equipment List, continued

ITEM #	MAJOR EQUIPMENT ITEM	UNIT CAPACITY	MATERIAL OF CONSTRUCTION	DESIGN CHARACTERISTICS
9	POWER CONDITIONING SYSTEM	17 MVA		1200 VDC IN / 440 VAC 3 PHASE OUT
10	DC CABLING	1250 AMPS/STACK MODULE	NI / CU	1200 VOLTS
11	MISC ELECTRICAL EQUIPMENT			
12	CONTROLS & INSTRUMENTS			
	AREA 2 - GAS TURBINE CYCLE			
13	GAS TURBINE			PACKAGED UNIT BY GAS TURBINE VENDOR
13.1	ELECTRIC GENERATOR	3.2MVA		
13.2	1ST STAGE COMPRESSOR	1849 hp		
13.3	2 ND STAGE COMPRESSOR	2882 hp		
13.4	3 RD STAGE COMPRESSOR	2787 hp		
13.5	HIGH PRESSURE TURBINE	10,864 hp		
13.6	LOW PRESSURE TURBINE	1568 hp		
14	E-10 FIRST INTERCOOLER	4.14 MMBTU/HR, EFFECTIVENESS= 0.79	ALUMINUM	OFF ENGINE- PLATE/FIN COUNTER FLOW,
15	E-11 SECOND INTERCOOLER	7.1 MMBTU/HR, EFFECTIVENESS= 0.86	ALUMINUM	OFF ENGINE -PLATE/FIN COUNTER FLOW,
16	E-12 COOLING UNIT	11.3 MMBTU/HR, EFFECTIVENESS= 0.81	COPPER	FINNED TUBES,, FORCED DRAFT, DRY , 15 - 1.5 HP FANS
17	COOLANT PUMP	100 GPM @ 25' HEAD		3 PUMPS, 2 OPERATING , ONE SPARE
18	AIR FILTER	31,200 SCFM		
19	START BURNER			
20	E-8 TURBINE INLET HEATER(2 REQUIRED)	9.08 MMBTU/HR, EFFECTIVENESS=.89	INCONEL 600	5/8" BARE TUBES, 3 PASS, CROSS COUNTER FLOW, 5378 FT ² SURFACE
21	E-9 TURBINE REHEATER (2 REQUIRED)	7.8 MMBTU/HR, EFFECTIVENESS=.84	STAINLESS STEEL	5/8" BARE TUBES, 3 PASS, CROSS COUNTER FLOW, 6239 FT ² SURFACE
22	PIPING & PIPE INSULATION			
	AREA 3- OVERALL PLANT			
23	OUTPUT TRANSFORMER	20.2 MVA		440 VAC IN / 13.8KV OUT
24	SWITCHGEAR & MISC ELECTRICAL			
25	POWER PLANT CONTROLLER			MICROPROCESSOR WITH CRT INTERFACE
26	UPS			5 kW FOR CONTROL POWER, 24 HOUR BATTERY

Figure 2.3: Preliminary Layout of 20MW Hybrid Powerplant

2.3 Task 3: Cost Estimating

The equipment list and preliminary layout were used to develop a capital cost estimate for the powerplant. The costs were estimated by various methods, depending on the type of cost element under consideration. For equipment costs, the following approaches were used:

- Budgetary estimates from potential suppliers were sought and obtained for equipment which was very specific to this powerplant design. This was done for the heat exchangers in the HRSG because of their unique duties and, in some cases, their high required effectiveness. This was also done for the dry cooling tower used to remove heat from the air compressors. The turbine section of the powerplant was also estimated this way, using a budgetary estimate provided by Allison Engine Company. Allison provided a "fully learned" estimate of \$600 per kW of turbine output. It was noted that this does not include first-of-a-kind development and tooling costs, which would impact early units depending on how those costs are covered and what production volumes are achieved. The fully learned estimate is consistent with the cost methodology of this study ("mature technologies and integration techniques, not first-of-a kind" per the Reporting Requirements), and was used for the plant cost estimate.
- Richardson Process Plant Construction Estimating Standards was used to estimate the cost of piping and insulation, and to estimate construction and installation costs.
- ERC's 3 MW commercial power plant cost estimate was used as a source of equipment cost data for items which would simply scale up to the larger plant. The cost elements were escalated from 1995 dollars to mid 1998 dollars using escalation factors from the Plant Cost Index of Chemical Engineering magazine (McGraw-Hill). Scale-up to the size required for the 20MW power plant was done based on the ratio of process flow between the larger equipment and the 3MW plant equipment. This scaleup factor was used with a cost power factor to derive the cost of the larger unit. The cost power factors were obtained from "*Design Cost Factors for Scaling-up Engineering Equipment*", S.S. Remer & L.H. Chai, Chemical Engineering Progress, August 1990.
- ERC's Internal database of costs for similar equipment based on other ERC projects which are collecting similar types of equipment costs.

The estimate for equipment and installation costs is summarized Table 2.4. Additional costs (land, startup, training, etc) are not included in Table 2.4, but were estimated in the cost of energy calculation, described next.

Table 2.4: Power Plant Equipment and Installation Cost Summary

ITEM	MAJOR EQUIPMENT ITEM	1995\$ 3.08 MW COST	ESCLTMN FCTR 95 TO 98\$	EQUIP COST/UNIT	COST BASIS	20MW FACTOR	COST FACTOR	COST/UNIT 20MW PLANT	SALES TAX/ UNIT	FREIGHT COST/UNIT	'INSTALLATION MHR\$ COST/UNIT		TOTAL COST/UNIT	QUANTITY	TOTAL COST, \$	COST \$/KW	
AREA 1 - FUEL CELL SUB-SYSTEM																	
1	FUEL CELL MODULES	\$509,748	1.01	\$514,845	"3"	0.96	0.96	\$493,119	\$0	\$14,250	1000	\$31,000	\$538,369	12	\$6,460,428	316.5	
2	DESULFURIZER	\$25,500	1.05	\$26,775	"3"	6.25	3.48	\$93,101	\$0	\$1,250	16	\$496	\$94,847	1	\$94,847	4.6	
3	ANODE EXHAUST OXIDIZER	\$105,352	1.05	\$110,620	"3"	3.27	2.24	\$247,422	\$0	\$1,250	40	\$1,240	\$249,912	2	\$499,823	24.5	
4	HEAT RECOVERY UNIT, INCLUDES				"1"			\$440,000	\$0	\$5,004	200	\$6,200	\$451,204	2	\$902,408	44.2	
4.1	E-1 SECOND STAGE AIR HEATER															0.0	
4.2	E-2 PROCESSED FUEL HEATER															0.0	
4.3	E-3 FUEL /STEAM HEATER															0.0	
4.4	E-4 FIRST STAGE AIR HEATER															0.0	
4.5	E-5 BOILER															0.0	
4.6	E-6 NATURAL GAS HEATER															0.0	
4.7	E-7 FEEDWATER HEATER															0.0	
5	WATER TREATMENT SYSTEM	\$41,087	1.05	\$43,141	"3"	5.95	4.09	\$176,585	\$0	\$6,000	100	\$3,100	\$185,685	1	\$185,685	9.1	
6	STEAM JET EJECTOR	\$13,200	1.05	\$13,860	"3"	2.71	1.68	\$23,281	\$0	\$100	4	\$124	\$23,505	2	\$47,011	2.3	
7	PRECONVERTER	\$17,172	1.05	\$18,031	"3"	2.71	1.97	\$35,527	\$0	\$200	16	\$496	\$36,223	2	\$72,446	3.5	
8	SS PIPING& PIPE INSULATION				"2"			\$595,856	\$0	INCLUDED		INCLUDED	\$595,856	2	\$1,191,712	58.4	
9	POWER CONDITIONING SYSTEM	\$397,539	1.03	\$409,465	"3"	2.76	2.49	\$1,020,936	\$0	\$6,000	100	\$3,100	\$1,026,936	2	\$2,053,871	100.6	
10	DC CABLING/CONDUIT				"2"			\$22,500	\$0	INCLUDED		INCLUDED	\$22,500	1	\$22,500	1.1	
11	MISC ELECTRICAL EQUIPMENT	\$15,000	1.03	\$15,450	"3"	2.76	2.04	\$31,444	\$0	\$500	100	\$3,100	\$34,544	2	\$69,088	3.4	
12	CONTROL VALVES &INSTRUMENTS	\$123,600	0.99	\$122,364	"3"	2.71	1.97	\$241,104	\$0	\$3,000	140	\$4,340	\$245,444	2	\$490,888	24.1	
AREA 2 - GAS TURBINE CYCLE																	
1	GAS TURBINE/GENERATOR				"1"			\$3,000,000	\$0	\$5,200	200	\$6,200	\$3,011,400	1	\$3,011,400	147.5	
2	E-10 FIRST INTERCOOLER				"4"			\$30,000	\$0	\$341	20	\$620	\$30,961	1	\$30,961	1.5	
3	E-11 SECOND INTERCOOLER				"4"			\$40,000	\$0	\$278	20	\$620	\$40,898	1	\$40,898	2.0	
4	E-12 COOLING TOWER				"1"			\$25,000	\$0	\$1,500	60	\$1,860	\$28,360	1	\$28,360	1.4	
5	COOLANT PUMP				"2"			\$4,950	\$0	\$28	23	\$713	\$5,691	3	\$17,073	0.8	
6	AIR FILTER			\$4,052	"5"	11.00	5.11	\$20,693	\$0	\$400	24	\$744	\$21,837	1	\$21,837	1.1	
7	OFF ENGINE START BURNER			\$6,876	"5"	5.20	3.07	\$21,097	\$0	\$1,000	40	\$1,240	\$23,337	2	\$46,674	2.3	
8	TURBINE INLET HEATER UNITS															0.0	
8.1	E-8 TURBINE INLET HEATER				"4"			\$210,000	\$0	\$1,600	20	\$620	\$212,220	2	\$424,440	20.8	
8.2	E-9 TURBINE REHEATER				"4"			\$120,000	\$0	\$1,800	20	\$620	\$122,420	2	\$244,840	12.0	
9	SS PIPING& PIPE INSULATION				"2"			\$261,710	\$0	INCLUDED		INCLUDED	\$261,710	1	\$261,710	12.8	
AREA 3 - OVERALL PLANT																	
1	TRANSFORMERS&MISC ELEC	\$26,400	1.03	\$27,192	"3"	5.52	3.31	\$89,902	\$0	\$6,000	250	\$7,750	\$103,652	1	\$103,652	5.1	
2	MOTOR CONTROL CENTER	\$15,700	1.03	\$16,171	"3"	1.00	1.00	\$16,171	\$0	\$2,000	8	\$248	\$18,419	1	\$18,419	0.9	
3	BULKS(ELECTRICAL)	\$74,200	1.03	\$76,426	"3"	5.52	3.31	\$252,679	\$0	\$3,000	2000	\$62,000	\$317,679	1	\$317,679	15.6	
4	POWER PLANT CONTROLLER	\$28,000	0.99	\$27,720	"3"	3.00	3.00	\$83,160	\$0	\$500	500	\$15,500	\$99,160	1	\$99,160	4.9	
5	CONTROL CONDUIT	\$11,400	0.97	\$11,058	"3"	5.52	3.31	\$36,560	\$0	\$1,000	500	\$15,500	\$53,060	1	\$53,060	2.6	
6	UPS	\$25,500	1.03	\$26,265	"3"	3.00	2.16	\$56,671	\$0	\$2,000	50	\$1,550	\$60,221	1	\$60,221	3.0	
7	I&C AIR SYSTEM	\$34,766	1.08	\$37,547	"3"	5.42	3.55	\$133,421	\$0	\$600	50	\$1,550	\$135,571	1	\$135,571	6.6	
8	MISC CRBN STEEL PIPNG&VALVES	\$10,500	0.99	\$10,395	"3"	5.42	3.16	\$32,815	\$0	2000	125	\$3,875	\$38,690	1	\$38,690	1.9	
9	FIRE PROTECTION(SUBCNTRCT)	\$51,100	1.01	\$51,611	"3"	6.00	3.38	\$174,536	\$0	INCLUDED		INCLUDED	\$174,536	1	\$174,536	8.6	
10	CIVIL WORKS	\$85,400	1.04	\$88,816	"3"	6.00	3.38	\$300,354	\$0	INCLUDED		INCLUDED	\$300,354	1	\$300,354	14.7	
TOTALS																\$17,520,242	858.4

Cost Basis Code Explanations:

"1" - Budgetary estimate from potential supplier

"2" - Estimate from conceptual design based on Richardson Process Plant Construction Estimating Standards

"3" - Estimate based on appropriate factor from detailed design of ERC 3 MW commercial plant design and cost, independently reviewed in DOE due diligence activity.

"4" - Estimate based on design sizing and cost /lb from quotations on similar equipment

"5"-Estimate based on appropriate factor from design and cost estimate for other application

The process model of the power plant was used to estimate process consumables, and manning forecasts were made to estimate operating labor costs. Fixed and variable cost estimates for the plant are summarized in Table 2.5

Table 2.5: Hybrid Plant Operating & Maintenance Costs

Fixed Costs	<u>Annual Cost</u>
Operating Labor (1000 Mhrs/Yr)	40,800
Maintenance Labor (1000 Mhrs/Yr)	45,600
Management Labor (400 Mhrs/Yr)	24,000
Other (non-staff)	10,000

Total Fixed Costs	120,400
Variable Costs	
Consumables	
Water & Sewer (\$1.7/1000 gallons)	12,520
Catalysts and Sorbents	48,130
Water Treatment	50,900
Miscellaneous	17,400
Maintenance During Plant Shutdown	160,968
(estimated @ 2% of fuel cell BOP cost)	-----
Total Variable Costs	289,918
Total Fixed and Variable Costs	410,318
Fuel Cell Module Replacement (5 year intervals)	6,460,428
25 year annualized cost of 4 replacements	1,033,668

These O&M costs were combined with estimates of additional cost, such as start-up, training, home office costs, etc to derive a cost of energy (COE) estimate, using the calculation assumptions specified in the Reporting Requirements. A summary of the COE calculation is shown in Table 2.6.

A cost of energy of 4.45 ¢/kWh was estimated, and this is the value used in comparisons and sensitivity studies in Task 4. An additional exercise was done in which a learning-type cost reduction occurred over the course of the 25-year analysis period, reducing the cost of replacement stack modules. The estimated COE with replacement stacks on 95, 90, and 85 percent learning curves is shown at the bottom of Table 2.6.

Table 2.6: Cost of Energy Calculation for Hybrid Power Plant

		\$	\$/KW
EQUIPMENT & INSTALLATION COST, TABLE 2.4		\$17,520,242	858.4
START-UP & TRAINING (1)	2%	\$350,405	17.2
LAND 1 ACRE		\$10,000	0.5
HOME OFFICE COST(2)	3%	\$525,607	25.8
ALLOCABLE COSTS(3)	3%	\$525,607	25.8
PROFIT	5%	\$876,012	42.9
TOTAL CAPITAL REQUIREMENT		\$19,807,873	970.5
CAPITAL CHARGE RATE (25 YEAR),%		15	
CAPACITY FACTOR ,%		92	
FUEL COST, \$/MMBtu(HHV)		3	
EFFICIENCY (LHV),%		71.1	
EFFICIENCY (HHV),%		63.99	
FUEL CELL RATING, kW		17002.00	
GAS TURBINE RATING,kW		3440.00	
POWER PLANT RATING, KW		20,411	
YEARLY POWER GENERATION, KWHRS		164,496,331	
COST OF ELECTRICITY(25 YEAR LEVELIZED)Constant \$			
		Yearly Cost	Cents/kWhr
CAPITAL RECOVERYCOST		\$2,971,181	1.806
OPERATING & MAINTENANCE COST			
FC FIXED (Tab. 2.5)		\$120,400	0.073
FC VARIABLE (Table 2.5)		\$289,918	0.176
GAS TURBINE, 0.01\$/kWhr		\$277,236	0.169
FUEL COST		\$2,632,219	1.600
FUEL CELL MODULE REPLACEMENT(ANNUALIZED)		\$1,033,668	0.628
COST OF ENERGY	TOTALS	\$7,324,623	4.453
TOTAL COST OF ELECTRICITY	LEARNING RATE(4)		
WITH PRODUCTION LEARNING	5%	\$7,143,154	4.3424
ON REPLACEMENT MODULES	10%	\$6,959,767	4.2310
	15%	\$6,864,745	4.1732
(1) Includes plant commissioning, acceptance testing, & Plant Buyers O&M staff training			
(2) Includes project management, Engineering & Site specific design, procurement, finance, clerical			
(3) Includes costs for legal and contracting, insurance, marketing, warranty reserves, interest, royalty			
(4) Production cost reduced by learning rate for each doubling of production quantity			

2.4 Task 4: Financial Analysis and Risk Analysis

2.4.1 Comparison and Sensitivity Analyses

This task called for the development of a financial analysis of the system. This includes an analysis of sensitivity of the cost of electricity (COE) to fuel price assuming the natural pipeline gas costs of \$1 to \$6 per million British thermal units (BTU's). A risk assessment of the costs estimated for each major cost category was also done.

In addition to the COE sensitivity analysis identified in the project SOW, the Reporting Requirements also asked for comparison with conventional generating equipment using \$3/MMBtu fuel. In this task a COE for a combined cycle power plant of similar size was estimated, and the estimate is summarized in Table 2.7. In this analysis the COE for the conventional powerplant is 5 to 12 percent more than the hybrid system, depending on whether or not replacement stack learning cost reduction is considered.

These same sets of assumptions were used in the sensitivity analysis, in which the conventional combined cycle plant was compared to the hybrid (in this case without replacement stack learning cost reduction) over a range of fuel costs. The results of this analysis are shown in Figure 2.4, which shows that even without stack cost reduction from learning assumptions, the plant COE is competitive with the conventional combined cycle plant for gas costs of \$2/MMBtu and higher.

Figure 2.4
Cost of Energy Comparison with Conventional Combined Cycle

Table 2.7
Cost of Energy Calculation for Conventional Plant

Reference: 1. Gas Turbine World Handbook General Electric STEG LM 116			
		\$	\$/KW
EQUIPMENT & INSTALLATION COST, Ref 1.		\$15,418,200	840.0
NATURAL GAS COMPRESSOR		\$594,700	32.4
START-UP & TRAINING (1)	2%	\$320,258	17.4
LAND 1 ACRE		\$10,000	0.5
HOME OFFICE COST(2)	3%	\$480,387	26.2
ALLOCABLE COSTS(3)	3%	\$480,387	26.2
PROFIT	5%	\$800,645	43.6
TOTAL CAPITAL REQUIREMENT		\$18,104,577	986.4
CAPITAL CHARGE RATE (25 YEAR),%		15	
CAPACITY FACTOR ,%		92	
FUEL COST, \$/MMBtu(HHV)		3	
EFFICIENCY (LHV),%(5)		48.8	
EFFICIENCY (HHV),%		43.93	
GAS TURBINE RATING,kW		13100.00	
STEAM PLANT RATING, kW		5300.00	
POWER PLANT RATING, kW(5)		18,355	
YEARLY POWER GENERATION, KWHRS		147,926,616	
COST OF ELECTRICITY(25 YEAR LEVELIZED)Constant \$			
		Yearly Cost	Cents/kWhr
CAPITAL RECOVERYCOST		\$2,715,687	1.836
OPERATING & MAINTENANCE COST			
FIXED (4.7\$/kW yr)(4)		\$86,269	0.058
VARIABLE (4.7MILLS/KWHR)		\$695,255	0.470
FUEL COST		\$3,447,757	2.331
TOTALS		\$6,944,967	4.695
(1) Includes plant commisioning,acceptance testing, & Plant Buyers O&M staff training			
(2) Includes project management, Engineering & Site specific design,procurement, finance, clerical			
(3) Includes costs for legal and contracting, insurance, marketing, warranty reserves, interest, royalty			
(4) Based EPRI TAG for 120MW combined cycle escalated 20% for 1998 \$			
(5) Corrected for 345kW gas compressor power consumption			

2.4.2 Risk Analysis

The cost estimate was evaluated in a risk analysis effort, in which probabilistic curves were assigned to individual cost estimate line items, and the impact on projected system cost was assessed. Crystal Ball^(TM) software was used for this analysis. This software is a Microsoft EXCEL^(TM) add-in which runs over the spreadsheet based cost estimate and uses a Monte-Carlo simulation to evaluate the overall probability of meeting or exceeding the projected cost. A similar analysis is done annually on the cost projection for the baseline simple cycle commercial product as part of the due diligence activity under the PDI program.

For the purposes of this project, the cost risks were estimated at the level of line items in the summary forecast, as presented earlier in Table 2.4. Each of the line item estimates was evaluated in terms of the probability of the actual cost meeting or exceeding the estimate. Distribution curves were developed for each line item, and fed into a Monte-Carlo simulation to assess the probability of achieving the projected cost. The rationale for the major line item probability assumptions was as follows:

Fuel Cell Modules

In the case of the fuel cell modules, the distribution which was selected was a Wiebull distribution which was shaped to resemble the distribution derived from a detailed analysis in the last due-diligence effort. That effort evaluated cost probabilities for components and raw materials in the stacks and stack modules, and developed a predicted probability curve. That analysis indicated a slight probability that the stack module costs could be less than the value used in the estimate due to the impact of potential cost reductions from ongoing manufacturing process development.

Gas Turbine System

A normal distribution (with a 10% standard deviation) was assigned to the gas turbine part of the hybrid plant. The additional complexities of the multi-stage system provide upward cost pressure, but on the other hand the unit runs at low temperature with clean air as the working fluid. Additional design detail will be needed to identify a clear likelihood for the costs to go one way or the other beyond the current estimate provided by Allison Engine Company.

BOP Equipment Quoted Specifically For This Project

In cases where equipment cost quotes were obtained from vendors, a Wiebull curve with a shape factor of 8 was used for the probability distribution. This curve essentially provides a fairly narrow distribution, with a limited likelihood for the cost to be above the mean estimate.

BOP Equipment Scaled From The 3MW System Design

For equipment which was scaled for the 3MW design, the probability distributions used in the last PDI due diligence study were used for this study (with the exception of the anode exhaust oxidizer). These were usually normal distributions with 5% standard deviations.

Anode Exhaust Oxidizer

In the case of the anode exhaust oxidizer, a Wiebull curve with a shape factor of 3 was used, and the location was set to indicate a likelihood that the cost could be higher than the estimate. This is due to the specialized application of the oxidizer in the hybrid system - operating at higher temperature and requiring a staged mixing between tail gas and air.

Once these probability assumptions were fed into the model, the program ran a Monte-Carlo simulation with 20,000 trials to assess how the assigned probability impacted the \$17,520,242 estimated installed equipment cost. The simulation projected an 80.8% probability that the system cost would be equal to or lower than the overall estimated cost. The probability distribution output from the program is shown in Figure 2.5.

Figure 2.5
Risk Analysis Probability Distribution for Powerplant Cost

3.0 CONCLUSION

The HEFPP project has identified an attractive DFC/Turbine hybrid design which offers the promise of >70% efficiency in a relatively simple system. The system is a significant improvement on earlier hybrid configurations due to the elimination of developmental equipment and the fact that it is a much more simple system. Cathode gas and anode gas recycle blowers have been eliminated. All gas supply is accomplished with the normal 15 psig natural gas pressure and the air compressor associated with the turbine equipment. High efficiency is obtained without the use of an additional steam bottoming cycle, eliminating the need for high pressure boilers (a concern with respect to unattended operation).

Two key features of our baseline simple cycle DFC product are retained: Ambient pressure operation and internal reforming, providing higher efficiency and avoiding the need for a high pressure boiler to provide fuel cell steam. The system allows the turbine pressure ratio to be set independently of fuel cell pressure considerations. Although in normal hybrid operation the turbine will be unfired, a burner could be provided in the turbine to allow independent operation of the turbine, replacing the existing fuel cell startup burner. Conceptually, the turbine section could be producing power during the fuel cell startup. Also conceptually, the turbine section can be used to load follow - utilizing stored kinetic energy - while the fuel cell is efficiently operated at constant power. This is only possible because of the decoupled nature of the fuel cell and turbine sections of the plant.

The cost estimate developed for the powerplant projected an installed equipment cost of \$858/kW. Cost of energy calculations indicated the system is cost competitive with conventional combined cycle equipment for cases where natural gas cost was \$2/MMBtu or greater. This analysis did not take into account cost credits or penalties which may be imposed as a matter of policy due to environmental or energy independence concerns. Given the likelihood of such policy items in the future, the possibility of providing products with these types of efficiencies is even more attractive than this simple cost analysis suggests.

Given the excellent results of this first phase of work, some consideration has been given to what the next steps should be. Additional work which was out of the scope of the study and which needs to be done on the design includes:

- Analysis of performance issues at partload.
- Startup, shutdown, and control issues.
- Evaluation of possible independent operation of turbine, e.g. during fuel cell heatup
- Evaluation of near-term options for proof-of-concept, field trials, and early market entry, including:

- Smaller systems. Hybrids based on ERC's market entry sub-MW, 1.5 MW and 3 MW power plants could offer very high efficiencies at their respective size levels.
- Field-test powerplants, consisting of hybrids with ERC's 200kW, 1MW, or 2MW near term field test systems.
- Detailed design of system-specific components. An advantage of the selected system is that there are no long term developmental issues in the balance of plant (like the high temperature heat exchanger from the earlier system). Certain components are unique to the hybrid system, and will require additional design specification. Examples include the anode exhaust oxidizer (which operates at a slightly higher temperature than the one in the baseline simple cycle system) and the multi-stage turbomachinery.

With these activities as a next logical step, a path forward is possible which can lead to the availability of very high efficiency power plants in the near future.

***High Efficiency Fossil Power Plants (HEFPP)
Conceptualization Program***

DE-RA26-97FT34164

**Final Report
Volume II**

**DESCRIPTION AND USER'S MANUAL OF
SYSTEM SIMULATION BASED ON EXCEL^(TM)
SPREADSHEET**

March, 1999

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Preface

Volume II of the HEFPP final report provides a description of the EXCEL^(TM) model which was developed for the selected powerplant design, and gives instructions for its use. The basic system design is reviewed, along with some of the fundamental math and logic which is used in the model.

In describing the model and procedures for using it, example printouts are provided in the main body and in appendices. Some of these may look slightly different from the output from the version of the model distributed with the final report. For example, late changes in the model code were made to refine the modeling of the turbomachinery, based on feedback from Allison Engine Company. This includes changing the compressor and expander efficiencies and adding inlet and ducting losses specified by Allison.

These types of changes will be noticed by the model user, but they should not be a problem in interpreting the model or interpreting this volume of the report.

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1.0 Introduction

This report provides a description of the simulation model for the 20 MW high efficiency power plant system devised by Energy Research Corporation under US Department of Energy contract DE-AC26-98FT34164, High Efficiency Fossil Power Plants (HEFPP) Conceptual Program. The simulation model is based on an [EXCEL 5 spread sheet](#) and provided on a 3½ disk.

2.0 Model Description

The following sections describe the high efficiency power plant system and the simulation model, which quantifies the performance characteristics of the system.

2.1 Power plant system

The high Efficiency System is shown in Figure 1. The system includes a direct carbonate fuel cell, a gas turbine with three compressor stages and two turbine stages, an oxidizer and twelve heat exchangers. The Fuel cells are operated at near atmospheric pressure and the gas turbine has a nominal pressure ratio of 15:1. In the system natural gas is treated to remove sulfur and preheated in E-6 by system exhaust. The preheated fuel is mixed with steam from boiler E-5 and recycled processed gas, heated further in E-3, and flows to an adiabatic prereformer. In the prereformer any heavy hydrocarbons are reformed. The fuel/steam mix is heated further in E-2 and flows to the direct carbonate fuel cell stacks. In the fuel cell stacks the stream is partially reformed in special reformer units, RUs. The partially reformed fuel stream then flows to the fuel cell anodes where the methane is completely converted and the hydrogen is electrochemically converted to DC power. Heat for the endothermic reform reaction in the RUs and anodes is provided by the excess heat generated in the fuel cell reaction. The anode vent stream contains unused hydrogen, CO and CO₂, from the incoming fuel as well as product water formed in the cells and additional CO₂ transferred from the cathodes. The anode vent stream flows to a catalytic oxidizer, or burner, where the unreacted hydrogen and CO are burned with process air from the turbine exhaust. The burner exhaust flows through heat exchangers E-8 and E-9 to the cathodes where oxygen and CO₂ react, completing the electrochemical carbonate cell process. Cathode exhaust flows through heat exchangers E-1, 2,3,4,5,6, and 7 and is exhausted from the system. Process air is compressed in the three gas turbine compressor stages. Intercooler E-10 and E-11 is provided to minimize compressor work. The intercoolers are cooled by a circulating water loop that rejects heat to the atmosphere in a natural draft cooling tower.

Figure 1 Process Flow Diagram for 20MW High Efficiency Power Plant

The cathode exhaust in heat exchangers E-4 and E-1 heats the compressed process air. It then flows through E-8 where it is heated further and expands through the first turbine stage. The exhaust from the first turbine stage is reheated in E-9 and flows to the second stage turbine. Exhaust air from the second stage turbine flows to the burner.

2.2 General model description

The power plant system model has the EXCEL work book file name "hefplant.xls". The model establishes power output, efficiency, key process performance parameters, process flows, gas compositions, temperatures, pressures, and mass and energy balances throughout the power plant system. The model is programmed with certain power plant design assumptions which are presented as reference values. The sensitivity to each of the key reference design assumptions can be evaluated by an input parameter deviation. The model iterates the flows and energies in the system based on the parameter deviation until there is energy, atom and mass balance convergence throughout the system. The resulting impact on the power plant performance and conditions throughout the system is displayed.

The model consists of 8 linked files or sheets bound in work book file "hefplant.xls". The 8 sheets are listed in Table 1 and are displayed as tabs along the bottom of the active work book. A print out of each of the 8 sheets is presented in the appendices. The following are descriptions of each sheet presented in order as listed in Table 1. Following the description of each of the sheets is a section describing the procedures for using the EXCEL simulation model.

Table 1
Sheets included in EXCEL work book hefplant.xls

Summary
Boiler
Prereformer
Cell stack
Compressor
Air preheater
Turbine
Burner

2.3 Summary sheet

The summary sheet in the model, shown in Appendix 1 has four parts. The first part shown on the left of page 1 is a listing of the input parameters to the model, with a reference value for each parameter. The list starts at the top with the methane flow to the power plant and fuel heating value, and proceeds downward to include key parameters for the environment, the fuel cell stack, fuel processing, the gas turbine and the heat exchangers.

For certain of the heat exchangers the key parameter is the process cold side temperature, for others the parameter is heat exchanger effectiveness.

The second part of the summary sheet, shown on the right side of page one, is a listing of the output characteristics of the power plant. The list starts at the top with the net AC output power in kilowatts and proceeds downward displaying certain key characteristics for the fuel cells, and the gas turbine. Also listed are the heat exchanger effectiveness values needed to achieve the cold side output temperatures listed in the input parameter list.

The third part of the summary sheet shown on page 2 & 3 is a table listing the process conditions at each of the stations in the power plant systems as identified in Figure 1. The process parameters include the temperature, enthalpy, stream composition, the mole flow in lb-moles/hr and the mass flow in lb/hr.

The last part of the summary sheet shown on page 3 is a heat, mass and mole balance around the entire system displaying the input, output, error and error percentage. Since heat, mole and mass balances are performed in the modeling of each process in the system the overall is very accurate. There is a slight, <1000 Btu/hr out of about 20 million Btu/hr, imbalance of energy resulting from the convergence logic at each process step as described later.

Only the summary sheet is used to evaluate the effect of changing a system input parameter. The other sheets are all linked to the summary sheet and to each other. The stream characteristics, flow rate, composition and temperature, developed in one sheet are linked to the other sheets which use those characteristics. In the following file descriptions, parameters that can be varied in the model are presented in *Italics*.

2.4 Boiler sheet

The boiler file, shown in appendix 2, includes the boiler E-5, the fuel preheater E-6 and the feedwater preheater E-7. The boiler hot side exit temperature is calculated based on the assigned *steam temperature*, and an energy balance.

The fuel preheater hot side exit temperature is calculated based on the assigned values for the *cold side inlet and outlet temperatures*. The feedwater hot side exit temperature is calculated based on the assigned values of the *cold side inlet and outlet temperatures*.

2.5 Prereformer sheet

The prereformer file, shown in appendix 3, starts with the preheated fuel, steam and recycled processed gas and calculates the mixed conditions at the entrance to the prereformer inlet heater E-3. The file calculates E-3 hot side exit temperature based on the hot side inlet conditions from E-2, and the assigned *cold side exit temperature*. The file then takes the stream conditions to the prereformer and calculates the process conditions and temperature leaving the adiabatic prereformer.

This is accomplished as follows. The file uses the following two equations for reforming and shift conversion to establish the exit composition.

Reforming Reaction “x”

$$\text{Where the reforming equilibrium constant } K_{pr} = \frac{(\text{PH}_2)^3 \times \text{PCO}}{\text{PCH}_4 \times \text{PH}_2\text{O}}$$

Shift Conversion “y”

$$\text{Where the shift equilibrium constant } K_{ps} = \frac{\text{PH}_2 \times \text{PCO}_2}{\text{PH}_2\text{O} \times \text{PCO}}$$

The combined reform and shift equations are solved simultaneously as follows. A CH₄ conversion fraction "x" is assumed. Then the shift fraction "y" is calculated from the above equations. Based on the calculated "y" and assumed "x", a composition at the exit of the prereformer is calculated. From this composition the reform constant is calculated and compared to the actual K_{pr} calculated at the prereformer exit temperature. The file then performs a dual iteration changing the assumed reform conversion fraction "x" each iteration until the proper K_{pr} is achieved within the preset convergence criteria.

At the same time the file iterates the exit temperature to achieve an energy balance. The iteration of reform fraction and exit temperature proceed simultaneously until both energy balance and equilibrium constant match are satisfied.

The prereformer file also includes models for the inlet preheater, E-3, and the exit heater E-2. In each heat exchanger model the hot side exit temperatures are calculated to achieve the assigned *cold side exit temperatures*.

2.6 Cell stack sheet

The cell stack file, shown in appendix 4, starts with the gas stream to the internal reforming units, RUs, within the cell stack. The file calculates an exit equilibrium composition from the RUs as previously discussed for the prereformer. The file calculates a shift and reform equilibrium constant based on the assigned *prereformer exit temperature*. The file assumes a reform conversion fraction and calculates a shift equilibrium composition based on the assumed conversion fraction. Next the file calculates a reform equilibrium constant based on the composition and compares it to the calculated constant based on the temperature. The file then iterates the assumed reform fraction until the calculated reform constant based on composition equals the reform constant based on temperature. An energy balance is then provided determining the endothermic reaction energy required by the internal reformer units.

The next step in the cell stack file is the calculation of the DC power output, hydrogen consumption, oxygen consumption and CO₂ transfer through the cells based on the assigned *cell voltage* and *current density* and the number and size of the cells. The file then takes the anode inlet stream and calculates a non equilibrium composition based on the hydrogen consumption, the water production and the CO₂ transfer within the cells. It also assumes 100% conversion of the methane in the anodes. Based on an assigned *anode exit temperature* the file calculates a shift equilibrium composition leaving the anodes. Next the file takes the cathode inlet composition, removes the oxygen consumed, and the CO₂ transferred to the anode, and calculates the cathode exit stream composition. The final step is an energy balance on the cell stack which includes the DC power output, the *stack heat loss*, the energy absorbed in the RU endothermic reaction, and the energy in the anode and cathode streams. The energy balance is used to determine the cathode exit temperature.

2.7 Compressor sheet

The compressor file shown in appendix 5 includes the models for each of the three compressor stages, the intercoolers the intercooler cooling loop pump and the natural draft cooling tower. It starts with certain parameter definitions from the summary sheet such as *compression ratio*.

The file then takes the inlet air to the first stage compressor, calculates the energy requirement for adiabatic compression and calculates the exit temperature based on an energy balance. The file then calculates the hot side and cold side exit temperatures for the first stage intercooler based on the assumed *effectiveness*. Next the file calculates the input pressure to the second stage compressor based on *intercooler pressure loss*, calculates the second stage compressor adiabatic energy requirement and calculates exit temperature by an energy balance as for the first stage. This is repeated for the second intercooler and third stage compressor. The file then determines the mix temperature of the water coolant streams from the two intercoolers and determines the hot side and cold side exit temperatures from the cooling tower based on the assigned *effectiveness*.

The last part of the compressor file calculates the parasitic power for the water pump in the intercooler cooling loop, and the overall parasitic power based on an assumed value of *parasitic power for the controls and miscellaneous* from the summary file.

2.8 Air preheater sheet

The air preheater file defines the conditions around the two air preheaters E-4 and E-1. For each heat exchanger it calculates the hot side exit temperature based on the established *cold side exit temperatures* and calculates the associated heat exchanger effectiveness.

2.9 Turbine sheet

The turbine file shown in appendix 7 includes the models for each of the two turbine stages. It starts with *turbine efficiencies* from the summary sheet. The file calculates pressure to the first turbine stage based on pressure from the compressor third stage and assumed *pressure losses through heat exchangers E-4, E-1 and E-8* from the summary sheet. It calculates the inlet pressure to the second stage turbine based on the loss through E-9. Pressure at the exit of the second turbine stage is established by pressure losses in the cell stack and the burner as well as hot side losses through heat exchangers E-8, E-9, and E-1 through E-7.

The file takes the stream characteristics to the first stage turbine, calculates the adiabatic expansion energy output in the first turbine stage, and calculates the exit temperature based on an energy balance. Next the file takes the input stream characteristics to the second turbine, calculates the second stage turbine adiabatic

energy output and calculates exit temperature by an energy balance similar to the first stage.

The file then summarizes the energy requirements for the compressor stages and the turbine stages, calculates the net energy and applies the *generator efficiency* to establish the AC output of the turbine generator.

2.10 Burner sheet

The burner sheet file, shown in appendix 8, starts with the stack anode exit stream and the stream from the exit of the second turbine stage. The mix stream temperature is calculated from an energy balance. The file then calculates the burner exhaust conditions assuming all of the hydrogen and CO are consumed. Exit temperature is calculated from an energy balance. The file establishes the exit temperature from heat exchanger E-8 based on a reference *E-8 cold side exit temperature* from the summary sheet. The file also calculates the required E-8 heat exchanger effectiveness. In a similar manner the file then establishes the exit temperature from heat exchanger E-9 based on a reference *E-9 cold side exit temperature* from the summary sheet. The file also calculates the required E-9 heat exchanger effectiveness.

3.0 Procedures for Model Use

3.1 Loading the program and initial EXCEL 5 settings

The following steps are recommended for use of the High Efficiency System EXCEL model.

1. Open EXCEL 5 work book file hefplant.xls provided on 3½ disk.
2. Before changing any parameter: click TOOLS, then OPTIONS
3. Under OPTIONS VIEW tab check
 - Show formula bar
 - Show status bar
 - Comments-none
 - Objects show all
4. Under WINDOWS OPTIONS tab check
 - Gridlines
 - Row and column headings
 - Outline symbols
 - Zero values
 - Horizontal scroll bar
 - Vertical scroll bar

Sheet tabs

Color automatic

5. Under the CALCULATION tab check

Manual

Put in 10,000 iterations

Maximum change 0.001

“No check mark” for recalculate before saving

Update remote references

Save external link values

3.2 Parameter sensitivity studies

1. Assign a new case number to the top slot on the summary sheet. This case number will be included on all 3 pages of the summary.

2. To evaluate an input parameter deviation: Put the + or minus deviation value in the deviation column adjacent the reference value. For example, to show the impact of increasing the turbine efficiency to 90% and the compressor efficiency to 85% as projected by Allison, put a deviation of 2 adjacent the compressor efficiency and a deviation of 7 adjacent the turbine efficiency. The result is shown in Appendix 8. An air flow deviation of –330 has been included to maintain the cell stack cathode exhaust temperature within established limits. Any number of deviations can be placed at one time, and deviations can remain from one case to another.

3. Press the F9 function key to start the calculation. The calculation will iterate until all the composition and energy balances are satisfied and the iteration stops. The new characteristics for the system will be displayed on the summary sheet.

iterate until all of the composition and energy balances are satisfied. This may take 1000 to 10,000 iterations depending on the parameter deviation input. Repeat as necessary until the parameters are steady and the iteration stops.

4. Read and print the outputs at the new input conditions in the summary file.

3.3 Cautions

1. When returning to the reference value enter the deviation as “zero”. If the cell is entered as blank the program will turn all parameters to “#VALUE” from which there the recovery is very tedious.

2. If the deviation is excessive or the deviation is not returned as a zero. The file may become unstable and all numbers will appear as "#VALUE". If this occurs *close the hefplant.xlw file without saving.* Reopen the file and reduce the magnitude of the input deviation or insure the reference value is returned by putting a zero in the deviation cell.

4.0 Description of key technical models

4.1 Specific Enthalpies of Gases

The enthalpy of for each of the gases is based on a 5th order polynomial fit from reference 6, providing excellent agreement with values from reference 2. The specific enthalpy for each of the gases is in Btu/lb mole and has an enthalpy base of 0 °K. Coefficients for the 5th order polynomial equations are listed in Table 1.

TABLE 1
POLYNOMIAL ENTHALPY CORRELATION,
Enthalpy, Btu/ lb mole = $a + bT + cT^2 + dT^3 + eT^4 + fT^5$
Where T is in degrees fahrenheit

Gas	a	b	c	d	e	f
O2	3192	6.88	0.0005923	1.452E-07	-1.38E-10	2.596E-14
H2	3112	6.888	0.0002101	-2.164E-07	1.376E-10	-2.452E-14
N2	3193	6.984	-0.0001687	5.128E-07	-1.948E-10	2.489E-14
H2O	-99150	7.959	0.0003182	3.576E-07	-1.074E-10	1.048E-14
CO	-45770	6.966	-9.373E-05	5.408E-07	-2.314E-10	3.306E-14
CO2	-165800	8.425	0.003167	-1.093E-06	2.497E-10	-2.7E-14
CH4	-25080	7.806	0.003597	5.746E-07	-4.709E-10	7.48E-14

The equation for the enthalpy of liquid water at temperature “t” is:

$$H \text{ Btu/lb mole} = -114,844 - 18.016 \times (212 - t)$$

4.2 Reforming and shift equilibrium

Where initial mole flows or composition fractions are:

$$\begin{array}{ll} \text{H}_2 = A & \text{CO} = D \\ \text{H}_2\text{O} = B & \text{CO}_2 = E \\ \text{CH}_4 = C & x = \text{The CH}_4 \text{ converted} \end{array}$$

The equation for the reforming equilibrium constant is:

$$K_{pr} = \frac{(P_{\text{H}_2})^3 \times P_{\text{CO}}}{P_{\text{CH}_4} \times P_{\text{H}_2\text{O}}}$$

The reforming equilibrium constant is calculated from:

$$\text{Log}_{10} K_{pr} = 13.50732392 - 22297.20008 (V) + 972312.7798(V)^2$$

$$\text{Where } V = 1 / (T + 460)$$

$$T = \text{Exit temperature} - \Delta T \text{ (approach to equilibrium)}$$

$$\begin{aligned} \text{The total moles} &= (C-x) + (B-x+y) + (3x+A-y) + (D+x+y) + (E-y) \\ &= A+B+C+D+E+2x \end{aligned}$$

The 2x term reflects the fact that 2 extra moles are produced for each mole of CH₄ Which is converted to H₂. The partial pressure of a constituent is therefore:

$$P_{\text{H}_2} = \frac{P_t (3x + A - y)}{(A+B+C+D+E+2x)} \quad \text{example}$$

The equation for K_{pr} is therefore:

$$K_{pr} = \frac{(3x + A - y)^3 (D + x + y)}{(C - x) (B - x + y)} \times \frac{(P_t)^2}{(A + B + C + D + E + 2x)^2}$$

where P_t = total pressure, atms

Shift Conversion

Where the shift equilibrium constant $K_{ps} = \frac{P_{H_2} \times P_{CO_2}}{P_{H_2O} \times P_{CO}}$

$$K_{ps} = \frac{(3x + A - y)(E - y)}{(B - x + y)(D + x + y)}$$

Solving this equation for y results in a quadratic equation

$$y = \frac{b^2 \pm (b^2 - 4ac)^{1/2}}{2a}$$

where $a = K_{ps} - 1$

$$b = K_{ps}(B + D) + E + A + 3x$$

$$c = K_{ps}(BD + Bx - Dx - x^2) - 3Ex - AE$$

The shift equilibrium constant as function of temperature is:

$$\text{Log}_{10}K_{ps} = -1.460864117 + 2492.498443(V) - 696866.2019(V)^2$$

$$V = 1 / (T + 460)$$

T = actual exit temperature, °F

The reform and shift equations are solved for x and y in the following way.

A CH₄ conversion fraction "x" is assumed. Then the shift fraction "y" is calculated from the quadratic shift equation. Based on the calculated "y" and assumed "x", a composition at the exit of the SHR is calculated. From this composition the reform constant is calculated and compared to the actual K_{pr} calculated at the SHR exit temperature. The model file then performs a dual iteration changing the assumed reform conversion fraction "x" until the proper K_{pr} is achieved, and adjusting the exit temperature to achieve an enthalpy balance.

4.3 Compressor

The compressor model for each compressor stage is based on an adiabatic compression of the inlet gas stream. The ideal energy for the adiabatic compression from Reference 4 is:

$$W_{ci} = wRT_{ci}(k/k-1)(P_o/P_i)^{((k/k-1)-1)}$$

Where W_{ci} = Ideal compressor work in Btu/hr
 w = Flow, lb moles /hr
 $R = 1544/778.3 = 1.984$ Btu/lb mole °R
 T_{ci} = Inlet temperature in °R
 k = Compressibility constant for the gas stream = c_p/c_v
 P_o/P_i = Compressor pressure ratio

The actual work is determined by applying an adiabatic efficiency to the process

$$W_{c(actual)} = W_{ci}/\text{efficiency}$$

An energy balance around the compressor stage accounting for the heat loss determines the exit temperature from the compressor.

4.4 Turbine

The model for each turbine stage is based on an adiabatic expansion of the inlet gas stream. The ideal energy for the adiabatic expansion from Reference 4 is:

$$W_{ti} = wRT_{ti}(k/k-1)(P_o/P_i)^{((k/k-1)-1)}$$

Where W_{ti} = Ideal turbine work in Btu/hr
 w = Flow, lb moles /hr
 $R = 1544/778.3 = 1.984$ Btu/lb mole °R
 T_{ti} = Inlet temperature in °R
 k = Compressibility constant for the gas stream = c_p/c_v
 P_o/P_i = Turbine expansion pressure ratio

The actual work output is determined by applying an adiabatic efficiency to the process

$$W_{t(actual)} = W_{ti} \times \text{efficiency}$$

An energy balance around the turbine stage accounting for the heat loss determines the exit temperature from the turbine.

An electrical efficiency for the turbine generator is applied to convert from rotational energy of the turbine to generator electrical output.

4.5 Burner

The method for determining the conditions at the exit of a burner is as follows. The composition of the streams entering a burner is determined by adding the flows. The temperature of the mixed stream before burning is determined by a temperature enthalpy iteration in which the mix temperature is changed until the enthalpy of the mix stream is balanced with the enthalpy of the combined inlet streams. After the mix temperature is determined, the conditions at the exit of the burner are determined as follows. The basic reactions for gases in a burner are:

From the above, the following equations are used to determine the composition of the gas at the exit of a burner where CO, CH₄ and H₂ conversion=100%.

1. $\text{CO}_{2\text{out}} = \text{CO}_{2\text{in}} + \text{CO}_{\text{in}} + \text{CH}_{4\text{in}}$
2. $\text{O}_{2\text{out}} = \text{O}_{2\text{in}} - \frac{1}{2}\text{CO}_{\text{in}} - 2\text{CH}_{4\text{in}} - \frac{1}{2}\text{H}_{2\text{in}}$
3. $\text{H}_2\text{O}_{\text{out}} = \text{H}_2\text{O}_{\text{in}} + \text{H}_{2\text{in}} + 2\text{CH}_{4\text{in}}$
4. $\text{CH}_{4\text{out}} = 0$
5. $\text{H}_{2\text{out}} = 0$
6. $\text{CO}_{\text{out}} = 0$
7. $\text{N}_{2\text{out}} = \text{N}_{2\text{in}}$

The temperature rise in the burner is determined by a temperature enthalpy iteration in which the exit temperature is changed until the enthalpy of the exit stream is balanced with the mix enthalpy before burning.

4.6 Heat exchangers

Heat exchanger performance is based on effectiveness “e” as defined in reference 1.

$$\text{where } e = \frac{WC_{ph}(\text{Thi} - \text{Tho})}{W_{cp\min}(\text{Thi} - \text{Tci})} \\ = \frac{WC_{pc}(\text{Tco} - \text{Tci})}{W_{cp\min}(\text{Thi} - \text{Tci})}$$

In the heat exchanger model $C_p = (\text{Hin} - \text{Hout})/(\text{Tin} - \text{Tout})$

Where H is the stream enthalpy in BTU/hr and T is the temperature in °F

4.7 Pressure losses

Pressure losses through the cell stack are representative of measured losses from operating cell stacks. Pressure losses through the burner, prereformer and heat exchangers are allocations for design based on previous more detailed design and construction experience. In some areas of the system the impact of pressure losses can be explored as deviations to a reference pressure loss.

4.8 Heat losses

The model includes provision at each energy balance point to include heat loss. Reference levels of heat loss are assigned for the cell stacks, based on an understanding of packaged cell stack designs, and for the gas turbine. Heat losses from other equipment are considered to be negligible relative to the magnitude of the stream energies.

4.9 Iteration technique for energy balances

The key iteration scheme used in the model is a mass and energy balance around a component. The exit temperature from the component is adjusted until an energy balance is established. This is accomplished using a logic scheme as shown in the following example from an EXCEL sheet.

	A	B	C	D	E	F	G	H	I	J	K
1	ENTHALPY IN, BTU/HR				-1.9800E+08			HEAT BALANCE			
2	ENTHALPY OUT, BTU/HR				-1.9800E+08			LOGIC FOR ITERAT			FALSE
3	HEAT LOSS, BTU/HR				0			ITERATION INCREM			0
4	NET ENTHALPY, IN-OUT, BTU/HR				400			TEMP CHANGE =			0

In cell E4 the net enthalpy is calculated from the in and out values.

Cell J2 has the logic: =AND (SIGN(E4)=1,E4>500), If E4 is greater than +500 the cell reads TRUE, if not, it reads FALSE.

Cell K2 has the logic: =AND(SIGN(E4)=-1,ABS(E4)>500), If E4 is negative and has an absolute value greater than 500, the cell reads TRUE, if not, it reads FALSE.

Cell J3 has the logic: =IF(J2=TRUE,J3+.01,J3), If J2 is true the value .01 is added to J3. If J2 is false J3 stays the same value.

Cell K3 has the logic: =IF(K2=TRUE,K3-.01,K3), If K2 is true the value .01 is subtracted from K3. If K2 is false K3 stays the same value.

Cell J4 has the logic: =J3+K3, This changing value is added to the temperature of the exit stream which changes its enthalpy.

The iteration proceeds with the temperature changing by 0.01°F until the net enthalpy is within 500 Btu/hr.

4.10 Cell stack

An MCFC has the following reactions at the electrodes.

The hydrogen consumed is based on Faraday's law as follows:

$$\text{WH}_2 = \frac{1.008 \text{ gm/gm eq} \times \text{lb} \times 3600 \text{ sec} \times 1000 \text{ watts} \times 1 \times 1}{96,500 \text{ amp sec/ gm eq} \times 453.59 \text{ gm hr} \times \text{kilowatts} \times \text{volts/cell} \times 2.016 \text{ lb/lbmole}}$$

$$\text{WH}_2, \text{ lb moles /kW-hr} = \frac{0.04112}{\text{volts / cell}}$$

The energy (Heating Value) available follows the reaction

From reference 2 @ 1200 °F

$$11,530 + \frac{1}{2}(12,332) = -88714 + \text{Energy}$$

$$\text{Heating value Energy(HV)} = 106,410 \text{ BTU/ Lb mole H}_2 \text{ consumed}$$

The waste heat removed by the circulating anode and cathode gas streams

$$\text{Waste heat, Btu/hr} = \text{HV} \times \text{WH}_2 - 3413 \times \text{DC Power out, kW} - \text{heat loss}$$

The waste heat is removed by the endothermic reaction heat in the RUs and anodes, and by the enthalpy change of the anode and cathode streams

5. REFERENCES

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2. Girdler Catalyst Document 1954, Physical and Thermodynamic Properties of various Elements and Compounds and other useful Information.
3. DOE/METC-94 /1006 Fuel Cells A Handbook (Revision 4) November 1998
4. Elementary Thermodynamics by Virgil Moring Faires
5. Microsoft EXCEL Version 5 Users Guide
6. DOE report on high efficiency system

APPENDICES

hefplant.xlw sheets

1. Summary sheet
2. Boiler sheet
3. Prereformer sheet
4. Cell stack sheet
5. Compressor sheet
6. Air preheater sheet
7. Turbine sheet
8. Burner
9. Summary sheet for example case

APPENDIX 1

SUMMARY SHEET

hefplant.xls

SUMMARY

3/29/99 15:29		HIGH EFFICIENCY POWER PLANT PROCESS SUMMARY									
EXCEL 97 WORK BOOK PREPARED BY R.A.Sanderson P.E.											
REVISION-- MARCH 19,1999											
		FIXED		CASE		1					
INPUT PARAMETERS		REFERENCE	+/- DEVIATION	OUTPUT CHARACTERISTICS						REQUIREMENT	
CH4 INPUT, MOLES/HR	=	296.00	0	NET AC POWER, KW						= 20,411 20,000.00	
PROCESS AIR FLOW, MOLES/HR	=	4570	0	NET AC LHV EFFICIENCY,%						= 71.08 > 70	
AMBIENT AIR TEMP F	=	59	0	PARASITIC LOSSES, KW						= 31	
WATER VAPOR PRESS, PSIA	=	0.147	0	GROSS AC POWER						= 20442	
FUEL LOWER HEATING VALUE,BTU/SCF	=	872.5	N/A	PLANT EXHAUST TEMP, F						= 208.6	
CELL STACK				FUEL CELLS							
CELL VOLTAGE	=	0.817	0	FUEL CELL AC OUTPUT, KW						= 17002	
CURRENT DENSITY, AMPS/CM^2	=	0.16	0	FUEL CELL DC POWER, KW						= 17619	
NUMBER STACKS		48	N/A	FUEL UTILIZATION,%						= 78.0	
NUMBER CELLS/STACK	=	360	N/A	CATHODE EXIT TEMP, DEG F						= 1173.6 1180 MAX	
CELL AREA, M^2	=	0.78	N/A	CATHODE INLET TEMP, DEG F						= 1035.8 1030 MIN	
ANODE EXIT TEMP DEG C	=	1170	0	FUEL CELL OXYGEN UTILIZATION, %						= 54.0	
METHANE CONVERSION,%	=	100	N/A	FUEL CELL INLET O2 FRACTION						= 0.1236	
STACK HEAT LOSS, MBTU/HR	=	700,000	0	OUTLET O2 FRACTION						= 0.0711	
INVERTER EFFICIENCY,%	=	96.5	0	FUEL CELL INLET CO2 FRACTION						= 0.1793	
PARASITIC PWR, CONTROLS&MISC,KW	=	10	0	OUTLET CO2 FRACTION						= 0.0572	
FUEL PROCESSING	=			GAS TURBINE							
PREREFORM EXIT PRES, PSIA	=	17	0	TURBINE GEN NETPOWER AC, KW						= 3440	
PREREFORM INLET TEMP, C	=	752	E-3 EXIT	HI TURBINE INLET TEMPERATURE, F						= 1526	
STEAM/CH4 RATIO	=	1.5	0	LOW TURBINE INLET TEMPERATURE, F						= 1220	
RU TEMP, F	=	1100	0	HIGH TURBINE POWER, KW						= 8035	
GAS TURBINE			0	LOW TURBINE POWER, KW						= 1315	
GAS TURBINE COMP EFF, %	=	85	0	1ST STAGE COMPRESSOR WORK, KW						= 1827	
GAS TURBINE TURB EFF, %	=	90	0	2ND STAGE COMPRESSOR WORK, KW						= 2002	
GT COMPRESSOR PRESS RATIOS	=	2.63	0	3RD STAGE COMPRESSOR WORK,KW						= 2011	
EACH INTERCOOLER PRESS LOSS, PSI	=	1	0								
COMPRESSOR DUCT AND INLET LOSSES, %	=	3	0								
PRESSURE LOSS, COMP OUT TO TURB IN,PSI	=	4	0	PROCESS WATER TOTAL, GAL/MIN						= 16	
PRESSURE LOSS, E-9 ,PSI	=	1	0								
IST TURBINE STAGE OUTLET PRESSURE, PSIA	=	23	0								
TURBINE EXIT DIFFUSER LOSSES, %, 1ST/2N	=	7/5	0								
GENERATOR EFFICIENCY,%	=	98	0								
TURBINE STAGE HEAT LOSS, BTU/HR	=	10000	0								
HEAT EXCHANGER SUMMARY		REFERENCE DEG F	+/- DEVIATION	REQUIRED HEAT EXCHANGER EFFECTIVENESS							
E-1 SECOND STAGE AIR HEATER	COLD SIDE EXIT	1058	0	E-1	=	0.68					
E-2 PROCESSED FUEL REHEATER	COLD SIDE EXIT	932	0	E-2	=	0.86					
E-3 FUEL /STEAM HEATER	COLD SIDE EXIT	752	0	E-3	=	0.75					
E-4 FIRST STAGE AIR HEATER	COLD SIDE EXIT	816	0	E-4	=	0.96					
E-5 BOILER	COLD SIDE EXIT	239	0	E-5	=	0.81					
E-6 NATURAL GAS HEATER	COLD SIDE EXIT	212	0	E-6	=	0.81					
E-7 FEEDWATER HEATER	COLD SIDE EXIT	203	0	E-7	=	0.80					
E-8 TURBINE INLET HEATER	COLD SIDE EXIT	1526	0	E-8	=	0.89					
E-9 TURBINE INLET REHEATER	COLD SIDE EXIT	1220	0	E-9	=	0.84					
E-10 INTERCOOLER	HOT SIDE EXIT	122	0	E-10	=	0.79					
E-11 INTERCOOLER	HOT SIDE EXIT	122	0	E-11	=	0.86					
E-12 COOLING TOWER	COLD SIDE EXIT	86	0	E-12	=	0.81					

SUMMARY

3/29/99 15:29				HIGH EFFICIENCY POWER PLANT PROCESS SUMMARY									
						PROCESS CONDITIONS							
					CASE	1							
	STATION	TEMP	PRESS	ENTHALPY	COMPOSITION, VOLUME FRACTION							FLOW	FLOW
NO	DESCRIPTION	DEG F	PSIA	BTU/HR	H2	H2O	CH4	CO	CO2	O2	N2	MOLES/HR	LBS/HR
1	NATURAL GAS IN	59.0	29.70	-7,949,462	0.0000	0.0000	0.9600	0.0000	0.0200	0.0000	0.0200	296.00	4,985
2	DESULFUR NAT GAS	15.0	24.70	-7,949,462	0.0000	0.0000	0.9600	0.0000	0.0200	0.0000	0.0200	296.00	4,985
3	HEATED NAT GAS	212.0	24.60	-7,551,780	0.0000	0.0000	0.9600	0.0000	0.0200	0.0000	0.0200	296.00	4,985
4	FUEL/STEAM/RECYCLE	233.7	19.20	-51,454,583	0.0012	0.5991	0.3835	0.0000	0.0083	0.0000	0.0080	752.02	13,187
5	PREREFORMER INLET	752.0	19.00	-47,683,213	0.0012	0.5991	0.3835	0.0000	0.0083	0.0000	0.0080	752.02	13,187
6	PREREFORMER EXIT	611.3	17.00	-47,682,947	0.0748	0.5448	0.3539	0.0001	0.0264	0.0000	0.0078	774.58	13,187
7	FUEL/STEAM TO E-2	611.3	16.71	-46,948,630	0.0742	0.5406	0.3511	0.0001	0.0262	0.0000	0.0077	768.57	12,984
8	FUEL/STEAM FROM E-2	932.0	16.61	-44,362,077	0.0742	0.5406	0.3511	0.0001	0.0262	0.0000	0.0077	768.57	12,984
9	FUEL TO STACK Rus	932.0	16.60	-44,362,077	0.0746	0.5357	0.3555	0.0001	0.0264	0.0000	0.0078	768.57	12,984
10	FUEL FROM Rus	1100.0	15.60	-30,524,258	0.5111	0.2058	0.1256	0.0734	0.0784	0.0000	0.0057	1,045.61	12,984
11	FUEL TO ANODES	1100.0	15.60	-30,524,258	0.5111	0.2058	0.1256	0.0734	0.0784	0.0000	0.0057	1,045.61	12,984
12	FUEL FROM ANODES	1170.0	15.52	-243,537,416	0.0715	0.3897	0.0000	0.0423	0.4938	0.0000	0.0027	2,195.09	66,199
13	FUEL TO BURNER	1170.0	15.50	-243,537,416	0.0715	0.3897	0.0000	0.0423	0.4938	0.0000	0.0027	2,195.09	66,199
14	BURNER PREMIX	1124.5	15.45	-199,470,000	0.0232	0.1332	0.0000	0.0137	0.1623	0.1398	0.5278	6,765.09	197,763
15	BURNER EXHAUST	1585.3	15.35	-199,469,960	0.0000	0.1593	0.0000	0.0000	0.1793	0.1236	0.5377	6,640.15	197,763
16	EXHAUST FROM E-8	1309.4	15.27	-216,413,113	0.0000	0.1593	0.0000	0.0000	0.1793	0.1236	0.5377	6,640.15	197,763
17	CATHODE INLET	1035.8	15.19	-232,676,546	0.0000	0.1593	0.0000	0.0000	0.1793	0.1236	0.5377	6,640.15	197,763
18	CATHODE EXHAUST	1173.6	15.11	-94,334,827	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
19	EXHAUST TO E-1	1173.6	15.15	-94,334,827	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
20	EXHAUST TO E-2	982.1	15.05	-102,758,864	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
21	EXHAUST TO E-3	922.3	15.00	-105,345,417	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
22	EXHAUST TO E-4	834.1	14.95	-109,116,787	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
23	EXHAUST TO E-5	446.8	14.85	-125,129,423	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
24	EXHAUST TO E-6	248.1	14.80	-133,024,387	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
25	EXHAUST TO E-7	238.0	14.75	-133,422,069	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
26	EXHAUST FROM SYSTEM	208.6	14.70	-134,573,876	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
27	PROCESS AIR IN	59.0	14.70	9,478,849	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
28	1ST STAGE COMP IN	59.0	14.45	9,478,849	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
29	1ST STAGE COMP OUT	253.7	38.01	15,714,514	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
30	2ND STAGE COMP IN	122.0	35.87	11,490,654	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
31	2ND STAGE COMP OUT	334.3	94.35	18,322,460	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
32	3RD STAGE COMP IN	122.0	90.52	11,490,654	0.0000	0.0583	0.0000	0.0000	0.0091	0.6273	2.3636	4,570.00	131,564
33	3RD STAGE COMP OUT	335.4	238.06	18,355,459	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
34	PREHEATER E-4 OUT	816.0	236.73	34,368,095	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
35	PREHEATER E-1 OUT	1058.0	235.39	42,792,133	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
36	1ST TURBINE IN	1526.0	231.68	59,735,286	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
37	1ST TURBINE OUT	755.6	25.81	32,302,844	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
38	2ND TURBINE IN	1220.0	23.00	48,566,276	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
39	2ND TURBINE OUT	1094.0	15.50	44,067,163	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564

SUMMARY

3/29/99 15:29				HIGH EFFICIENCY POWER PLANT PROCESS SUMMARY										
						PROCESS CONDITIONS								
						CASE	1							
STATION		TEMP	PRESS	ENTHALPY	COMPOSITION, VOLUME FRACTION								FLOW	FLOW
NO	DESCRIPTION	DEG F	PSIA	BTU/HR	H2	H2O	CH4	CO	CO2	O2	N2	MOLES/HR	LBS/HR	
40	PROCESS WATER IN	59.0	30	-52,214,531	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	444.00	7,999	
41	TREATED WATER	59.0	24	-52,214,531	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	444.00	7,999	
42	FEEDWATER HTR OUT	203.0	23	-51,062,724	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	444.00	7,999	
43	STEAM FROM BOILER	239.0	22	-43,167,937	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	444.00	7,999	
44	STEAM TO EJECTOR	239.0	21.5	-43,167,937	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	444.00	7,999	
45	FUEL RECYCLE	611.3	17	-734,257	0.0742	0.5407	0.3512	0.0000	0.0262	0.0000	0.0077	12.02	203	
46	STEAM/ RECYCLE MIX	250.1	19.5	-43,902,319	0.0020	0.9879	0.0093	0.0000	0.0007	0.0000	0.0002	456.02	8,202	
47	COOLING PMP INLET	86.0	29	-608,992,883	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	5,200.00	93,683	
48	COOLING PMP EXIT	86.0	36	-608,992,883	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	5,200.00	93,683	
49	COOLANT TO E-10	86.0	36	-234,228,032	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	2,000.00	36,032	
50	COOLANT FROM E-10	203.2	34	-230,004,171	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	2,000.00	36,032	
51	COOLANT TO E-11	86.0	36	-374,764,851	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	3,200.00	57,651	
52	COOLANT FROM E-11	204.5	34	-367,933,045	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	3,200.00	57,651	
53	COOLANT TO E-12	204.0	34	-597,937,610	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	5,200.00	93,683	
54	AIR TO TOWER	59.0	14.7	248,897,563	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	119,999.88	3,454,636	
55	AIR FROM TOWER	72.2	14.7	259,952,836	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	119,999.88	3,454,636	

APPENDIX 2

BOILER SHEET

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BOILER

3/29/99 15:29			BOILER																	
			BOILER, FEEDWATER HEATER & FUEL PREHEATER																	
STATION	23	TEMP, F	446.8	230.4	DEG C															
HOT SIDE INLET																				
	GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL										
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	27.222										
MOLE FLOW, LB MOLES/HR			0.00	1058.02	0.00	0.00	303.79	377.67	3570.52	5309.99										
MOLE FRACTION			0.000	0.199	0	0.000	0.057	0.071	0.672	1.000										
SPEC ENTH, BTU/LB MOLE			6,217	-95,503	-20,840	-42,637	-161,492	6,392	6,318											
ENTHALPY, BTU/HR			0	-101,043,667	0	0	-49,059,051	2,414,093	22,559,203	-125,129,423										
MASS FLOW LB/HR			0	19,061	0	0	13,370	12,085	100,032	144,548										
STATION	24	TEMP, F	248.1	120.1	DEG C															
HOT SIDE EXIT																				
	GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL										
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	27.222										
MOLE FLOW, LB MOLES/HR			0.00	1058.02	0.00	0.00	303.79	377.67	3570.52	5309.99										
MOLE FRACTION			0.000	0.199	0	0.000	0.057	0.071	0.672	1.000										
SPEC ENTH, BTU/LB MOLE			4,831	-97,150	-22,915	-44,040	-163,530	4,937	4,923											
ENTHALPY, BTU/HR			0	-102,787,147	0	0	-49,678,406	1,864,661	17,576,505	-133,024,387										
MASS FLOW LB/HR			0	19,061	0	0	13,370	12,085	100,032	144,548										
STATION	42	TEMP, F	203.0	95.0	DEG C	51128820				-7,894,787										
COLD SIDE INLET																				
	GAS		H2	H2O(LIQUID)	CH4	CO	CO2	O2	N2	TOTAL										
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	18.016										
MOLE FLOW, LB MOLES/HR			0.000	444.000	0.000	0.000	0.000	0.000	0.000	444										
MOLE FRACTION			0.000	1.000	0.000	0.000	0.000	0.000	0.000	1.000										
SPEC ENTH, BTU/LB MOLE			4,517	-115,006	-23,343	-44,356	-163,968	4,614	4,608											
ENTHALPY, BTU/HR			0	-51,062,724	0	0	0	0	0	-51,062,724										
MASS FLOW LB/HR			0	7,999	0	0	0	0	0	7,999										
STATION	43	TEMP, F	239.0	115.0	DEG C															
COLD SIDE EXIT																				
	GAS		H2	H2O(VAPOR)	CH4	CO	CO2	O2	N2	TOTAL										
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	18.016										
MOLE FLOW, LB MOLES/HR			0.000	444.000	0.000	0.000	0.000	0.0000	0.000	444										
MOLE FRACTION			0.000	1.000	0.000	0.000	0.000	0.000	0.000	1.000										
SPEC ENTH, BTU/LB MOLE			4,768	-97,225	-23,003	-44,104	-163,620	4,872	4,859											
ENTHALPY, BTU/HR			0	-43,167,937	0	0	0	0	0	-43,167,937										
MASS FLOW LB/HR			0	7,999	0	0	0	0	0	7,999										
ENTHALPY IN, BTU/HR				-1.7619E+08			HEAT BALANCE													
ENTHALPY OUT, BTU/HR				-1.7619E+08			LOGIC FOR ITERATION				FALSE		FALSE							
HEAT LOSS, BTU/HR				0			ITERATION INCREMENT				803.86		-805.73							
NET ENTHALPY, IN - OUT, BTU/HR				177			24 TEMP CHANGE =				-1.87									
CALCULATED EFFECTIVENESS FOR BOILER BASEDON HOT SIDE					e =(Thi-Tho))/(Thi-Tci)				0.815											

BOILER

				FUEL PREHEATER E-6								
STATION	24	TEMP, F	248	120.1	DEG C			FUEL FLOW LBMOLES/HR =		296		
HOT SIDE INLET												
GAS		H2		H2O	CH4	CO	CO2	O2	N2	TOTAL		
MOLECULAR WEIGHT		2.016		18.016	16.042	28.01	44.01	32	28.016	27.222		
MOLE FLOW, LB MOLES/HR		0.00		1058.02	0.00	0.00	303.79	377.67	3570.52	5309.99		
MOLE FRACTION		0.000		0.199	0	0.000	0.057	0.071	0.672	1.000		
SPEC ENTH, BTU/LB MOLE		4,831		-97,150	-22,915	-44,040	-163,530	4,937	4,923			
ENTHALPY, BTU/HR		0		-102,787,147	0	0	-49,678,406	1,864,661	17,576,505	-133,024,387		
MASS FLOW LB/HR		0		19,061	0	0	13,370	12,085	100,032	144,548		
STATION	25	TEMP, F	238.0	114.4	DEG C							
HOT SIDE EXIT												
GAS		H2		H2O	CH4	CO	CO2	O2	N2	TOTAL		
MOLECULAR WEIGHT		2.016		18.016	16.042	28.01	44.01	32	28.016	27.222		
MOLE FLOW, LB MOLES/HR		0.00		1058.02	0.00	0.00	303.79	377.67	3570.52	5309.99		
MOLE FRACTION		0.000		0.199	0	0.000	0.057	0.071	0.672	1.000		
SPEC ENTH, BTU/LB MOLE		4,761		-97,233	-23,012	-44,111	-163,629	4,865	4,852			
ENTHALPY, BTU/HR		0		-102,874,715	0	0	-49,708,514	1,837,171	17,323,989	-133,422,069		
MASS FLOW LB/HR		0		19,061	0	0	13,370	12,085	100,032	144,548		
STATION	2	TEMP, F	59.0	15.0	DEG C							
COLD SIDE INLET (FUEL IN)												
GAS		H2		H2O	CH4	CO	CO2	O2	N2	TOTAL		
MOLECULAR WEIGHT		2.016		18.016	16.042	28.01	44.01	32	28.016	16.841		
MOLE FLOW, LB MOLES/HR		0.00		0.00	284.16	0.00	5.92	0.00	5.92	296.00		
MOLE FRACTION		0.000		0.000	0.960	0.000	0.020	0.000	0.020	1.000		
SPEC ENTH, BTU/LB MOLE		3,519		-98,679	-24,607	-45,359	-165,292	3,600	3,605			
ENTHALPY, BTU/HR		0		0	-6,992,272	0	-978,529	0	21,339	-7,949,462		
MASS FLOW LB/HR		0		0	4,558	0	261	0	166	4,985		
STATION	3	TEMP, F	212.0	100.0	DEG C							
COLD SIDE EXIT												
GAS		H2		H2O	CH4	CO	CO2	O2	N2	TOTAL		
MOLECULAR WEIGHT		2.016		18.016	16.042	28.01	44.01	32	28.016	16.841		
MOLE FLOW, LB MOLES/HR		0.00		0.00	284.16	0.00	5.92	0.00	5.92	296.00		
MOLE FRACTION		0.000		0.000	0.960	0.000	0.020	0.000	0.020	1.000		
SPEC ENTH, BTU/LB MOLE		4,580		-97,445	-23,259	-44,293	-163,881	4,678	4,671			
ENTHALPY, BTU/HR		0		0	-6,609,252	0	-970,178	0	27,650	-7,551,780		
MASS FLOW LB/HR		0		0	4,558	0	261	0	166	4,985		
		STATION	STATION		ASSUMED EFFECTIVENESS, e =			0.809				
Thi, F	=	24	248	25	CALCULATED Tho F =			238.0		=Thi-eCmin/Ch(Thi-Tci)		
Tci, F	=	2	59	3	SELECTED COLD SIDE EXIT			212.0				
Ch=DH/DT	=	BTU/HR F	39261		ENERGY BALANCE			IN	OUT	DELTA		
Cc=DH/DT	=	BTU/HR F	2599		HOT SIDE			-133,024,387	-133,422,069	397,682		
Cmin	=	BTU/HR F	2599		COLD SIDE			-7,949,462	-7,551,780	-397,682		
Thi-Tci, F	=		189		TOTAL			-140,973,849	-140,973,849	0		
Cmin/Cmax	=		0.07									

BOILER

			FEEDWATER HEATER E-7						WATER /FUEL RATIO		1.5				
STATION	25	TEMP, F	238	114.4	DEG C				PROCESS WATER FLOW, LB MOLES /HR		444				
HOT SIDE INLET															
GAS		H2		H2O		CH4		CO		CO2		O2	N2	TOTAL	
MOLECULAR WEIGHT		2.016		18.016		16.042		28.01		44.01		32		28.016	27.222
MOLE FLOW, LB MOLES/HR		0.00		1058.02		0.00		0.00		303.79		377.67		3570.52	5309.99
MOLE FRACTION		0.000		0.199		0		0.000		0.057		0.071		0.672	1.000
SPEC ENTH, BTU/LB MOLE		4,761		-97,233		-23,012		-44,111		-163,629		4,865		4,852	
ENTHALPY, BTU/HR		0		-102,874,715		0		0		-49,708,514		1,837,171		17,323,989	-133,422,069
MASS FLOW LB/HR		0		19,061		0		0		13,370		12,085		100,032	144,548
STATION	26	TEMP, F	208.6	98.1	DEG C	PLANT EXHAUST									
HOT SIDE EXIT															
GAS		H2		H2O		CH4		CO		CO2		O2	N2	TOTAL	
MOLECULAR WEIGHT		2.016		18.016		16.042		28.01		44.01		32		28.016	27.222
MOLE FLOW, LB MOLES/HR		0.00		1058.02		0.00		0.00		303.79		377.67		3570.52	5309.99
MOLE FRACTION		0.000		0.199		0		0.000		0.057		0.071		0.672	1.000
SPEC ENTH, BTU/LB MOLE		4,556		-97,473		-23,291		-44,316		-163,914		4,654		4,647	
ENTHALPY, BTU/HR		0		-103,128,198		0		0		-49,795,018		1,757,675		16,591,665	-134,573,876
MASS FLOW LB/HR		0		19,061		0		0		13,370		12,085		100,032	144,548
STATION	41	TEMP, F	59.0	15.0	DEG C										
COLD SIDE INLET															
GAS		H2		H2O(LIQUID)		CH4		CO		CO2		O2	N2	TOTAL	
MOLECULAR WEIGHT		2.016		18.016		16.042		28.01		44.01		32		28.016	18.016
MOLE FLOW, LB MOLES/HR		0.00		444.00		0.00		0.00		0.00		0.00		0.00	444.00
MOLE FRACTION		0.000		1.000		0.000		0.000		0.000		0.000		0.000	1.000
SPEC ENTH, BTU/LB MOLE		3,519		-117,600		-24,607		-45,359		-165,292		3,600		3,605	
ENTHALPY, BTU/HR		0		-52,214,531		0		0		0		0		0	-52,214,531
MASS FLOW LB/HR		0		7,999		0		0		0		0		0	7,999
STATION	42	TEMP, F	203.0	95.0	DEG C										
COLD SIDE EXIT															
GAS		H2		H2O(LIQUID)		CH4		CO		CO2		O2	N2	TOTAL	
MOLECULAR WEIGHT		2.016		18.016		16.042		28.01		44.01		32		28.016	18.016
MOLE FLOW, LB MOLES/HR		0.00		444.00		0.00		0.00		0.00		0.00		0.00	444.00
MOLE FRACTION		0.000		1.000		0.000		0.000		0.000		0.000		0.000	1.000
SPEC ENTH, BTU/LB MOLE		4,517		-115,006		-23,343		-44,356		-163,968		4,614		4,608	
ENTHALPY, BTU/HR		0		-51,062,724		0		0		0		0		0	-51,062,724
MASS FLOW LB/HR		0		7,999		0		0		0		0		0	7,999
Thi, F	=	STATION 25	238	STATION 26	ASSUMED EFFECTIVENESS, e =					0.804					
Tci, F	=	41	59	42	CALCULATED Tho F =					208.6		=Thi-eCmin/Ch(Thi-Tci)			
Ch=DH/DT	=	BTU/HR F	39182		SELECTED COLD SIDE EXIT					203.0					
Cc=DH/DT	=	BTU/HR F	7999		ENERGY BALANCE					IN		OUT		DELTA	
Cmin	=	BTU/HR F	7999		HOT SIDE					-133,422,069		-134,573,876		1,151,807	
Thi-Tci, F	=		179		COLD SIDE					-52,214,531		-51,062,724		-1,151,807	
Cmin/Cmax	=		0.20		TOTAL					-185,636,600		-185,636,600		0	
CALCULATED EFFECTIVENESS FOR SELECTED Tco					e = Cc(Tco-Tci))/ Cmin(Thi-Tci)					0.804					

APPENDIX 3

PREREFORMER SHEET

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PREREFORMER

3/29/99 15:29`			PREREFORMER												
			FUEL/STEAM/RECYCLE MIX, E-3, PREREFORMER, E-2												
STATION	3	TEMP, F	212.0	100.0	DEG C										
FUEL FROM FUEL PREHEATER E-6															
	GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL					
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	16.841					
MOLE FLOW, LB MOLES/HR			0.00	0.00	284.16	0.00	5.92	0.00	5.92	296.00					
MOLE FRACTION			0.000	0.000	0.960	0.000	0.020	0.000	0.020	1.000					
SPEC ENTH, BTU/LB MOLE			4,580	-97,445	-23,259	-44,293	-163,881	4,678	4,671						
ENTHALPY, BTU/HR			0	0	-6,609,252	0	-970,178	0	27,650	-7,551,780					
MASS FLOW LB/HR			0	0	4,558	0	261	0	166	4,985					
STATION	43	TEMP, F	239.0	115.0	DEG C										
STEAM FROM BOILER															
	GAS		H2	H2O(VAPOR)	CH4	CO	CO2	O2	N2	TOTAL					
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	18.016					
MOLE FLOW, LB MOLES/HR			0.000	444.000	0.000	0.000	0.000	0.000	0.000	444					
MOLE FRACTION			0.000	1.000	0.000	0.000	0.000	0.000	0.000	1.000					
SPEC ENTH, BTU/LB MOLE			4,768	-97,225	-23,003	-44,104	-163,620	4,872	4,859						
ENTHALPY, BTU/HR			0	-43,167,937	0	0	0	0	0	-43,167,937					
MASS FLOW LB/HR			0	7,999	0	0	0	0	0	7,999					
				RECYCLE FROM PREREFORMER											
				RECYCLE FRACTION						0.0154					
STATION	45	TEMP, F	611.3	321.9	DEG C										
RECYCLE FROM PREREFORMER EXIT															
	GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL					
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	16.892					
MOLE FLOW, LB MOLES/HR			0.892	6.499	4.221	0.000	0.315	0.000	0.093	12.020					
MOLE FRACTION			0.074	0.541	0.351	0.000	0.026	0.000	0.008	1.000					
SPEC ENTH, BTU/LB MOLE			7,369	-94,098	-18,892	-41,452	-159,683	7,635	7,492						
ENTHALPY, BTU/HR			6,576	-611,552	-79,743	0	-50,232	0	694	-734,257					
MASS FLOW LB/HR			2	117	68	0	14	0	3	203					
STATION	46	TEMP, F	250.1	121.2	DEG C										
RECYCLE/STEAM MIX															
	GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL					
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	17.986					
MOLE FLOW, LB MOLES/HR			0.892	450.499	4.221	0.000	0.315	0.000	0.093	456.020					
MOLE FRACTION			0.002	0.988	0.009	0.000	0.001	0.000	0.000	1.000					
SPEC ENTH, BTU/LB MOLE			4,845	-97,135	-22,896	-44,026	-163,511	4,951	4,936						
ENTHALPY, BTU/HR			4,323	-43,759,019	-96,644	0	-51,437	0	457	-43,902,319					
MASS FLOW LB/HR			2	8,116	68	0	14	0	3	8,202					

PREREFORMER

ENTHALPY IN , BTU/HR	44+45		-4.390E+07			HEAT BALANCE								
ENTHALPY OUT, BTU/HR			-4.390E+07			LOGIC FOR ITERATION	FALSE	FALSE						
HEAT LOSS, BTU/HR			0			ITERATION INCREMENT	21.68	-21.6						
NET ENTHALPY, IN - OUT , BTU/HR			125			TEMP CHANGE =	0.08							
STATION	4	TEMP, F	233.7	112.1	DEG C									
FUEL/STEAM/RECYCLE MIX														
GAS			H2	H2O	CH4	CO	CO2	O2	N2	TOTAL				
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	17.535				
MOLE FLOW, LB MOLES/HR			0.8924	450.4991	288.3811	0.0000	6.2346	0.0000	6.0126	752.020				
MOLE FRACTION			0.001	0.599	0.383	0.000	0.008	0.000	0.008	1.000				
SPEC ENTH, BTU/LB MOLE			4,731	-97,268	-23,053	-44,141	-163,671	4,834	4,822					
ENTHALPY, BTU/HR			4,222	-43,819,269	-6,648,108	0	-1,020,421	0	28,993	-51,454,583				
MASS FLOW LB/HR			2	8,116	4,626	0	274	0	168	13,187				
ENTHALPY IN , BTU/HR	46+3		-5.145E+07			HEAT BALANCE								
ENTHALPY OUT, BTU/HR			-5.145E+07			LOGIC FOR ITERATION	FALSE	FALSE						
HEAT LOSS, BTU/HR			0			ITERATION INCREMENT	26.07	-20.36						
NET ENTHALPY, IN - OUT , BTU/HR			484			TEMP CHANGE =	5.71							
			PREREFORMER INLET HEATER, E-3											
STATION	21	TEMP, F	922	494.6	DEG C									
HOT SIDE INLET														
GAS			H2	H2O	CH4	CO	CO2	O2	N2	TOTAL				
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	27.222				
MOLE FLOW, LB MOLES/HR			0.00	1058.02	0.00	0.00	303.79	377.67	3570.52	5309.99				
MOLE FRACTION			0.000	0.199	0	0.000	0.057	0.071	0.672	1.000				
SPEC ENTH, BTU/LB MOLE			9,557	-91,329	-14,661	-39,146	-156,031	10,072	9,769					
ENTHALPY, BTU/HR			0	-96,628,123	0	0	-47,400,183	3,804,003	34,878,886	-105,345,417				
MASS FLOW LB/HR			0	19,061	0	0	13,370	12,085	100,032	144,548				
STATION	22	TEMP, F	834	445.6	DEG C									
HOT SIDE EXIT														
GAS			H2	H2O	CH4	CO	CO2	O2	N2	TOTAL				
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	27.222				
MOLE FLOW, LB MOLES/HR			0	1058.02	0.00	0.00	303.79	377.67	3570.52	5309.99				
MOLE FRACTION			0.000	0.199	0	0.000	0.057	0.071	0.672	1.000				
SPEC ENTH, BTU/LB MOLE			8,935	-92,130	-15,931	-39,810	-157,094	9,371	9,114					
ENTHALPY, BTU/HR			0	-97,475,644	0	0	-47,723,036	3,538,977	32,542,916	-109,116,787				
MASS FLOW LB/HR			0	19,061	0	0	13,370	12,085	100,032	144,548				

PREREFORMER

STATION	4	TEMP, F	233.71	112.1	DEG C																				
COLD SIDE INLET																									
GAS			H2	H2O	CH4	CO	CO2	O2	N2	TOTAL															
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	17.535															
MOLE FLOW, LB MOLES/HR			0.89	450.50	288.38	0.00	6.23	0.00	6.01	752.02															
MOLE FRACTION			0.001	0.599	0.383	0.000	0.008	0.000	0.008	1.000															
SPEC ENTH, BTU/LB MOLE			4,731	-97,268	-23,053	-44,141	-163,671	4,834	4,822																
ENTHALPY, BTU/HR			4,222	-43,819,269	-6,648,108	0	-1,020,421	0	28,993	-51,454,583															
MASS FLOW LB/HR			2	8,116	4,626	0	274	0	168	13,187															
STATION	5	TEMP, F	752.00	400.0	DEG C																				
COLD SIDE EXIT																									
GAS			H2	H2O	CH4	CO	CO2	O2	N2	TOTAL															
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	17.535															
MOLE FLOW, LB MOLES/HR			0.89	450.50	288.38	0.00	6.23	0.00	6.01	752.02															
MOLE FRACTION			0.001	0.599	0.383	0.000	0.008	0.000	0.008	1.000															
SPEC ENTH, BTU/LB MOLE			8,357	-92,865	-17,064	-40,421	-158,065	8,725	8,511																
ENTHALPY, BTU/HR			7,457	-41,835,438	-4,920,940	0	-985,468	0	51,175	-47,683,213															
MASS FLOW LB/HR			2	8,116	4,626	0	274	0	168	13,187															
STATION		STATION		ASSUMED EFFECTIVENESS, e =								0.753													
Thi, F	=	21	922	22	CALCULATED Tho F =								834.1		=Thi-eCmin/Ch(Thi-Tci)										
Tci, F	=	4	234	5	CALCULATED Tco F								752.0		=Tci+eCmin/Cc(Thi-Tci)										
Ch=DH/DT	=	BTU/HR F	42775.43	ENERGY BALANCE								IN		OUT		DELTA									
Cc=DH/DT	=	BTU/HR F	7276.56	HOT SIDE								-105,345,417		-109,116,787		3,771,370									
Cmin	=	BTU/HR F	7276.56	COLD SIDE								-51,454,583		-47,683,213		-3,771,370									
Thi-Tci, F	=		689	TOTAL								-156800000		-156800000		0									
Cmin/Cmax	=		0.17																						
CALCULATED EFFECTIVENESS FOR SELECTED Tco										e = Cc(Tco-Tci))/ Cmin(Thi-Tci)										0.753					

PREREFORMER

BACK SHIFT CONVERSION FRACTION = y			REFORM CONVERSION FRACTION = X								
a =	28.7035501		ASSUME VALUE X =			0.01900		1+2x=	1.0300		
b =	17.860		CALCULATED FRACTIONS								
c =	0.326846952		H2	H2O	CH4	CO	CO2				
y =	-0.018872439		0.07705	0.56118	0.36448	0.00013	0.02716		1.0300		
CALCULATED CONVERSION CONSTANT Kp			CALCULATE Kp =			3.48316E-07					
			Kp ACT-Kp(X) =			3.40178E-11					
			ITERATION FOR CLOSURE ON X								
			LOGIC FOR ITERATION				FALSE	FALSE			
			ITERATION INCREMENT				0.013757	-0.014759			
			REVISED VALUE X =				-1.00E-03				
			CH4 CONVERTED IN RU ,LB MOLES / HR				14.2869				
			CH4 CONVERSION =				0.0495				
STATION	6	TEMP, F	611.3	321.9	DEG C						
PREREFORMER EXIT											
	GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL	
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	17.025	
MOLE FLOW, LB MOLES/HR			57.945	422.020	274.094	0.094	20.427	0.000	6.013	774.581	
MOLE FRACTION			0.0748	0.5448	0.3539	0.0001	0.0264	0.0000	0.0078	1.008	
SPEC ENTH, BTU/LB MOLE			7,369	-94,098	-18,892	-41,452	-159,683	7,635	7,492		
ENTHALPY, BTU/HR			427,006	-39,711,144	-5,178,092	-3,914	-3,261,848	0	45,044	-47,682,947	
MASS FLOW LB/HR			117	7,603	4,397	3	899	0	168	13,187	
ENTHALPY IN, BTU/HR				-4.7683E+07			HEAT BALANCE				
ENTHALPY OUT, BTU/HR				-4.7683E+07			LOGIC FOR ITERATION	FALSE	FALSE		
HEAT LOSS, BTU/HR				0			ITERATION INCREMENT	137.98	-140.04		
NET ENTHALPY, IN - OUT, BTU/HR				-266			TEMP CHANGE =	-2.06			
PREREFORMER EXIT RECYCLE TO RECYCLE EJECTOR											
RECYCLE FRACTION =		0.0154									
FRACTION TO E-2		0.9846									
PREREFORMER EXIT HEATER, E-2											
STATION	20	TEMP, F	982	527.8	DEG C						
HOT SIDE INLET											
	GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL	
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	27.222	
MOLE FLOW, LB MOLES/HR			0.00	1058.02	0.00	0.00	303.79	377.67	3570.52	5309.99	
MOLE FRACTION			0.000	0.199	0	0.000	0.057	0.071	0.672	1.000	
SPEC ENTH, BTU/LB MOLE			9,980	-90,778	-13,770	-38,692	-155,299	10,553	10,217		
ENTHALPY, BTU/HR			0	-96,045,070	0	0	-47,177,799	3,985,531	36,478,474	-102,758,864	
MASS FLOW LB/HR			0	19,061	0	0	13,370	12,085	100,032	144,548	

PREREFORMER

STATION		21	TEMP, F		922	494.6	DEG C										
HOT SIDE EXIT																	
GAS				H2	H2O	CH4	CO	CO2	O2	N2	TOTAL						
MOLECULAR WEIGHT				2.016	18.016	16.042	28.01	44.01	32	28.016	27.222						
MOLE FLOW, LB MOLES/HR				0.00	1058.02	0.00	0.00	303.79	377.67	3570.52	5309.99						
MOLE FRACTION				0.000	0.199	0	0.000	0.057	0.071	0.672	1.000						
SPEC ENTH, BTU/LB MOLE				9,557	-91,329	-14,661	-39,146	-156,031	10,072	9,769							
ENTHALPY, BTU/HR				0	-96,628,123	0	0	-47,400,183	3,804,003	34,878,886	-105,345,417						
MASS FLOW LB/HR				0	19,061	0	0	13,370	12,085	100,032	144,548						
STATION	7	TEMP, F	611.34	321.9	DEG C												
COLD SIDE INLET																	
GAS				H2	H2O	CH4	CO	CO2	O2	N2	TOTAL						
MOLECULAR WEIGHT				2.016	18.016	16.042	28.01	44.01	32	28.016	16.894						
MOLE FLOW, LB MOLES/HR				57.05	415.52	269.87	0.09	20.11	0.00	5.92	768.57						
MOLE FRACTION				0.074	0.541	0.351	0.000	0.026	0.000	0.008	1.000						
SPEC ENTH, BTU/LB MOLE				7,369	-94,098	-18,892	-41,452	-159,683	7,635	7,492							
ENTHALPY, BTU/HR				420,431	-39,099,592	-5,098,349	-3,854	-3,211,616	0	44,350	-46,948,630						
MASS FLOW LB/HR				115	7,486	4,329	3	885	0	166	12,984						
STATION	8	TEMP, F	932.00	500.0	DEG C												
COLD SIDE EXIT																	
GAS				H2	H2O	CH4	CO	CO2	O2	N2	TOTAL						
MOLECULAR WEIGHT				2.016	18.016	16.042	28.01	44.01	32	28.016	16.894						
MOLE FLOW, LB MOLES/HR				57.05	415.52	269.87	0.09	20.11	0.00	5.92	768.57						
MOLE FRACTION				0.074	0.541	0.351	0.000	0.026	0.000	0.008	1.000						
SPEC ENTH, BTU/LB MOLE				9,626	-91,240	-14,518	-39,073	-155,912	10,150	9,841							
ENTHALPY, BTU/HR				549,164	-37,912,100	-3,917,989	-3,633	-3,135,780	0	58,260	-44,362,077						
MASS FLOW LB/HR				115	7,486	4,329	3	885	0	166	12,984						
		STATION		STATION	ASSUMED EFFECTIVENESS, e =				0.865								
Thi, F	=	20	982	21	CALCULATED Tho F =				922.3	=Thi-eCmin/Ch(Thi-Tci)							
Tci, F	=	7	611	8	CALCULATED Tco F				932.0	=Tci+eCmin/Cc(Thi-Tci)							
Ch=DH/DT	=	BTU/HR F	43227.58	<u>ENERGY BALANCE</u>				IN	OUT	DELTA							
Cc=DH/DT	=	BTU/HR F	8066.34	HOT SIDE				-102,758,864	-105,345,417	2,586,553							
Cmin	=	BTU/HR F	8066.34	COLD SIDE				-46,948,630	-44,362,077	-2,586,553							
Thi-Tci, F	=		371	TOTAL				-149,707,494	-149,707,494	0							
Cmin/Cmax	=		0.19														
CALCULATED EFFECTIVENESS FOR SELECTED Tco				e = Cc(Tco-Tci))/ Cmin(Thi-Tci)				0.865									

APPENDIX 4

CELL STACK SHEET

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CELL STACK

3/29/99 15:29 `		CELL STACK (INCLUDES INTERNAL REFORMER UNITS)								
STATION	9	TEMP, F	932.0	500.0	DEG C		PRESSURE , ATM =	1.041		
RU INLET = E-2 EXIT(8)							ATM^2 =	1.083		
	GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	16.894
MOLE FLOW, LB MOLES/HR			57.05	415.52	269.87	0.09	20.11	0.00	5.92	768.57
MOLE FRACTION			0.0742	0.5406	0.3511	0.0001	0.0262	0.0000	0.0077	1.000
SPEC ENTH, BTU/LB MOLE			9,626	-91,240	-14,518	-39,073	-155,912	10,150	9,841	
ENTHALPY, BTU/HR			549,164	-37,912,100	-3,917,989	-3,633	-3,135,780	0	58,260	-44,362,077
MASS FLOW LB/HR			115	7,486	4,329	3	885	0	166	12,984
SHIFT EQUILIBRIUM CONSTANT			REFORM EQUILIBRIUM CONSTANT							
TEMP, DEG F		1100	TEMP,DEG F =			1100	APPROACH TEMP		0	
TEMP, DEG R		1560	TEMP, DEG R =			1560				
V=1/Tr =		0.00064103	V=1/Tr =			0.000641026				
log10 Kp=		0.42324328	Log10 Kp =			-0.386216245				
Kp=		2.6500	Kp =			0.41095				
BACK SHIFT CONVERSION FRACTION = y			REFORM CONVERSION FRACTION = X							
a =	1.649984202		ASSUME VALUE X =			0.18023	1+2x=		1.3605	
b =	2.074		CALCULATED FRACTIONS							
c =	0.156157948		H2	H2O	CH4	CO	CO2			
y =	-0.080436604		0.69536	0.27997	0.17091	0.09991	0.10661	1.3528		
CALCULATED CONVERSION CONSTANT Kp			CALCULATE Kp =			0.410921919				
			Kp ACT-Kp(X) =			2.31329E-05				
			ITERATION FOR CLOSURE ON X							
			LOGIC FOR ITERATION			FALSE		FALSE		
			ITERATION INCREMENT			0.07271		-0.09248		
			REVISED VALUE X =			0.18023				
			CH4 CONVERTED IN RU ,LB MOLES / HR			138.5198				
			CH4 CONVERSION =			0.5133				
STATION	10	TEMP, F	1100.0	593.3	DEG C					
RU OUT										
	GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	12.418
MOLE FLOW, LB MOLES/HR			534.43	215.18	131.35	76.79	81.93	0.00	5.92	1045.61
MOLE FRACTION			0.511	0.206	0.126	0.073	0.078	0.000	0.006	1.000
SPEC ENTH, BTU/LB MOLE			10,817	-89,674	-11,945	-37,787	-153,833	11,510	11,109	
ENTHALPY, BTU/HR			5,780,950	-19,296,115	-1,569,044	-2,901,683	-12,604,130	0	65,763	-30,524,258
MASS FLOW LB/HR			1,077	3,877	2,107	2,151	3,606	0	166	12,984
ENTHALPY IN , BTU/HR			-4.4362E+07							
ENTHALPY OUT, BTU/HR			-3.0524E+07							
RU HEAT LOSS, BTU/HR			0			INCLUDED IN STACK LOSS				
NET ENTHALPY, IN - OUT , BTU/HR			-13,837,819			TO BE SUBTRACTED FROM STACK WASTE HEAT IN STACK BALANCE				

CELL STACK

						FUEL UTILIZATION=H2 CONSUMED/H2 CONSUMED +H2 EXIT +CO EXIT =					
CELL STACK POWER, KW			17619.00	60,133,647	BTU/HR						0.780
CELL VOLTAGE , V/C			0.817								
FUEL CELL EFFICIENCY ,%			0.637								
CURRENT DENSITY,A/CM2			0.16	H2	CO2 TRNSFER	H2O FORMED	O2				
CONSUMPTIONS, LB MOLES/HR				886.773	886.773	886.773	443.386				
STATION			11	TEMP, F	1100.0	593.3	DEG C				
ANODE IN											
GAS				H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT				2.016	18.016	16.042	28.010	44.010	32.000	28.016	12.418
MOLE FLOW, LB MOLES/HR				534.43	215.18	131.35	76.79	81.93	0.00	5.92	1045.61
MOLE FRACTION				0.511	0.206	0.126	0.073	0.078	0.000	0.006	1.000
SPEC ENTH, BTU/LB MOLE				10,817	-89,674	-11,945	-37,787	-153,833	11,510	11,109	
ENTHALPY, BTU/HR				5,780,950	-19,296,115	-1,569,044	-2,901,683	-12,604,130	0	65,763	-30,524,258
MASS FLOW LB/HR				1,077	3,877	2,107	2,151	3,606	0	166	12,984
Cp, BTU/LB MOLE/DEG F				7.16	9.46	15.82	7.73	12.54	8.16	7.62	0.09
NON EQUILIB OUT FLOWS OF H2, CO, H2O,&CO2(ASSUMING 100% CH4 CONVERSION)											
MOLE FLOW, LB MOLES/HR				41.721	970.599		208.145	968.706			2189.171
MOLE FRACTION				0.019	0.443		0.095	0.442			1.000
				C	B		A	D			
EQUILIB OUT FLOWS											
MOLE FRACTION				0.072	0.391		0.042	0.495			1.000
MOLE FLOW, LB MOLES/HR				156.955	855.365		92.911	1083.940			2189.171
EQUILIB TEMP				=	1170.00		FRACTION CO CONVERTED =			0.052638075	
SHIFT EQUILIBRIUM CONSTANT =					2.141		a=	1.14073			
							b=	-1.61422			
							c=	0.08181			
SHIFT EQUILIBRIUM CALCULATION FOR ANODE EXIT STREAM											
TEMP, DEG F =				1170.00							
TEMP, DEG R =				1630							
V=1/Tr=				0.000613497							
log10Kp=				0.330561478							
Kp =				2.1407							
STATION			12	TEMP, F	1170.00	632.2	DEG C				
ANODE OUT FLOW											
GAS				H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT				2.016	18.016	16.042	28.010	44.010	32.000	28.016	30.158
MOLE FLOW, LB MOLES/HR				156.95	855.37	0.00	92.91	1083.94	0.00	5.92	2195.09
MOLE FRACTION				0.072	0.390	0.000	0.042	0.494	0.000	0.003	1.000
SPEC ENTH, BTU/LB MOLE				11,316	-89,008	-10,821	-37,243	-152,949	12,083	11,644	
ENTHALPY, BTU/HR				1,776,111	-76,134,278	0	-3,460,291	-165,787,892	0	68,934	-243,537,416
MASS FLOW LB/HR				316	15,410	0	2,602	47,704	0	166	66,199
Cp, BTU/LB MOLE/DEG F				7.13	9.58	16.23	7.79	12.68	8.21	7.67	0.11

CELL STACK

CATHODE						CATHODE OXYGEN UTILIZATION= O2 CONSUMED/O2 IN=	0.54		
STATION	17	TEMP, F	1035.8	557.6	DEG C				
CATHODE IN FLOW FROM E-9									
GAS			H2	H2O	CH4	CO	CO2	O2	N2
MOLECULAR WEIGHT			2.016	18.016	16.042	28.010	44.010	32.000	28.016
MOLE FLOW, LB MOLES/HR			0.00	1058.02	0.00	0.00	1190.56	821.05	3570.52
MOLE FRACTION			0.000	0.159	0.000	0.000	0.179	0.124	0.538
SPEC ENTH, BTU/LB MOLE			10,360	-90,279	-12,950	-38,282	-154,636	10,987	10,621
ENTHALPY, BTU/HR			0	-95,516,844	0	0	-184,102,916	9,020,778	37,922,435
MASS FLOW LB/HR			0	19,061	0	0	52,397	26,274	100,032
Cp, BTU/LB MOLE/DEG F			7.18	9.35	15.44	7.67	12.41	8.11	7.56
STATION	18	TEMP, F	1173.6	634.2	DEG C				
CATHODE OUT FLOW									
GAS			H2	H2O	CH4	CO	CO2	O2	N2
MOLECULAR WEIGHT			2.016	18.016	16.042	28.010	44.010	32.000	28.016
MOLE FLOW, LB MOLES/HR			0.00	1058.02	0.00	0.00	303.79	377.67	3570.52
MOLE FRACTION			0.000	0.199	0.000	0.000	0.057	0.071	0.672
SPEC ENTH, BTU/LB MOLE			11,342	-88,973	-10,762	-37,215	-152,903	12,113	11,672
ENTHALPY, BTU/HR			0	-94,135,253	0	0	-46,449,993	4,574,753	41,675,666
MASS FLOW LB/HR			0	19,061	0	0	13,370	12,085	100,032
Cp, BTU/LB MOLE/DEG F			7.13	9.58	16.25	7.79	12.68	8.21	7.68
STACK HEAT BALANCE									
POWER OUTPUT, BTU/HR			=	60,133,647		STREAM DELTA T =		70.00	137.89
HEAT LOSS FROM STACKS , BTU/HR			=	700,000					
HEAT ABSORBED BY RU			=	-13,837,819		LOGIC FOR CATHODE EXIT TEMP TO BALANCE STACK ENERGY			
ANODE STREAM ENTHALPY IN, BTU/HR			=	-30,524,258		CATHODE EXIT TEMP CORRECTION TO NULL ENTHALPY DELTA			
ANODE STREAM ENTHALPY OUT, BTU/HR			=	-243,537,416		LOGIC FOR ITERATION	FALSE	FALSE	
CATH STRM ENTHALPY IN , BTU/HR			=	-232,676,546		ITERATION INCREMENT	13.24	-19.6	
CATH STRM ENTHALPY OUT, BTU/HR			=	-94,334,827		TEMP CHANGE =	-6.36		
NET ENTHALPY , BTU/HR		IN-OUT	=	-28					

APPENDIX 5

COMPRESSOR SHEET

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GTCOMPRESSOR

3/29/99 15:29					THREE STAGE COMPRESSOR WITH INTERCOOLER E-10 & E-11 AND COOLING UNIT E-12					
					1ST STAGE	COMPRESSORS PRESSURE RATIO				2.63
						COMPRESSOR EFFICIENCY,				0.84
1ST STAGE COMPRESSOR				'INTAKE LOSS ,% = 1		PROCESS AIR FLOW, LB MOLES /HR =				4570
						WATER VAPOR PRESSURE,PSIA				0.147
STATION 28	TEMP, F	59.0	15.0	DEG C		DRY FRACTION				0.99
1ST STAGE COMPRESSOR INLET AIR						PRESSURE IN , PSIA =				14.454
	GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	28.789
MOLE FLOW, LB MOLES/HR			0.00	45.70	0.00	0.00	13.71	945.99	3564.60	4570.00
MOLE FRACTION			0.00000	0.01000	0.00000	0.00000	0.00300	0.20700	0.78000	1.000
SPEC ENTH, BTU/LB MOLE			3,519	-98,679	-24,607	-45,359	-165,292	3,600	3,605	
ENTHALPY, BTU/HR			0	-4,509,641	0	0	-2,265,928	3,405,559	12,848,860	9,478,849
MASS FLOW LB/HR			0	823	0	0	603	30,272	99,866	131,564
COMPRESSOR EXIT TEMP & PRESSURE					FIRST STAGE COMPRESSR WORK, Wci, (ADIABATIC)					
To = Ti (Po/Pi)^(K-1/K) +(Wca-Wci-HL)/wcCp -460					=(wc/m) RTci (k/k-1)/((Po/Pi)^(k-1/k)-1)					
					R =1544/778.3= 1.984 BTU /lb mole deg R					
			COMP-1		Wci =BTU/HR, kW		IDEAL WORK		5,238,069	1,535
TEMP IN , F	Tci		59.0	15.0	DEG C	wc = LB/HR				131,564
TEMP OUT, F	Tco		254.6	123.6	DEG C	m = molecular wt lbs/lb mole				28.789
PRESSURE RATIO	Po/Pi		2.630			Po/Pi = pressure ratio psi				2.630
EXIT PRESSURE, PSIA			38.01402			k = constant for gas, Cp/Cv				1.3978
						Cp , BTU/LB F				0.2435
						k-1/k				0.285
						n = ADIABATIC efficiency				0.84
						Wca= BTU/HR, kW		ACTUAL WORK		6,235,796
										1,827
STATION 29	TEMP, F	253.7	123.1	DEG C						
1ST STAGE COMPRESSOR AIR EXIT										
	GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	28.789
MOLE FLOW, LB MOLES/HR			0.00	45.70	0.00	0.00	13.71	945.99	3564.60	4570.00
MOLE FRACTION			0.00000	0.01000	0.00000	0.00000	0.00300	0.20700	0.78000	1.000
SPEC ENTH, BTU/LB MOLE			4,870	-97,105	-22,861	-44,001	-163,476	4,977	4,961	
ENTHALPY, BTU/HR			0	-4,437,710	0	0	-2,241,032	4,708,268	17,684,988	15,714,514
MASS FLOW LB/HR			0	823	0	0	603	30,272	99,866	131,564
ENTHALPY IN , BTU/HR				9.4788E+06			HEAT BALANCE			
COMPRESSOR POWER IN BT/HR				6.2358E+06			LOGIC FOR ITERATION		FALSE	FALSE
ENTHALPY OUT, BTU/HR				1.5715E+07			ITERATION INCREMENT		1.24	-2.14
HEAT LOSS, BTU/HR				0			TEMP CHANGE =		-0.9	
NET ENERGY, IN - OUT , BTU/HR				130						

GTCOMPRESSOR

FIRST INTERCOOLER E-10										
STATION	49	TEMP, F	86.0	30.0	DEG C					
COLD SIDE INLET										
GAS			H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	18.016
MOLE FLOW, LB MOLES/HR			0.00	2000.00	0.00	0.00	0.00	0.00	0.00	2000.00
MOLE FRACTION			0.00000	1.00000	0.00000	0.00000	0.00000	0.00000	0.00000	1.000
SPEC ENTH, BTU/LB MOLE			3,706	-117,114	-24,382	-45,171	-165,053	3,788	3,793	
ENTHALPY, BTU/HR			0	-2.34E+08	0	0	0	0	0	-234,228,032
MASS FLOW LB/HR			0	36,032	0	0	0	0	0	36,032
STATION	50	TEMP, F	203.2	95.1	DEG C					
COLD SIDE EXIT										
GAS			H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	18.016
MOLE FLOW, LB MOLES/HR			0.00	2000.00	0.00	0.00	0.00	0.00	0.00	2000.00
MOLE FRACTION			0.00000	1.00000	0.00000	0.00000	0.00000	0.00000	0.00000	1.000
SPEC ENTH, BTU/LB MOLE			4,519	-115,002	-23,341	-44,354	-163,966	4,616	4,609	
ENTHALPY, BTU/HR			0	-230,004,171	0	0	0	0	0	-230,004,171
MASS FLOW LB/HR			0	36,032	0	0	0	0	0	36,032
		STATION	STATION		ASSUMED EFFECTIVENESS, e =		0.785			
Thi, F	=	29	254	30	CALCULATED Tho F =		122.0		=Thi-eCmin/Ch(Thi-Tci)	
Tci, F	=	49	86	50	COLD SIDE EXIT		203.2			
Ch=DH/DT	=	BTU/HR F	32082	ENERGY BALANCE		IN	OUT	DELTA		
Cc=DH/DT	=	BTU/HR F	36032	HOT SIDE		15,714,514	11,490,654	4,223,861		
Cmin	=	BTU/HR F	32082	COLD SIDE		-234,228,032	-230,004,171	-4,223,861		
Thi-Tci, F	=		168	TOTAL		-218,513,518	-218,513,518	0		
Cmin/Cmax	=		1.12							
CALCULATED EFFECTIVENESS FOR SELECTED Tco				e = Cc(Tco-Tci)/ Cmin(Thi-Tci)				0.785		
CALCULATED EFFECTIVENESS FOR SELECTED Tho				e = Ch(Thi-Tho)/ Cmin(Thi-Tci)				0.785		
2ND STAGE COMPRESSOR										
					2ND STAGE	COMPRESSORS PRESSURE RATIO				2.63
						COMPRESSOR EFFICIENCY,				0.86
STATION	30	TEMP, F	122.0	50.0	DEG C	PRESSURE LOSS, DUCT/ INLET, %				3.0
2ND STAGE COMPRESSOR INLET AIR					PRESSURE IN , PSIA =					35.87
GAS			H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	28.789
MOLE FLOW, LB MOLES/HR			0.00	45.70	0.00	0.00	13.71	945.99	3564.60	4570.00
MOLE FRACTION			0.00000	0.01000	0.00000	0.00000	0.00300	0.20700	0.78000	1.000
SPEC ENTH, BTU/LB MOLE			3,955	-98,174	-24,073	-44,921	-164,727	4,040	4,043	
ENTHALPY, BTU/HR			0	-4,486,535	0	0	-2,258,181	3,822,170	14,413,199	11,490,654
MASS FLOW LB/HR			0	823	0	0	603	30,272	99,866	131,564

GTCOMPRESSOR

COMPRESSOR EXIT TEMP					SECOND STAGE COMPRESSR WORK, Wci, (ADIABATIC)					
To = Ti (Po/Pi)^(K-1/K) +(Wca-Wci-HL)/wcCp -460					=(wc/m) RTci (k/k-1)/((Po/Pi)^(k-1/k)-1)					
					R =1544/778.3= 1.984 BTU /lb mole deg R					
					Wci =BTU/HR, kW			IDEAL WORK	5,875,271	1,721
TEMP IN , F	Tci		122.0	50.0	DEG C	wc = LB/HR			131,564	
TEMP OUT, F	Tco		336.4	169.1	DEG C	m = molecular wt lbs/lb mole			28.789	
PRESSURE RATIO	Po/Pi		2.630			Po/Pi = pressure ratio psi			2.630	
EXIT PRESSURE, PSIA=			94.35			k = constant for gas, Cp/Cv			1.3987	
					Cp , BTU/LB F			0.2446		
					k-1/k			0.285		
					n = ADIABATIC efficiency			0.86		
					Wca= BTU/HR, kW			ACTUAL WORK	6,831,710	2,002
STATION	31	TEMP, F	334.3	168.0	DEG C					
2ND STAGE COMPRESSOR AIR EXIT										
GAS			H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	28.789
MOLE FLOW, LB MOLES/HR			0.00	45.70	0.00	0.00	13.71	945.99	3564.60	4570.00
MOLE FRACTION			0.00000	0.01000	0.00000	0.00000	0.00300	0.20700	0.78000	1.000
SPEC ENTH, BTU/LB MOLE			5,432	-96,441	-22,052	-43,434	-162,667	5,562	5,526	
ENTHALPY, BTU/HR			0	-4,407,371	0	0	-2,229,942	5,261,821	19,697,952	18,322,460
MASS FLOW LB/HR			0	823	0	0	603	30,272	99,866	131,564
ENTHALPY IN , BTU/HR			1.1491E+07			HEAT BALANCE				
COMPRESSOR POWER IN BT/HR			6.8317E+06			LOGIC FOR ITERATION			FALSE	FALSE
ENTHALPY OUT, BTU/HR			1.8322E+07			ITERATION INCREMENT			9.06	-11.16
HEAT LOSS, BTU/HR			0			TEMP CHANGE =			-2.1	
NET ENERGY, IN - OUT , BTU/HR			-96							
SECOND INTERCOOLER E-11										
STATION	51	TEMP, F	86.0	30.0	DEG C					
COLD SIDE INLET										
GAS			H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	18.016
MOLE FLOW, LB MOLES/HR			0.00	3200.00	0.00	0.00	0.00	0.00	0.00	3200.00
MOLE FRACTION			0.00000	1.00000	0.00000	0.00000	0.00000	0.00000	0.00000	1.000
SPEC ENTH, BTU/LB MOLE			3,706	-117,114	-24,382	-45,171	-165,053	3,788	3,793	
ENTHALPY, BTU/HR			0	-374,764,851	0	0	0	0	0	-374,764,851
MASS FLOW LB/HR			0	57,651	0	0	0	0	0	57,651
STATION	52	TEMP, F	204.5	95.8	DEG C					
COLD SIDE EXIT										
GAS			H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	18.016
MOLE FLOW, LB MOLES/HR			0.00	3200.00	0.00	0.00	0.00	0.00	0.00	3200.00
MOLE FRACTION			0.00000	1.00000	0.00000	0.00000	0.00000	0.00000	0.00000	1.000
SPEC ENTH, BTU/LB MOLE			4,528	-114,979	-23,329	-44,345	-163,954	4,625	4,618	
ENTHALPY, BTU/HR			0	-367,933,045	0	0	0	0	0	-367,933,045
MASS FLOW LB/HR			0	57,651	0	0	0	0	0	57,651

GTCOMPRESSOR

		STATION		STATION	ASSUMED EFFECTIVENESS, e =		0.855		
Thi, F	=	31	334	32	CALCULATED Tho F =		122.0	=Thi-eCmin/Ch(Thi-Tci)	
Tci, F	=	51	86	52	COLD SIDE EXIT		204.5		
Ch=DH/DT	=	BTU/HR F	32174		ENERGY BALANCE	IN	OUT	DELTA	
Cc=DH/DT	=	BTU/HR F	57651		HOT SIDE	18,322,460	11,490,654	6,831,806	
Cmin	=	BTU/HR F	32174		COLD SIDE	-374,764,851	-367,933,045	-6,831,806	
Thi-Tci, F	=		248		TOTAL	-356,442,391	-356,442,391	0	
Cmin/Cmax	=		1.79						
CALCULATED EFFECTIVENESS FOR SELECTED Tco					e = Cc(Tco-Tci))/ Cmin(Thi-Tci)		0.855		
CALCULATED EFFECTIVENESS FOR SELECTED Tho					e = Ch(Thi-Tho))/ Cmin(Thi-Tci)		0.855		
3RD STAGE COMPRESSOR					3RD STAGE	COMPRESSORS PRESSURE RATIO			2.63
STATION	32	TEMP, F	122.0	50.0	DEG C	COMPRESSOR EFFICIENCY,			0.857
3RD STAGE COMPRESSOR AIR INLET						PRESSURE LOSS,DUCT/ INLET,%			3.0
						PRESSURE IN , PSIA =			90.52
	GAS		H2	H2O	CH4	CO	CO2	O2	N2
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016
MOLE FLOW, LB MOLES/HR			0.00	45.70	0.00	0.00	13.71	945.99	3564.60
MOLE FRACTION			0.00000	0.05830	0.00000	0.00000	0.00909	0.62727	2.36364
SPEC ENTH, BTU/LB MOLE			3,955	-98,174	-24,073	-44,921	-164,727	4,040	4,043
ENTHALPY, BTU/HR			0	-4,486,535	0	0	-2,258,181	3,822,170	14,413,199
MASS FLOW LB/HR			0	823	0	0	603	30,272	99,866
COMPRESSOR EXIT TEMP					THIRD STAGE COMPRESSR WORK, Wci, (ADIABATIC)				
To = Ti (Po/Pi)^(K-1/K) +(Wca-Wci-HL)/wcCp -460					=(wc/m) RTci (k/k-1)/((Po/Pi)^(k-1/k)-1)				
					R =1544/778.3= 1.984 BTU /lb mole deg R				
					Wci =BTU/HR, kW				
					IDEAL WORK				
TEMP IN , F	Tci		122.0	50.0	DEG C	wc = LB/HR			5,883,296
TEMP OUT, F	Tco		339.2	170.7	DEG C	m = molecular wt lbs/lb mole			1,724
PRESSURE RATI	Po/Pi		2.630			Po/Pi = pressure ratio psi			131,564
PRESSURE OUT ,PSIA =			238.06			k = constant for gas, Cp/Cv			28,789
						Cp , BTU/LB F			2,630
						k-1/k			1.404
						n = ADIABATIC efficiency			0.2446
						Wca= BTU/HR, kW			0.288
						ACTUAL WORK			0.857
									6,864,990
STATION	33	TEMP, F	335.4	168.5	DEG C				2,011
3RD STAGE COMPRESSOR AIR EXIT									
	GAS		H2	H2O	CH4	CO	CO2	O2	N2
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016
MOLE FLOW, LB MOLES/HR			0.00	45.70	0.00	0.00	13.71	945.99	3564.60
MOLE FRACTION			0.00000	0.01000	0.00000	0.00000	0.00300	0.20700	0.78000
SPEC ENTH, BTU/LB MOLE			5,439	-96,433	-22,042	-43,427	-162,657	5,570	5,533
ENTHALPY, BTU/HR			0	-4,406,986	0	0	-2,229,800	5,268,852	19,723,393
MASS FLOW LB/HR			0	823	0	0	603	30,272	99,866
TOTAL									131,564

GTCOMPRESSOR

ENTHALPY IN , BTU/HR			1.1491E+07		HEAT BALANCE				
COMPRESSOR POWER IN BT/HR			6.8650E+06		LOGIC FOR ITERATION		FALSE	FALSE	
ENTHALPY OUT, BTU/HR			1.8355E+07		ITERATION INCREMENT		12.7	-16.57	
HEAT LOSS, BTU/HR			0		TEMP CHANGE =		-3.87		
NET ENERGY, IN - OUT , BTU/HR			185						
INTERCOOLER COOLANT LOOP AND COOLING UNIT									
STATION	53	TEMP, F	204.0	95.6	DEG C				
COOLANT MIX FROM INTERCOOLERS									
GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT		2.016	18.016	16.042	28.01	44.01	32	28.016	18.016
MOLE FLOW, LB MOLES/HR		0.00	5200.00	0.00	0.00	0.00	0.00	0.00	5200.00
MOLE FRACTION		0.00000	1.00000	0.00000	0.00000	0.00000	0.00000	0.00000	1.000
SPEC ENTH, BTU/LB MOLE		4,522	-114,988	-23,536	-44,364	-164,052	4,589	4,602	
ENTHALPY, BTU/HR		0	-597,937,610	0	0	0	0	0	-597,937,610
MASS FLOW LB/HR		0	93,683	0	0	0	0	0	93,683
ENTHALPY IN , BTU/HR		50+52	-5.9794E+08		HEAT BALANCE				
ENTHALPY OUT, BTU/HR			-5.9794E+08		LOGIC FOR ITERATION		FALSE	FALSE	
HEAT LOSS, BTU/HR			0		ITERATION INCREMENT		129.157	-130.15	
NET ENTHALPY, IN - OUT , BTU/HR			393		TEMP CHANGE =		-0.993		
STATION	47	TEMP, F	86.0	30.0	DEG C				
COOLANT FROM COOLING TOWER									
GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT		2.016	18.016	16.042	28.01	44.01	32	28.016	18.016
MOLE FLOW, LB MOLES/HR		0.00	5200.00	0.00	0.00	0.00	0.00	0.00	5200.00
MOLE FRACTION		0.00000	1.00000	0.00000	0.00000	0.00000	0.00000	0.00000	1.000
SPEC ENTH, BTU/LB MOLE		3,697	-117,114	-24,867	-45,207	-165,317	3,709	3,766	
ENTHALPY, BTU/HR		0	-6.09E+08	0	0	0	0	0	-6.09E+08
MASS FLOW LB/HR		0	93,683	0	0	0	0	0	93,683
COOLING UNIT AIR									
					COOLING UNIT AIR FLOW, LB MOLES /HR =			120000	
					WATER VAPOR PRESSURE,PSIA			0.147	
STATION	54	TEMP, F	59.0	15.0	DEG C	DRY FRACTION			0.99000
COOLING TOWER INLET AIR									
GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT		2.016	18.016	16.042	28.01	44.01	32	28.016	28.789
MOLE FLOW, LB MOLES/HR		0.00	1200.00	0.00	0.00	359.96	24839.89	93600.03	120000
MOLE FRACTION		0.00000	0.01000	0.00000	0.00000	0.00300	0.20700	0.78000	1.000
SPEC ENTH, BTU/LB MOLE		3,519	-98,679	-24,607	-45,359	-165,292	3,600	3,605	
ENTHALPY, BTU/HR		0	-118,415,087	0	0	-59,499,213	89,423,859	337,388,005	248,897,563
MASS FLOW LB/HR		0	21,619	0	0	15,842	794,877	2,622,298	3,454,636

GTCOMPRESSOR

STATION	55	TEMP, F	72.2	22.3	DEG C		DRY FRACTION			1.00000
COOLING TOWER EXIT AIR										
GAS			H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	28.789
MOLE FLOW, LB MOLES/HR			0.00	1200.00	0.00	0.00	359.96	24839.89	93600.03	119999.88
MOLE FRACTION			0.00000	0.01000	0.00000	0.00000	0.00300	0.20700	0.78000	1.000
SPEC ENTH, BTU/LB MOLE			3,610	-98,574	-24,498	-45,267	-165,176	3,692	3,697	
ENTHALPY, BTU/HR			0	-118,288,339	0	0	-59,457,292	91,704,765	345,993,703	259,952,836
MASS FLOW LB/HR			0	21,619	0	0	15,842	794,877	2,622,298	3,454,636
STATION			STATION		ASSUMED EFFECTIVENESS, e =			0.814		
Thi, F	=	53	204	47	SELECTED Tho			86.0	=Thi-eCmin/Ch(Thi-Tci)	
Tci, F	=	54	59	55	COLD SIDE EXIT			72.2		
Ch=DH/DT	=	BTU/HR F	93683	ENERGY BALANCE			IN	OUT	DELTA	
Cc=DH/DT	=	BTU/HR F	837887	HOT SIDE			-597937610	-608992883	11,055,273	
Cmin	=	BTU/HR F	93683	COLD SIDE			248897563	259952836	-11,055,273	
Thi-Tci, F	=		145	TOTAL			-349040047	-349040047	0	
Cmin/Cmax	=		0.11							
CALCULATED EFFECTIVENESS FOR SELECTED Tco				e = Cc(Tco-Tci)/ Cmin(Thi-Tci)				0.814		
CALCULATED EFFECTIVENESS FOR SELECTED Tho				e = Ch(Thi-Tho)/ Cmin(Thi-Tci)				0.814		
COOLING LOOP COOLANT PUMP										
FLOW, LBMOLES/HR =			5200							
FLOW LBS/HR			93683.2							
FLOW, GPM			187.1662222							
HEAD, PSI =			10							
EFFICIENCY			0.7							
PUMP POWER, KW =		FLOW(GPM)XHEAD(PSI)/ 2300 =			1.16					
		EFFICIENCY								
PARASITIC POWER FOR COOLING UNIT CONTROLS AND MISC					30		KW			
TOTAL PARASITIC POWER					31.16		KW			

APPENDIX 6

AIR PREHEATER SHEET

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AIR PREHEATERS

3/29/99 15:29									
AIR PREHEATER E-4 AND E-1									
HEAT EXCHANGER E-4									
STATION	22	TEMP, F	834	445.6	DEG C				
HOT SIDE INLET									
GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT		2.016	18.016	16.042	28.01	44.01	32	28.016	27.222
MOLE FLOW, LB MOLES/HR		0	1058.02	0.00	0.00	303.79	377.67	3570.52	5309.99
MOLE FRACTION		0.000	0.199	0	0.000	0.057	0.071	0.672	1.000
SPEC ENTH, BTU/LB MOLE		8,935	-92,130	-15,931	-39,810	-157,094	9,371	9,114	
ENTHALPY, BTU/HR		0	-97,475,644	0	0	-47,723,036	3,538,977	32,542,916	-109,116,787
MASS FLOW LB/HR		0	19,061	0	0	13,370	12,085	100,032	144,548
STATION 23 TEMP, F 446.8 230.4 DEG C									
HOT SIDE EXIT									
GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT		2.016	18.016	16.042	28.01	44.01	32	28.016	27.222
MOLE FLOW, LB MOLES/HR		0.00	1058.02	0.00	0.00	303.79	377.67	3570.52	5309.99
MOLE FRACTION		0.000	0.199	0	0.000	0.057	0.071	0.672	1.000
SPEC ENTH, BTU/LB MOLE		6,217	-95,503	-20,840	-42,637	-161,492	6,392	6,318	
ENTHALPY, BTU/HR		0	-101,043,667	0	0	-49,059,051	2,414,093	22,559,203	-125,129,423
MASS FLOW LB/HR		0	19,061	0	0	13,370	12,085	100,032	144,548
STATION 33 TEMP, F 335.4 168.5 DEG C									
COLD SIDE INLET FROM GT 3RD STAGE COMPRESSOR									
GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT		2.016	18.016	16.042	28.01	44.01	32	28.016	28.789
MOLE FLOW, LB MOLES/HR		0.000	45,700	0.000	0.000	13,709	945,986	3564,601	4569,995476
MOLE FRACTION		0.000	0.010	0.000	0.000	0.003	0.207	0.780	1.000
SPEC ENTH, BTU/LB MOLE		5,439	-96,433	-22,042	-43,427	-162,657	5,570	5,533	
ENTHALPY, BTU/HR		0	-4,406,986	0	0	-2,229,800	5,268,852	19,723,393	18,355,459
MASS FLOW LB/HR		0	823	0	0	603	30,272	99,866	131,564
STATION 34 TEMP, F 816.0 435.6 DEG C									
COLD SIDE EXIT									
GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT		2.016	18.016	16.042	28.01	44.01	32	28.016	28.789
MOLE FLOW, LB MOLES/HR		0.00	45,70	0.00	0.00	13,71	945,99	3564,60	4570,00
MOLE FRACTION		0.000	0.010	0.000	0.000	0.003	0.207	0.780	1.000
SPEC ENTH, BTU/LB MOLE		8,807	-92,293	-16,185	-39,945	-157,309	9,228	8,981	
ENTHALPY, BTU/HR		0	-4,217,795	0	0	-2,156,496	8,729,148	32,013,238	34,368,095
MASS FLOW LB/HR		0	823	0	0	603	30,272	99,866	131,564
Thi, F	=	22	834	23	ASSUMED EFFECTIVENESS, e =		0.964		
Tci, F	=	33	335	34	CALCULATED Tho F =		446.8	=Thi-eCmin/Ch(Thi-Tci)	
Ch=DH/DT	=	BTU/HR F	41344		CALCULATED Tco F		816.0	=Tci+eCmin/Cc(Thi-Tci)	
Cc=DH/DT	=	BTU/HR F	33315		ENERGY BALANCE	IN	OUT	DELTA	
Cmin	=	BTU/HR F	33315		HOT SIDE	-109,116,787	-125,129,423	16,012,636	
Thi-Tci, F	=		499		COLD SIDE	18,355,459	34,368,095	-16,012,636	
Cmin/Cmax	=		0.81		TOTAL	-90,761,328	-90,761,328	0	
CALCULATED EFFECTIVENESS FOR SELECTED Tco e = Cc(Tco-Tci)/ Cmin(Thi-Tci) 0.964									

AIR PREHEATERS

HEAT EXCHANGER E-1										
STATION	19	TEMP, F	1,174	634.2	DEG C					
HOT SIDE INLET										
GAS			H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	27.222
MOLE FLOW, LB MOLES/HR			0.00	1058.02	0.00	0.00	303.79	377.67	3570.52	5309.99
MOLE FRACTION			0.000	0.199	0	0.000	0.057	0.071	0.672	1.000
SPEC ENTH, BTU/LB MOLE			11,342	-88,973	-10,762	-37,215	-152,903	12,113	11,672	
ENTHALPY, BTU/HR			0	-94,135,253	0	0	-46,449,993	4,574,753	41,675,666	-94,334,827
MASS FLOW LB/HR			0	19,061	0	0	13,370	12,085	100,032	144,548
STATION	20	TEMP, F	982.1	527.8	DEG C					
HOT SIDE EXIT										
GAS			H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	27.222
MOLE FLOW, LB MOLES/HR			0.00	1058.02	0.00	0.00	303.79	377.67	3570.52	5309.99
MOLE FRACTION			0.000	0.199	0	0.000	0.057	0.071	0.672	1.000
SPEC ENTH, BTU/LB MOLE			9,980	-90,778	-13,770	-38,692	-155,299	10,553	10,217	
ENTHALPY, BTU/HR			0	-96,045,070	0	0	-47,177,799	3,985,531	36,478,474	-102,758,864
MASS FLOW LB/HR			0	19,061	0	0	13,370	12,085	100,032	144,548
STATION	34	TEMP, F	816.0	435.6	DEG C					
COLD SIDE INLET										
GAS			H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	28.789
MOLE FLOW, LB MOLES/HR			0.00	45.70	0.00	0.00	13.71	945.99	3564.60	4570.00
MOLE FRACTION			0.000	0.010	0.000	0.000	0.003	0.207	0.780	1.000
SPEC ENTH, BTU/LB MOLE			8,807	-92,293	-16,185	-39,945	-157,309	9,228	8,981	
ENTHALPY, BTU/HR			0	-4,217,795	0	0	-2,156,496	8,729,148	32,013,238	34,368,095
MASS FLOW LB/HR			0	823	0	0	603	30,272	99,866	131,564
STATION	35	TEMP, F	1058.0	570.0	DEG C					
COLD SIDE EXIT										
GAS			H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	28.789
MOLE FLOW, LB MOLES/HR			0.00	45.70	0.00	0.00	13.71	945.99	3564.60	4570.00
MOLE FRACTION			0.000	0.010	0.000	0.000	0.003	0.207	0.780	1.000
SPEC ENTH, BTU/LB MOLE			10,518	-90,070	-12,605	-38,111	-154,359	11,168	10,789	
ENTHALPY, BTU/HR			0	-4,116,216	0	0	-2,116,046	10,564,298	38,460,096	42,792,133
MASS FLOW LB/HR			0	823	0	0	603	30,272	99,866	131,564
Thi, F	=	STATION 19	1,174	STATION 20	ASSUMED EFFECTIVENESS, e =		0.677			
Tci, F	=	34	816	35	CALCULATED Tho F =		982.1			=Thi-eCmin/Ch(Thi-Tci)
Ch=DH/DT	=	BTU/HR F	43981		CALCULATED Tco F		1058.0			=Tci+eCmin/Cc(Thi-Tci)
Cc=DH/DT	=	BTU/HR F	34810		ENERGY BALANCE	IN	OUT	DELTA		
Cmin	=	BTU/HR F	34810		HOT SIDE		-94,334,827	-102,758,864	8,424,038	
Thi-Tci, F	=		358		COLD SIDE		34,368,095	42,792,133	-8,424,038	
Cmin/Cmax	=		0.79		TOTAL		-59,966,732	-59,966,732	0	
CALCULATED EFFECTIVENESS FOR SELECTED Tco				e = Cc(Tco-Tci)/ Cmin(Thi-Tci)				0.677		

APPENDIX 7

TURBINE SHEET

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Page 1

3/29/99 15:29		TURBINE EXPANDER/GENERATOR							
P1,= FIRST TURBINE INLET PRESSURE, PSIA =		231.68			IST STAGE	TURBINE EFFICIENCY =		0.899	
P2 = IST STAGE OUTLET PRESSURE		25.81			TURBINE INLET LOSSES,%			1	
P3 = NOMINAL 2ND STAGE INLET PRESSURE		23.00			IST STAGE TURBINE EXIT DIFFUSER LOSS, %=			7.00	
P3'= ACTUAL 2ND STAGE INLET		22.77			EXPANSION RATIO IST TURBINE =			8.98	
P4= 2ND STAGE OUTLET PRESSURE, PSIA =		16.32			2ND STAGE TURBINE EFFICIENCY=			0.922	
P5 = EXIT PRESSURE FROM 2ND TURBINE		15.5			2ND STAGE TURBINE EXIT DIFFUSER LOSS, %			5.00	
					EXPANSION RATIO SECOND TURBINE =			1.40	
STATION	36	TEMP, F	1526.0	830.0	DEG C				
FIRST STAGE TURBINE INLET									
	GAS	H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT		2.016	18.016	16.042	28.01	44.01	32	28.016	28.789
MOLE FLOW, LB MOLES/HR		0.000	45.700	0.000	0.000	13.709	945.986	3564.601	4570.00
MOLE FRACTION		0.000	0.010	0.000	0.000	0.003	0.207	0.780	1.000
SPEC ENTH, BTU/LB MOLE		13,887	-85,489	-4,685	-34,418	-148,322	15,053	14,430	
ENTHALPY, BTU/HR		0	-3,906,825	0	0	-2,033,291	14,239,562	51,435,839	59,735,286
MASS FLOW LB/HR		0	823	0	0	603	30,272	99,866	131,564
					FIRST STAGE TURBINE WORK, Wt, (ADIABATIC)				
					=(wt/m) R Tti (K/K-1)(1-(Po/Pi)^(k-1/k))				
TURBINE EXIT TEMP					R = 1.9838BTU /lb mole R				
To = Ti (Po/Pi)^(K-1/K) +(Wti-Wta-HL)/wtCp -460					Wtideal = BTU/HR,kW		30,503,125	8,937	
					wt = LB/HR		131,564		
TEMP IN , F		1526			m = molecular wt lbs/lb mole		28.789		
TEMP OUT , F		781.3			Po/Pi = pressure ratio		0.1114		
					k = constant for gas, Cp/Cv		1.3281		
					Cp , BTU/LB F		0.271		
					k-1/k		0.247		
					n = ADIABATIC efficiency		0.899		
					Wtactual= BTU/HR, kW		27,422,309	8,035	
STATION	37	TEMP, F	755.6	402.0	DEG C	SECOND STAGE OUTLET PRESSURE, PSIA =		16.31578947	
FIRST STAGE TURBINE EXIT									
	GAS	H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT		2.016	18.016	16.042	28.01	44.01	32	28.016	28.789
MOLE FLOW, LB MOLES/HR		0.000	45.700	0.000	0.000	13.709	945.986	3564.601	4570.00
MOLE FRACTION		0.000	0.010	0.000	0.000	0.003	0.207	0.780	1.000
SPEC ENTH, BTU/LB MOLE		8,382	-92,833	-17,016	-40,394	-158,023	8,752	8,537	
ENTHALPY, BTU/HR		0	-4,242,473	0	0	-2,166,283	8,279,599	30,432,001	32,302,844
MASS FLOW LB/HR		0	823	0	0	603	30,272	99,866	131,564
ENTHALPY IN , BTU/HR		36	5.9735E+07			HEAT BALANCE			
ENTHALPY OUT, BTU/HR		37+TURBINE POWER	5.9725E+07			LOGIC FOR ITERATION		FALSE	FALSE
HEAT LOSS, BTU/HR			10000			ITERATION INCREMENT		226.87	-277.32
NET ENTHALPY, IN - OUT , BTU/HR			133			TEMP CHANGE =		-62.45	

GTTURBINE

STATION	38	TEMP, F	1220.0	660.0	DEG C					
SECOND STAGE TURBINE INLET										
	GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	28.789
MOLE FLOW, LB MOLES/HR			0.000	45.700	0.000	0.000	13.709	945.986	3564.601	4570.00
MOLE FRACTION			0.000	0.010	0.000	0.000	0.003	0.207	0.780	1.000
SPEC ENTH, BTU/LB MOLE			11,674	-88,527	-10,001	-36,852	-152,312	12,495	12,029	
ENTHALPY, BTU/HR			0	-4,045,669	0	0	-2,087,992	11,820,365	42,879,572	48,566,276
MASS FLOW LB/HR			0	823	0	0	603	30,272	99,866	131,564
SECOND STAGE TURBINE WORK, Wt, (ADIABATIC)										
=(wt/m) R Tti (K/K-1)(1-(Po/Pi)^(k-1/k))										
R = 1.9838BTU /lb mole R										
To = Ti (Po/Pi)^(K-1/K) +(Wti-Wta-HL)/wtCp -460										
Wtideal = BTU/HR,kW										
wt = LB/HR										
m = molecular wt lbs/lb mole										
Po/Pi = pressure ratio										
k = constant for gas, Cp/Cv										
Cp , BTU/LB F										
k-1/k										
n = ADIABATIC efficiency										
Wtactual= BTU/HR, kW										
STATION	39	TEMP, F	1094.0	590.0	DEG C					
SECOND STAGE TURBINE EXIT										
	GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	28.789
MOLE FLOW, LB MOLES/HR			0.000	45.700	0.000	0.000	13.709	945.986	3564.601	4570.00
MOLE FRACTION			0.000	0.010	0.000	0.000	0.003	0.207	0.780	1.000
SPEC ENTH, BTU/LB MOLE			10,774	-89,731	-12,040	-37,833	-153,908	11,461	11,063	
ENTHALPY, BTU/HR			0	-4,100,705	0	0	-2,109,869	10,841,890	39,435,847	44,067,163
MASS FLOW LB/HR			0	823	0	0	603	30,272	99,866	131,564
ENTHALPY IN , BTU/HR			4.8566E+07			HEAT BALANCE				
ENTHALPY OUT, BTU/HR			4.8556E+07			LOGIC FOR ITERATION			FALSE	FALSE
HEAT LOSS, BTU/HR			10000			ITERATION INCREMENT			2.46	-3.21
NET ENTHALPY, IN - OUT , BTU/HR			11			TEMP CHANGE =			-0.75	
TURBINE GENERATOR WORK SUMMARY										
WORK, KW										
FIRST STAGE COMPRESSOR										
SECOND STAGE COMPRESSOR										
THIRD STAGE COMPRESSOR										
FIRST STAGE TURBINE										
SECOND STAGE TURBINE										
NET POWER										
GENERATOR EFFICIENCY										
NET AC POWER										

APPENDIX 8

BURNER SHEET

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3/29/99 15:29			BURNER CATALYTIC BURNER , E-8 PLUS E-9							
STATION	13	TEMP, F	1,170.00	632.2 DEG C						
STACK ANODE OUT FLOW TO BURNER										
GAS			H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.010	44.010	32.000	28.016	30.158
MOLE FLOW, LB MOLES/HR			156.955	855.365	0.000	92.911	1083.940	0.000	5.920	2195.09
MOLE FRACTION			0.072	0.390	0.000	0.042	0.494	0.000	0.003	1.000
SPEC ENTH, BTU/LB MOLE			11,316	-89,008	-10,821	-37,243	-152,949	12,083	11,644	
ENTHALPY, BTU/HR			1,776,111	-76,134,278	0	-3,460,291	-165,787,892	0	68,934	-243,537,416
MASS FLOW LB/HR			316	15,410	0	2,602	47,704	0	166	66,199
STATION	39	TEMP, F	1,094.03	590.0 DEG C						
2ND STAGE TURBINE EXIT										
GAS			H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.010	44.010	32.000	28.016	28.789
MOLE FLOW, LB MOLES/HR			0.000	45.700	0.000	0.000	13.709	945.986	3564.601	4570.00
MOLE FRACTION			0.000	0.010	0.000	0.000	0.003	0.207	0.780	1.000
SPEC ENTH, BTU/LB MOLE			10,774	-89,731	-12,040	-37,833	-153,908	11,461	11,063	
ENTHALPY, BTU/HR			0	-4,100,705	0	0	-2,109,869	10,841,890	39,435,847	44,067,163
MASS FLOW LB/HR			0	823	0	0	603	30,272	99,866	131,564
STATION	14	TEMP, F	1,124.5	606.9 DEG C						
MIXED ANODE EXHAUST AND TURBINE EXHAUST (BEFORE BURNING										
GAS			H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.010	44.010	32.000	28.016	29.233
MOLE FLOW, LB MOLES/HR			156.955	901.065	0.000	92.911	1097.649	945.986	3570.521	6765.09
MOLE FRACTION			0.023	0.133	0.000	0.014	0.162	0.140	0.528	1.000
SPEC ENTH, BTU/LB MOLE			10,991	-89,442	-11,555	-37,597	-153,525	11,710	11,295	
ENTHALPY, BTU/HR			1,725,139	-80,593,368	0	-3,493,173	-168,516,677	11,077,348	40,330,732	-199,470,000
MASS FLOW LB/HR			316	16,234	0	2,602	48,308	30,272	100,032	197,763
ENTHALPY IN , BTU/HR				-1.9947E+08			HEAT BALANCE			
ENTHALPY OUT, BTU/HR				-1.9947E+08			LOGIC FOR ITERATION		FALSE	FALSE
HEAT LOSS, BTU/HR				0			ITERATION INCREMENT		1.49	-0.01
NET ENTHALPY, IN - OUT , BTU/HR				-253			TEMP CHANGE =		1.48	
STATION	15	TEMP, F	1,585.3	863.0 DEG C						
CATALYTIC BURNER EXHAUST										
GAS			H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.010	44.010	32.000	28.016	29.783
MOLE FLOW, LB MOLES/HR			0.000	1058.020	0.000	0.000	1190.560	821.053	3570.521	6640.15
MOLE FRACTION			0.000	0.159	0.000	0.000	0.179	0.124	0.538	1.000
SPEC ENTH, BTU/LB MOLE			14,321	-84,881	-3,601	-33,938	-147,532	15,554	14,903	
ENTHALPY, BTU/HR			0	-89,806,144	0	0	-175,645,886	12,771,023	53,211,047	-199,469,960
MASS FLOW LB/HR			0	19,061	0	0	52,397	26,274	100,032	197,763
ENTHALPY IN , BTU/HR				-1.9947E+08			HEAT BALANCE			
ENTHALPY OUT, BTU/HR				-1.9947E+08			LOGIC FOR ITERATION		FALSE	FALSE
HEAT LOSS, BTU/HR				0			ITERATION INCREMENT		0.345	-0.02
NET ENTHALPY, IN - OUT , BTU/HR				-40			TEMP CHANGE =		0.325	

		HEAT EXCHANGER E-8															
STATION	15	TEMP, F	1,585	863.0	DEG C												
HOT SIDE INLET(CAT BURNER EXIT)																	
GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL								
MOLECULAR WEIGHT		2.016	18.016	16.042	28.01	44.01	32	28.016	29.783								
MOLE FLOW, LB MOLES/HR		0.000	1058.020	0.000	0.000	1190.560	821.053	3570.521	6640.153769								
MOLE FRACTION		0.000	0.159	0	0.000	0.179	0.124	0.538	1.000								
SPEC ENTH, BTU/LB MOLE		14,321	-84,881	-3,601	-33,938	-147,532	15,554	14,903									
ENTHALPY, BTU/HR		0	-89,806,144	0	0	-175,645,886	12,771,023	53,211,047	-199,469,960								
MASS FLOW LB/HR		0	19,061	0	0	52,397	26,274	100,032	197,763								
STATION	16	TEMP, F	1,309.4	709.7	DEG C												
HOT SIDE EXIT																	
GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL								
MOLECULAR WEIGHT		2.016	18.016	16.042	28.01	44.01	32	28.016	29.783								
MOLE FLOW, LB MOLES/HR		0.00	1058.02	0.00	0.00	1190.56	821.05	3570.52	6640.15								
MOLE FRACTION		0.000	0.159	0	0.000	0.179	0.124	0.538	1.000								
SPEC ENTH, BTU/LB MOLE		12,316	-87,655	-8,498	-36,148	-151,162	13,236	12,723									
ENTHALPY, BTU/HR		0	-92,741,245	0	0	-179,967,499	10,867,808	45,427,822	-216,413,113								
MASS FLOW LB/HR		0	19,061	0	0	52,397	26,274	100,032	197,763								
STATION	35	TEMP, F	1058.0	570.0	DEG C												
COLD SIDE INLET FROM E-1 EXIT																	
GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL								
MOLECULAR WEIGHT		2.016	18.016	16.042	28.01	44.01	32	28.016	28.789								
MOLE FLOW, LB MOLES/HR		0.000	45.700	0.000	0.000	13.709	945.986	3564.601	4570.00								
MOLE FRACTION		0.000	0.010	0.000	0.000	0.003	0.207	0.780	1.000								
SPEC ENTH, BTU/LB MOLE		10,518	-90,070	-12,605	-38,111	-154,359	11,168	10,789									
ENTHALPY, BTU/HR		0	-4,116,216	0	0	-2,116,046	10,564,298	38,460,096	42,792,133								
MASS FLOW LB/HR		0	823	0	0	603	30,272	99,866	131,564								
STATION	36	TEMP, F	1526.0	830.0	DEG C												
COLD SIDE EXIT																	
GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL								
MOLECULAR WEIGHT		2.016	18.016	16.042	28.01	44.01	32	28.016	28.789								
MOLE FLOW, LB MOLES/HR		0.000	45.700	0.000	0.000	13.709	945.986	3564.601	4570.00								
MOLE FRACTION		0.000	0.010	0.000	0.000	0.003	0.207	0.780	1.000								
SPEC ENTH, BTU/LB MOLE		13,887	-85,489	-4,685	-34,418	-148,322	15,053	14,430									
ENTHALPY, BTU/HR		0	-3,906,825	0	0	-2,033,291	14,239,562	51,435,839	59,735,286								
MASS FLOW LB/HR		0	823	0	0	603	30,272	99,866	131,564								
		STATION	STATION	ASSUMED EFFECTIVENESS, e =			0.887										
Thi, F	=	15	1,585	16	CALCULATED Tho F =			1309.4		=Thi-eCmin/Ch(Thi-Tci)							
Tci, F	=	37	1,058	38	CALCULATED Tco F			1526.0		=Tci+eCmin/Cc(Thi-Tci)							
Ch=DH/DT	=	BTU/HR F	61405		ENERGY BALANCE			IN		OUT		DELTA					
Cc=DH/DT	=	BTU/HR F	36203		HOT SIDE			-199,469,960		-216,413,113		16,943,153					
Cmin	=	BTU/HR F	36203		COLD SIDE			42,792,133		59,735,286		-16,943,153					
Thi-Tci, F	=		527		TOTAL			-156,677,827		-156,677,827		0					
Cmin/Cmax	=		0.59														
CALCULATED EFFECTIVENESS FOR SELECTED Tco				e = Cc(Tco-Tci)/ Cmin(Thi-Tci)				0.887									

HEAT EXCHANGER E-9										
STATION	16	TEMP, F	1,309	709.7	DEG C					
HOT SIDE INLET										
	GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	29.783
MOLE FLOW, LB MOLES/HR			0.000	1058.020	0.000	0.000	1190.560	821.053	3570.521	6640.153769
MOLE FRACTION			0.000	0.159	0	0.000	0.179	0.124	0.538	1.000
SPEC ENTH, BTU/LB MOLE			12,316	-87,655	-8,498	-36,148	-151,162	13,236	12,723	
ENTHALPY, BTU/HR			0	-92,741,245	0	0	-179,967,499	10,867,808	45,427,822	-216,413,113
MASS FLOW LB/HR			0	19,061	0	0	52,397	26,274	100,032	197,763
STATION	17	TEMP, F	1,035.8	557.6	DEG C					
HOT SIDE EXIT										
	GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	29.783
MOLE FLOW, LB MOLES/HR			0.000	1058.020	0.000	0.000	1190.560	821.053	3570.521	6640.153769
MOLE FRACTION			0.000	0.159	0	0.000	0.179	0.124	0.538	1.000
SPEC ENTH, BTU/LB MOLE			10,360	-90,279	-12,950	-38,282	-154,636	10,987	10,621	
ENTHALPY, BTU/HR			0	-95,516,844	0	0	-184,102,916	9,020,778	37,922,435	-232,676,546
MASS FLOW LB/HR			0	19,061	0	0	52,397	26,274	100,032	197,763
STATION	37	TEMP, F	755.6	402.0	DEG C					
COLD SIDE INLET FROM 1ST STAGE TURBINE										
	GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	28.789
MOLE FLOW, LB MOLES/HR			0.000	45.700	0.000	0.000	13.709	945.986	3564.601	4570.00
MOLE FRACTION			0.000	0.010	0.000	0.000	0.003	0.207	0.780	1.000
SPEC ENTH, BTU/LB MOLE			8,382	-92,833	-17,016	-40,394	-158,023	8,752	8,537	
ENTHALPY, BTU/HR			0	-4,242,473	0	0	-2,166,283	8,279,599	30,432,001	32,302,844
MASS FLOW LB/HR			0	823	0	0	603	30,272	99,866	131,564
STATION	38	TEMP, F	1220.0	660.0	DEG C					
COLD SIDE EXIT										
	GAS		H2	H2O	CH4	CO	CO2	O2	N2	TOTAL
MOLECULAR WEIGHT			2.016	18.016	16.042	28.01	44.01	32	28.016	28.789
MOLE FLOW, LB MOLES/HR			0.000	45.700	0.000	0.000	13.709	945.986	3564.601	4570.00
MOLE FRACTION			0.000	0.010	0.000	0.000	0.003	0.207	0.780	1.000
SPEC ENTH, BTU/LB MOLE			11,674	-88,527	-10,001	-36,852	-152,312	12,495	12,029	
ENTHALPY, BTU/HR			0	-4,045,669	0	0	-2,087,992	11,820,365	42,879,572	48,566,276
MASS FLOW LB/HR			0	823	0	0	603	30,272	99,866	131,564
Thi, F	=	STATION 16	1,309	STATION 17	ASSUMED EFFECTIVENESS, e =			0.839		
Tci, F	=	39	756	40	CALCULATED Tho F =			1035.8	=Thi-eCmin/Ch(Thi-Tci)	
Ch=DH/DT	=	BTU/HR F	59432		CALCULATED Tco F			1220.0	=Tci+eCmin/Cc(Thi-Tci)	
Cc=DH/DT	=	BTU/HR F	35017		ENERGY BALANCE			IN	OUT	DELTA
Cmin	=	BTU/HR F	35017		HOT SIDE			-216,413,113	-232,676,546	16,263,433
Thi-Tci, F	=		554		COLD SIDE			32,302,844	48,566,276	-16,263,433
Cmin/Cmax	=		0.59		TOTAL			-184,110,270	-184,110,270	0
CALCULATED EFFECTIVENESS FOR SELECTED Tco					e = Cc(Tco-Tci)/ Cmin(Thi-Tci)			0.839		
CALCULATED EFFECTIVENESS FOR SELECTED Tho					e=Ch(Thi-Tho)/Cmin(Thi-Tci)			0.839		

APPENDIX 9

SUMMARY SHEET FOR EXAMPLE CASE

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SUMMARY

3/29/99 15:29		HIGH EFFICIENCY POWER PLANT PROCESS SUMMARY									
EXCEL 97 WORK BOOK PREPARED BY R.A.Sanderson P.E.				CASE	1						
REVISION-- MARCH 19,1999											
		FIXED									
INPUT PARAMETERS		REFERENCE	+/- DEVIATION	OUTPUT CHARACTERISTICS						REQUIREMENT	
CH4 INPUT, MOLES/HR	=	296.00	0	NET AC POWER, KW		=	20,411	20,000.00			
PROCESS AIR FLOW, MOLES/HR	=	4570	0	NET AC LHV EFFICIENCY,%		=	71.08	> 70			
AMBIENT AIR TEMP F	=	59	0	PARASITIC LOSSES, KW		=	31				
WATER VAPOR PRESS, PSIA	=	0.147	0	GROSS AC POWER		=	20442				
FUEL LOWER HEATING VALUE,BTU/SCF	=	872.5	N/A	PLANT EXHAUST TEMP, F		=	208.6				
CELL STACK				FUEL CELLS							
CELL VOLTAGE	=	0.817	0	FUEL CELL AC OUTPUT, KW		=	17002				
CURRENT DENSITY, AMPS/CM^2	=	0.16	0	FUEL CELL DC POWER, KW		=	17619				
NUMBER STACKS		48	N/A	FUEL UTILIZATION,%		=	78.0				
NUMBER CELLS/STACK	=	360	N/A	CATHODE EXIT TEMP, DEG F		=	1173.6	1180 MAX			
CELL AREA, M^2	=	0.78	N/A	CATHODE INLET TEMP, DEG F		=	1035.8	1030 MIN			
ANODE EXIT TEMP DEG C	=	1170	0	FUEL CELL OXYGEN UTILIZATION, %		=	54.0				
METHANE CONVERSION,%	=	100	N/A	FUEL CELL INLET O2 FRACTION		=	0.1236				
STACK HEAT LOSS, MBTU/HR	=	700,000	0	OUTLET O2 FRACTION		=	0.0711				
INVERTER EFFICIENCY,%	=	96.5	0	FUEL CELL INLET CO2 FRACTION		=	0.1793				
PARASITIC PWR, CONTROLS&MISC,KW	=	10	0	OUTLET CO2 FRACTION		=	0.0572				
FUEL PROCESSING	=			GAS TURBINE							
PREREFORM EXIT PRES, PSIA	=	17	0	TURBINE GEN NETPOWER AC, KW		=	3440				
PREREFORM INLET TEMP, C	=	752	E-3 EXIT	HI TURBINE INLET TEMPERATURE, F		=	1526				
STEAM/CH4 RATIO	=	1.5	0	LOW TURBINE INLET TEMPERATURE, F		=	1220				
RU TEMP, F	=	1100	0	HIGH TURBINE POWER, KW		=	8035				
GAS TURBINE			0	LOW TURBINE POWER, KW		=	1315				
GAS TURBINE COMP EFF,%	=	85	0	1ST STAGE COMPRESSOR WORK, KW		=	1827				
GAS TURBINE TURB EFF,%	=	90	0	2ND STAGE COMPRESSOR WORK, KW		=	2002				
GT COMPRESSOR PRESS RATIOS	=	2.63	0	3RD STAGE COMPRESSOR WORK,KW		=	2011				
EACH INTERCOOLER PRESS LOSS, PSI	=	1	0								
COMPRESSOR DUCT AND INLET LOSSES, %		3	0								
PRESSURE LOSS, COMP OUT TO TURB IN,PSI	=	4	0	PROCESS WATER TOTAL, GAL/MIN		=	16				
PRESSURE LOSS, E-9 ,PSI	=	1	0								
IST TURBINE STAGE OUTLET PRESSURE, PSIA	=	23	0								
TURBINE EXIT DIFFUSER LOSSES, %, 1ST/2N	=	7/5	0								
GENERATOR EFFICIENCY,%	=	98	0								
TURBINE STAGE HEAT LOSS, BTU/HR	=	10000	0								
HEAT EXCHANGER SUMMARY		REFERENCE DEG F	+/- DEVIATION	REQUIRED HEAT EXCHANGER EFFECTIVENESS							
E-1 SECOND STAGE AIR HEATER	COLD SIDE EXIT	1058	0	E-1	=	0.68					
E-2 PROCESSED FUEL REHEATER	COLD SIDE EXIT	932	0	E-2	=	0.86					
E-3 FUEL /STEAM HEATER	COLD SIDE EXIT	752	0	E-3	=	0.75					
E-4 FIRST STAGE AIR HEATER	COLD SIDE EXIT	816	0	E-4	=	0.96					
E-5 BOILER	COLD SIDE EXIT	239	0	E-5	=	0.81					
E-6 NATURAL GAS HEATER	COLD SIDE EXIT	212	0	E-6	=	0.81					
E-7 FEEDWATER HEATER	COLD SIDE EXIT	203	0	E-7	=	0.80					
E-8 TURBINE INLET HEATER	COLD SIDE EXIT	1526	0	E-8	=	0.89					
E-9 TURBINE INLET REHEATER	COLD SIDE EXIT	1220	0	E-9	=	0.84					
E-10 INTERCOOLER	HOT SIDE EXIT	122	0	E-10	=	0.79					
E-11 INTERCOOLER	HOT SIDE EXIT	122	0	E-11	=	0.86					
E-12 COOLING TOWER	COLD SIDE EXIT	86	0	E-12	=	0.81					

SUMMARY

3/29/99 15:29				HIGH EFFICIENCY POWER PLANT PROCESS SUMMARY									
						PROCESS CONDITIONS							
						CASE	1						
	STATION	TEMP	PRESS	ENTHALPY	COMPOSITION, VOLUME FRACTION							FLOW	FLOW
NO	DESCRIPTION	DEG F	PSIA	BTU/HR	H2	H2O	CH4	CO	CO2	O2	N2	MOLES/HR	LBS/HR
1	NATURAL GAS IN	59.0	29.70	-7,949,462	0.0000	0.0000	0.9600	0.0000	0.0200	0.0000	0.0200	296.00	4,985
2	DESULFUR NAT GAS	15.0	24.70	-7,949,462	0.0000	0.0000	0.9600	0.0000	0.0200	0.0000	0.0200	296.00	4,985
3	HEATED NAT GAS	212.0	24.60	-7,551,780	0.0000	0.0000	0.9600	0.0000	0.0200	0.0000	0.0200	296.00	4,985
4	FUEL/STEAM/RECYCLE	233.7	19.20	-51,454,583	0.0012	0.5991	0.3835	0.0000	0.0083	0.0000	0.0080	752.02	13,187
5	PREREFORMER INLET	752.0	19.00	-47,683,213	0.0012	0.5991	0.3835	0.0000	0.0083	0.0000	0.0080	752.02	13,187
6	PREREFORMER EXIT	611.3	17.00	-47,682,947	0.0748	0.5448	0.3539	0.0001	0.0264	0.0000	0.0078	774.58	13,187
7	FUEL/STEAM TO E-2	611.3	16.71	-46,948,630	0.0742	0.5406	0.3511	0.0001	0.0262	0.0000	0.0077	768.57	12,984
8	FUEL/STEAM FROM E-2	932.0	16.61	-44,362,077	0.0742	0.5406	0.3511	0.0001	0.0262	0.0000	0.0077	768.57	12,984
9	FUEL TO STACK Rus	932.0	16.60	-44,362,077	0.0746	0.5357	0.3555	0.0001	0.0264	0.0000	0.0078	768.57	12,984
10	FUEL FROM Rus	1100.0	15.60	-30,524,258	0.5111	0.2058	0.1256	0.0734	0.0784	0.0000	0.0057	1,045.61	12,984
11	FUEL TO ANODES	1100.0	15.60	-30,524,258	0.5111	0.2058	0.1256	0.0734	0.0784	0.0000	0.0057	1,045.61	12,984
12	FUEL FROM ANODES	1170.0	15.52	-243,537,416	0.0715	0.3897	0.0000	0.0423	0.4938	0.0000	0.0027	2,195.09	66,199
13	FUEL TO BURNER	1170.0	15.50	-243,537,416	0.0715	0.3897	0.0000	0.0423	0.4938	0.0000	0.0027	2,195.09	66,199
14	BURNER PREMIX	1124.5	15.45	-199,470,000	0.0232	0.1332	0.0000	0.0137	0.1623	0.1398	0.5278	6,765.09	197,763
15	BURNER EXHAUST	1585.3	15.35	-199,469,960	0.0000	0.1593	0.0000	0.0000	0.1793	0.1236	0.5377	6,640.15	197,763
16	EXHAUST FROM E-8	1309.4	15.27	-216,413,113	0.0000	0.1593	0.0000	0.0000	0.1793	0.1236	0.5377	6,640.15	197,763
17	CATHODE INLET	1035.8	15.19	-232,676,546	0.0000	0.1593	0.0000	0.0000	0.1793	0.1236	0.5377	6,640.15	197,763
18	CATHODE EXHAUST	1173.6	15.11	-94,334,827	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
19	EXHAUST TO E-1	1173.6	15.15	-94,334,827	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
20	EXHAUST TO E-2	982.1	15.05	-102,758,864	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
21	EXHAUST TO E-3	922.3	15.00	-105,345,417	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
22	EXHAUST TO E-4	834.1	14.95	-109,116,787	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
23	EXHAUST TO E-5	446.8	14.85	-125,129,423	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
24	EXHAUST TO E-6	248.1	14.80	-133,024,387	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
25	EXHAUST TO E-7	238.0	14.75	-133,422,069	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
26	EXHAUST FROM SYSTEM	208.6	14.70	-134,573,876	0.0000	0.1993	0.0000	0.0000	0.0572	0.0711	0.6724	5,309.99	144,548
27	PROCESS AIR IN	59.0	14.70	9,478,849	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
28	1ST STAGE COMP IN	59.0	14.45	9,478,849	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
29	1ST STAGE COMP OUT	253.7	38.01	15,714,514	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
30	2ND STAGE COMP IN	122.0	35.87	11,490,654	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
31	2ND STAGE COMP OUT	334.3	94.35	18,322,460	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
32	3RD STAGE COMP IN	122.0	90.52	11,490,654	0.0000	0.0583	0.0000	0.0000	0.0091	0.6273	2.3636	4,570.00	131,564
33	3RD STAGE COMP OUT	335.4	238.06	18,355,459	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
34	PREHEATER E-4 OUT	816.0	236.73	34,368,095	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
35	PREHEATER E-1 OUT	1058.0	235.39	42,792,133	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
36	1ST TURBINE IN	1526.0	231.68	59,735,286	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
37	1ST TURBINE OUT	755.6	25.81	32,302,844	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
38	2ND TURBINE IN	1220.0	23.00	48,566,276	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564
39	2ND TURBINE OUT	1094.0	15.50	44,067,163	0.0000	0.0100	0.0000	0.0000	0.0030	0.2070	0.7800	4,570.00	131,564

SUMMARY

[illegible]

High Efficiency Fossil Power Plant (HEFPP)

Conceptualization Program

Final Report

Start Date: May 28, 1998
End Date: October 31, 1999

Principal Author: J.L. Justice

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Abstract

This study confirms the feasibility of a natural gas fueled, 20 MW M-C Power integrated pressurized molten carbonate fuel cell combined in a topping cycle with a gas turbine generator plant. The high efficiency fossil power plant (HEFPP) concept has a 70 percent efficiency on a LHV basis. The study confirms the HEFPP has a cost advantage on a cost of electricity basis over the gas turbine based combined cycle plants in the 20 MW size range. The study also identifies the areas of further development required for the fuel cell, gas turbine generator, cathode blower, inverter, and power module vessel. The HEFPP concept offers an environmentally friendly power plant with minuscule emission levels when compared with the combined cycle power plant.

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1. EXECUTIVE SUMMARY

This study shows that Pressurized Molten Carbonate Fuel Cell technology can be combined with gas turbine technology to produce a high efficiency power plant with 70 percent low heating value efficiency when utilizing natural gas as the fuel in the 20 MW plant size. The study has also proven that the high efficiency fossil power plant (HEFPP) concept produces a lower cost of electricity upon comparison with a similar size combined cycle plant. In the specific study case, the HEFPP concept cost to produce electricity was 6.3 cents per kWh as compared to the 7.1 cents per kWh for the gas turbine combined cycle case. The HEFPP concept also produces a much more environmentally friendly power plant with very low emissions of NO_x, CO₂, and SO_x, when compared to the gas turbine combined cycle case.

This study acknowledges that investments of capital and time in the HEFPP components must be made to drive the component costs down and to eliminate or improve specific problem areas. Specific components requiring further investment and development time include fuel cell stacks, gas turbine generator, recycle blower, power module vessel, and inverter. It should be noted that significant work to validate the process concept at smaller scale has already been completed or is in progress based on DOE funding for M-C Power's Process Design and Improvement Project under contract DE-FC21-95MC30133 and the ongoing MCFC power plant testing at Miramar.

2. INTRODUCTION

M-C Power Corporation was the prime contractor for the High Efficiency Fossil Power Plant concept study with Bechtel National, Inc., Rolls-Royce Allison, and Institute of Gas Technology (IGT) as the subcontractors.

The High Efficiency Fossil Power Plant (HEFPP) proposed concept consists of several integrated molten carbonate fuel cells (MCFC) serving as a topping cycle to a variable speed gas turbine driving an air compressor and electric alternator. The power plant will produce approximately 20 MW of power when natural gas is used as the fuel source. The gas turbine compressor section provides compressed air at a pressure ratio of 6 to the MCFC. The MCFC provides hot exhaust gas for expansion in the gas turbine, which drives both the compressor section and the alternator. DC power from the fuel cells is run through inverters where the voltage is converted. The AC power is elevated to 4160VAC. The power from the gas turbine alternator is directed through a power conditioner whereby the frequency is corrected to 60 Hz and the voltage is changed to 4160VAC. It should be noted that the design concept for the HEFPP is based on M-C Power's 450 kW MCFC power plant.

The main goal of the HEFPP study is to have a minimum overall plant efficiency of 70 percent excluding the potential for excess steam generation for cogeneration purposes from the high temperature gas turbine exhaust. Other goals are to achieve low environmental emissions and have high reliability with no major technical barriers for commercialization by the year 2010. An economic goal is to have the HEFPP plant to achieve a reduction by 10 to 20% for the cost of electricity price below a conventional gas turbine combined cycle plant.

This study identifies the design and development issues associated with the proposed HEFPP, confirms the design parameters of the various plant components, provides the process description and plant performance summary, makes economic projections for the plant capital and operating costs, and establishes the plant life cycle costs and cost of electricity. The HEFPP concept has been compared to a similar size gas fired gas turbine based combined cycle power plant with similar environmental emissions to assess the economic viability of the project.

3. RESULTS AND DISCUSSTION

3.A Concept Verification

This report section addresses the design and development issues and design parameters for the HEFPP concept by looking at the fuel cell, gas turbine generator and the balance of plant. Optimization of the cycle operating pressure and the number of spare trains are also addressed in this section.

3.A.1. Design and Development Issues

3.A.1.a Fuel Cell Stack

The molten carbonate fuel cell (MCFC) stacks in the High Efficiency Fossil Power Plant (HEFPP) study are the next generation of the full-area commercial-size MCFC stacks. Verification of the new generation has begun at M-C Power's pressurized test facility with a 75 kW demonstration stack at Miramar Marine Corps Air Station near San Diego. The development plan for cell package engineering to meet the required commercialization cost, power density and life goals for the year 2000 to 2006 time frame is in place as part of the present DOE funded Process Design Improvement (PDI) program (DOE Contract No. DE-FC21-95MC30133). The following design and development issues are being considered for successful implementation of the MCFC for the high-efficiency power plants:

Endurance (stack life) under HEFPP operating conditions.

The projected stack life under HEFPP operating conditions is 40,000 hours. There are no physical laws or technical data that states that this goal is beyond reach and cannot be achieved by 2008-2010.

The issues limiting stack life include:

- (a) NiO cathode dissolution in the electrolyte and electrical shorting by Ni deposition in the matrix,
- (b) electrolyte loss by corrosion of the separator plate and metal current collectors, and electrolyte loss by evaporation,
- (c) matrix ability to retain electrolyte due to change of matrix pore structure and phase composition,
- (d) matrix crack-resistance,
- (e) separator plate corrosion,
- (f) wet-seal corrosion protection and efficiency of sealing.

NiO cathode dissolution in electrolyte and electrical shorting by Ni deposition in the matrix

The issue of cathode dissolution and Ni-shorting through the matrix is being addressed by using additives, such as CaO or CaCO₃, to the Li-Na carbonate electrolyte. A stabilized cathode also reduces the potential for shorting. The use of LiNa significantly reduces NiO cathode

dissolution. The Dutch experimental data suggests that MCFC stacks can operate between 30,000-50,000 hours before the shorting occurs. An Institute of Gas Technology (IGT) bench-scale fuel cell had operated for >17,000 hours with no shorting. Additional development work and testing are planned to confirm the stack life can be increased to the 40,000-hour goal.

Electrolyte losses

To abate electrolyte loss by corrosion of the separator plate and current collectors, material selection for the cell cathode side is being analyzed and design modifications are being considered. The 310S has better corrosion resistance than 316L; however, the analysis has shown that 310S stainless steel consumes more electrolyte for corrosion than 316L. The purpose of the design modifications is to reduce the metal surface area exposed to electrolyte and, by doing that, to reduce electrolyte losses by corrosion. Considerable reduction of the electrolyte loss by evaporation has already been achieved by changing from Li-K to Li-Na electrolytes per M-C Power and IGT bench scale testing. The 310 S material will be tested at Miramar in a 75 kW stack as part of the Process Design Improvement Project.

Matrix pore structure and phase composition change

The ability of the matrix to retain electrolyte depends on LiAlO_2 phase stability and particle growth. It was found that $\alpha\text{-LiAlO}_2$ is more stable than $\gamma\text{-LiAlO}_2$ at the normal operating temperature of $\sim 650^\circ\text{C}$. However, recent IHI experience has proven that $\gamma\text{-LiAlO}_2$ can be stable if very high purity constituents and very small uniform particles are used, which increases stack life to 40,000 hours. For its previous stacks, M-C Power has used $\gamma\text{-LiAlO}_2$ matrices; but both approaches are being evaluated for future matrix material.

Matrix crack-resistance

The matrices developed by M-C Power confirmed their ability to withstand multiple thermal cycles. A number of bench-scale cell tests, which have been performed at IGT, showed that cell performance, sealing, as well as matrix integrity were not compromised after 6-8 thermal cycles. Full area fuel cell thermal cycling tests are planned at Miramar as part of the PDI project.

Separator plate corrosion

Stainless steels 316L and 310S are the current choices for MCFC separator plate material. The nickel-clad layer is being used for the anode side because Ni is thermodynamically stable in the anode gas environment. With the Ni-coating protection, the anode side did not show any signs of corrosion attack after several thousand hours of operation. Based on published data [CFC Technology, 1993, V.93-3, p.158], the 0.002" thick Ni-layer may be sufficient for 40,000-hour operation. Economical considerations require elimination of the Ni-clad layer. M-C Power has been conducting studies of homogeneous materials that could be technically and economically suitable for the separator plate. This effort is ongoing.

Wet-seal corrosion protection

The wet-seal area simultaneously experiences reducing and oxidizing environments and requires a special corrosion protection. Aluminization of the stainless steel surface has been generally used for this purpose. M-C Power's plates have been aluminized by Al-spray with following heat-treatment to provide an inter-metallic corrosion-resistant coating. Based on projections of the experimental data [CFC Technology, 1993, Vol.93-3, p.160], the aluminized coating is sufficient for 40,000-hour operation provided that a consistent quality of coating is maintained. Further development of wet-seal aluminization is planned to improve the quality and to reduce the cost of the process.

Maximum hardware temperature

Another issue related to stack life is the maximum hardware temperature that is determined by the stack flow geometry. The lower the maximum hardware temperature, the longer stack life. For cross-flow stacks the maximum local hardware temperature is high ($> 700^{\circ}\text{C}$), especially at a high current density. This issue is being addressed through plate and stack hardware design.

Stack Life Verification Testing

The required 40,000-hour stack life has to be verified via a comprehensive development and testing program that will encompass further improvement and cost reduction of individual components, cell packages, and small stacks operating under power plant conditions – gas compositions, flows, temperatures, pressures. A part of this program will include testing of full area stacks, 1000 cm² and 100 cm² stacks at increasing current densities, fuel utilizations and operating pressures.

Stack Design Issues

The main MCFC stack design improvement issues include the following:

1. Development of a cost-effective, reliable and “simple” clamping system for the commercial design. The prototype clamping system, which will be implemented on M-C Power's 450 kW demonstration project, will eliminate use of the bellows and the associated bellows control system by use of springs.
2. Development of a cost-effective and “thin” cathode endplate that will improve thermal cyclability and sealing at the top by conforming to the stack. M-C Power is presently working on this problem as part of the PDI program.
3. Development of a commercial-design plenum to provide proper flow conditions at minimum cost and space. The initial designs will be tested in the 450 kW demonstration project. The plenum will require redesign at the 600 kPa operating pressure as the specific volume of gas is significantly different from the 300 kPa used on the 450 kW power plant.
4. Development of a commercial-design power-bus that satisfies electrical and mechanical requirements. The prototype will be tested in the 450 kW demonstration projected under the Process Design & Improvement Contract. The power bus will

- require rework in the HEFPP project as four stacks will be combined and equipment location and geometries will be different in the 600 kPa power module vessel.
5. Electrical connection of several stacks in series which will raise the total voltage going to the inverter, reduces the inverter cost, improves inverter electrical efficiency, and improves the overall plant efficiency.

Confirmation of predicted performance and endurance at the 600 kPa pressure level will require both bench scale and full cell area testing. With the increase in pressure over the 300 kPa PDI project, the stack pressure drops (plenum arrangement internal manifolding, and plate spacing) will require review and possible redesign to reach the desired pressure drops.

3.A.1.b Gas Turbine Generator

The 3.8 MW Rolls-Royce Allison gas turbine generator must be designed and developed prior to use in the HEFPP project. Specific issues are addressed individually below:

Direct Drive Alternators (DDA)

The advantage of the direct drive alternator is that it eliminates the need for a gearbox between the turbo-machinery and the generator that it drives. This, in turn, opens the way to totally oil-free system providing appropriate bearings are used. The best bearing system is considered to be an Active Magnetic Bearing (AMB) configuration.

The DDA is to be run asynchronously such that there is an option to adjust the turbogenerator speed to suit the selected plant power level. It is anticipated that the plant turndown efficiency will be enhanced if the turbogenerator flow can be decreased at low power level (to prevent overcooling of the stack and to diminish the exhaust energy drain). Although this could be achieved to a degree by using variable vanes at a fixed turbine generator speed, asynchronous running offers a physically simpler and more thermodynamically efficient approach.

DDAs have been developed up to about 75 kW for use with microturbines and individual units have been demonstrated up to approximately 300 kW but not up to the 3.8 MW level. High speed electric motors, which may be regarded as close mirror images of DDAs have been produced at high power [Drives and Controls magazine July/August 1998 which reports an 8 MW motor with a design speed of 10,000 RPM]. The 3.8 MW, 11,000 to 15,000 RPM DDA required for this application must be designed in a manner which is compatible with the overall plant control scheme and, to this extent a new design task is created. Although the solution will be new in detail, it will represent a focusing of available technologies rather than a step into the unknown.

The power conditioning system is anticipated to take high frequency AC from the DDA, rectify it, and then convert it to 60 Hz in a converter dedicated to the turbogenerator. The mechanical integration of the compressor, turbine and DDA will be much facilitated by the use of AMBs, because they permit tuning of equivalent support stiffness to tune out potential vibration modes.

Power Conditioning (Converter)

Converter design choices tend to be influenced by the best solid state electronic component values currently available. Presently, for instance, 250 kW is a good modular size, which is likely to increase. Influenced by the different control requirements of the fuel cell stack and turbogenerator power loops, it is expected that the turbogenerator converter will be kept separate from the inverter or inverter array handling the fuel cell stack power output. The hybrid fuel cell power plant is likely to become a driving force in the advancement of cost effective large power conditioners (converters).

The present capacity of the Insulated Gate Bipolar Transistors (IGBTs) is a limiting factor to a single unit power conditioner. Other technologies such as gate turnoff thyristors (GTOs) are

available but have inherent problems of lower efficiency and slower switching speeds when compared to the IGBT technology. Other emerging technologies, which require further study and development, include the MOS Turn-off Thyristors (MTOs) and the Integrated Gate Comutating Thyristor (IGCTs).

By 2010, the new IEEE electric grid interconnect standard (IEEE P1547) for distributed generation (currently in development) will likely require some modifications to the power conditioner in terms of additional protection and control requirements. It is envisioned the IEEE standard will require sweeping changes throughout the power conditioner industry as the industry moves toward a standard power conditioner configuration and away from each customer or contract having specific site performance requirements. The IEEE standard should help reduce the cost of the inverter performance and programming by minimizing the per contract engineering costs. At present, no power conditioner test protocol exists for testing the 3.8MW power conditioner. In addition, no manufacturer or third party testing organization is set up to full load test a power conditioner of this size. A test protocol and suitable test lab will have to be developed to accommodate the large power conditioner. Acceptance testing of the power conditioner will be critical to utility acceptance of the plant for reliability and safety reasons.

Power conditioner electrical efficiency presently at 95 percent must be improved to the 97.5 percent to improve the gas turbine generator output and the overall plant electrical efficiency.

Active Magnetic Bearings

Rolls-Royce Allison has proposed using magnetic bearings on the plant gas turbine generator. Magnetic bearings have several advantages, which include the following:

1. Long bearing life
2. Bearings are self-correcting to out of balance loads, which may arise through wear and tear or deposition.
3. No lubrication required (eliminates the need and weight of an entire lubrication system)
4. Reduced friction losses, as the only friction at bearings is air. (Reduced friction yields a more efficient machine)
5. The gas turbine is more reliable as the entire lubrication system has been eliminated, which can cause failure.
6. Bearings have inherent ability to reduce noise and vibration caused by rotor imbalance.

Active magnetic bearings (AMBs) with their associated control system inherently offers on-line monitoring and control of bearing forces and rotor displacements allowing early detection of faults. Further control system development work will be required in the programming and controls area to electronically locate the rotor position and to generate the current required for the magnetic bearings.

The control and stabilization of active magnetic bearings is now well understood and the gas turbine generator components can readily be designed to achieve good bearing performance without resort to new materials technology. This is because the bearings can be situated in a reasonably cool environment between 422 °K and 588 °K, which is compatible with normal

electromagnet materials. AMBs are not yet widely in use, so some learning experience must be anticipated during the early embodiment.

Combustor

A low risk conventional combustor will be used in the firing of the Rolls-Royce Allison gas turbine generator. Several considerations affect the design choices for the combustor which appear below:

1. Because the combustor will be fueled for a short time, it should not be required to have ultra-low emissions. At maximum temperature rise, the NO_x emissions will be in the 100-150 ppm range but will decrease as the plant heats up.
2. As the fuel cells rise in temperature, the combustor delivery temperature is held constant requiring less fuel to burn. Combustor stability is required over a substantial range of fuel flow.
3. To avoid complicating the hot ducting, the combustor must remain in the main flow path even though it is not functioning. The pressure drop across the combustor must be minimized.
4. The combustor and associated ducting will have to be insulated in a thermal blanket to avoid overheating adjacent plant equipment.

Rolls-Royce Allison has recommended that a cylindrical three liner type combustor be used. The design point temperature rise is 866 to 921 °K based on plant startup conditions. The normal capability of a single liner system is to turn down by stable means to approximately one quarter of its design temperature rise (135-149 °K). The provision of three liners allows individual liners to be shut down, such that while the last operating liner is still producing a 167 °K rise, the whole flow average rise is only 55.5 °K, permitting a smoother transition from fired to unfired during startup. If it is determined that a larger step change is acceptable, then the system can be simplified to a single liner design. An alternate for further study is to utilize a catalytic element in the combustor in addition to multiple liner operation.

Gas Turbine Generator Control System

The gas turbine generator control system would be used to startup the gas turbine with a start signal from the plant control system. Once the gas turbine has been started, the fuel cell heated, and the combustor switched off, the gas turbine control system fuel control logic would become dormant. The gas turbine control system would control turbine speed and monitor turbine generator supervisory control. The plant control system through the turbine generator programmable logic controller (PLC) would take over the operation of the gas turbine generator to match the overall plant process requirements of the turbine compressor and expander sections. The coordination between the gas turbine controls and the plant control system still requires further development and understanding between Rolls-Royce Allison, M-C Power and Bechtel National.

Modeling work will also be required to determine the load following, load shedding and turndown of the generator and power conditioner. The response to electrical system transients

such as loss of electric grid, ground faults in single phase or neutral, synchronization, and switch to island mode operation must be studied in more detail. Schemes for voltage control, current control, frequency control, power factor control and switching to and from voltage and current control modes quickly must also be studied.

Once the plant has reached the steady state performance, positive control of the turbogenerator running point will be primarily maintained in an asynchronous system by altering the DDA load characteristics. This requirement affects the electrical design of the generator and the associated power conditioning and electrical protection systems. The integration process is expected to require significant development time and capital.

3.A.1.c Balance of Plant

Much of the balance of plant (BOP) equipment (excluding the fuel cell and the gas turbine generator) has no design or development issues because the products available are considered to be mature. Specific pieces of equipment which will require additional design and development prior to use in the - 20 MW high efficiency pressurized molten carbonate fuel cell power plant appear below:

Recycle Blowers

At approximately 610 kPa operating pressure, only a limited number of blower manufacturers have the knowledge base and experience to properly design the recycle blower. The blower operates at a low pressure ratio, high temperature, and moderate pressure. The design issues resulting from the design constraints are as follows:

Shaft seals-The shaft seals must be designed to keep the higher pressure cathode gas from leaking out to atmosphere. Design, material selection, and endurance testing for design confirmation will be required.

Bearing design- The bearing design must accommodate the blower rotor high temperatures and the associated shaft thermal growth. Confirmation of the bearing and shaft design prior to the commercial unit will be required by shop testing of any proposed solution.

Materials of Construction- The materials selected for the blower housing, impeller, and shaft must be designed to prevent high temperature surface corrosion. This may be accomplished by the use of high temperature base materials or with protective spray coatings over a less costly base material. Endurance testing of the proposed materials must be verified by testing prior to installation in a commercial fuel cell power plant.

Power Module Vessel

Each power module vessel contains four (4) molten carbonate fuel cell stacks, four (4) reformer assemblies, and miscellaneous piping. Some of the design and development issues associated with the Power Module Vessel include the following:

Vessel Design-The design must accommodate the removal and reinstallation of the individual reformers and fuel cell stacks. Other requirements include the access requirements for maintenance and transportation of the vessel with equipment installed subject to weight and dimensional restrictions of land-based transportation. Wind loading, seismic loading, and transportation loading (lifting vessel assembly onto and from vehicle, and vessel stiffness to road potholes) design conditions must also be considered in the vessel design. Power module assembly methods and sequences must also be modeled and developed by the plant module fabricator prior to vessel fabrication commencement. A removal scheme for each fuel cell and reformer with special tooling required must be modeled and procedure developed prior to vessel fabrication. Nozzle orientations and allowable stresses must be studied and coordinated with both vessel interior and exterior piping. Electrical power and grounding connections to the power

module components must be routed and designed to the ANSI/NFPA 70, ANSI C2, IEEE 80, and IEEE 142. The instrumentation must be designed to take the high internal vessel temperature and penetrate through the pressure vessel wall without leaks. A supporting vessel structure such as saddles, which controls the vessel thermal growth, must also be developed. With the vessel in operation containing both natural gas and hydrogen, a blast analysis study must be completed to specify and locate the appropriate blast pressure relieving devices. M-C Power will perform most of these tasks on the 450 kW demonstration plant power module vessel. The PDI 450 kW power module vessel will operate at 300 kPa pressure, which results in larger diameter vessel and piping than the HEFPP vessel. Lessons learned will be implemented in the HEFPP power module vessel.

Vessel Insulation Design- The vessel insulation must be studied to allow acceptable heat loss during plant operation at the lowest and highest ambient temperature. The insulation should also be designed for personnel protection from burns due to hot insulation or jacketing, where accessible by plant personnel. The insulation will also be designed to protect the vessel internal metal surfaces from condensation and to allow the use of carbon steel pressure vessel materials.

Piping design- The power module piping design must be studied to eliminate the use of control valves inside the pressure vessel. Manual valve usage inside the pressure vessel will be studied to minimize the number of valves. Pipe thermal expansion inside the vessel will be studied to eliminate the need for expensive expansion joints. High temperature, low-pressure-drop check valves located inside the hot pressure vessel will have to be developed and tested to keep the cathode blower power consumption low. In general the piping will be smaller but heavier wall for the HEFPP 600 kPa operating pressure when compared with the PDI 450 kW piping.

Inverter

The initial design plan is for one inverter to be provided for each power module. Four fuel cells in one common pressure vessel would feed one inverter. The fuel cells will be connected electrically in series. The inverter takes the DC power from the fuel cells and then boosts the DC voltage. Initial thoughts were to have the higher voltage DC power inverted to AC power at nominally 480 VAC. With 9 inverters and a net expected output from the inverters to exceed 17 MW, the electric bus connecting the inverters together becomes large due to the high current handling requirements. The 480VAC inverter output will have to be transformed up to a higher voltage to minimize large bus duct costs and grid connection costs. The present scheme is to have the inverter output voltage changed from 480 VAC to 4160 VAC. A study to determine the most economical bus and type of step-up transformer will have to be completed in the preliminary engineering phases. With the 9 inverters planned, a master synchronization and paralleling scheme for connecting the individual inverters to the grid and to the turbine generator power conditioner (converter) for startup and shutdown of the plant must be developed. The plant electrical islanding scheme and reconnection scheme must also be developed, which will allow the plant to operate in island mode and then synchronize with the grid without dropping load. The synchronization of the plant must occur within a very few cycles.

A study to determine the appropriate emerging inverter technology (Integrated Gate Commutating Thyristors (IGCTs), Insulated Gate Bipolar Transistors (IGBTs), MOS Turn-off Thyristors (MTOs) or Gate Turnoff Thyristors (GTOs)) prior to order placement will be required. In the study, efficiency, cost, footprint size, voltage capacity, current capacity, additional auxiliary equipment/system requirements, and switching speed will be evaluated to select the best available technology. An optimization study will also be required to minimize the number of inverters. A limiting factor for the optimization study will be the maximum AC voltage the selected inverter technology can handle in the high frequency pulse width modulation mode. The pulse width modulation (PWM) technology may be a limiting factor as the maximum rating of present PWM technology is 690 VAC. It is anticipated that by 2003, IGBTs will be available to handle up to 1200 VAC in the pulse width modulation mode.

The present inverter technology in the size range required for the high efficiency power plant yields 95 percent electrical efficiency. Further developments to improve the efficiency to 97 to 98 percent are required. The most significant efficiency improvement may come from the use of trench type IGBTs, which will cut the conduction losses by 50 percent. Another improvement may come from the use of advanced pulse width modulation (PWM) generation algorithms. Implementation of the new algorithms may reduce the IGBT switching losses by approximately 30 percent without sacrificing bridge performance characteristics. Higher grade magnetic materials can be used in the reactors and chokes to reduce the iron losses or reduce the copper losses via shorter mean turn lengths. Two tradeoffs for the higher grade magnetic materials include higher material and component fabrication costs and reduction in the number of available core suppliers.

Fuel cell currents exceeding 1500 ADC present a unique design challenge for the inverter. The bus bar and cable losses generated at this current level are significant and can be difficult to control. Methods to reduce the losses include a) use of less resistive materials, b) use of broader cross-sections, and c) shorter circuit lengths. All of the above methods will be employed to some extent to repackage the inverter and improve the overall electrical efficiency.

By 2010, the new IEEE P1547 Interconnection standard for distributed generation (currently in development) will probably require modifications to the inverter in terms of additional protection and control requirements. At present no standard exists to test the inverters of this size. In addition, no manufacturer has the in-house testing facilities to perform a full load test of the inverters envisioned. Acceptance testing of inverters will be critical to obtaining electric utility acceptance of the plant for reliability and safety reasons. Development of a test protocol and test facility either at the manufacturer or a third party is required.

Decisions on whether the turbine generator power conditioner or the fuel cell inverter will have the master synchronizer to the electric grid must be made. Initial thoughts on this matter include that the turbine generator power conditioner will be the master as it is the first piece of equipment to start and last to operate under normal conditions. Development of the synchronization scheme requires further work.

Reformer

The reformer will also require some development work primarily due to the change in pressure from 300 kPa (used on the PDI project) to the 600 kPa HEFPP operating pressure.

Ishikawajima-Harima Heavy Industries (IHI) has developed a reformer for 1mPa operation for Japan MCFC projects and 300 kPa operation for Miramar. Operation at 600 kPa will require changes in the flow distribution path. Specific areas of change include piping (where pipe size will get smaller due to lower specific volume but pressure will require thicker pipe materials) and manifolding /plate spacing (where manifolds and plate spacing should get smaller due to lower specific volume). The reformer at 600 kPa operating pressure will also require some testing to confirm performance.

The HEFPP goal for reformer catalyst replacement is 10 years. IHI will develop and test their catalyst to obtain 40,000 HR life as part of the PDI project. Catalyst life is dependent on the amount of catalyst deposited per unit area on the reformer plates and the total plate area in the reformer. Some additional catalyst life development work will be required by IHI for the HEFPP project.

3.A.2. Confirmation of Design Parameters

3.A.2.a Fuel Cell

The MCFC stacks will be designed to operate at the following conditions for the HEFPP study:

Pressure (kPa)	605 (~6 atm)
Fuel utilization (per pass, %)	80.4
Current density (mA/cm ²)	200
Average cell voltage (mV)	758 (at the beginning of life).

When adjusted for operating conditions, this performance is close to the performance that has been achieved in bench-scale fuel cells at IGT and M-C Power. For example, fuel cell IM9705, tested at IGT for approximately 5,000 hours, had a voltage of 730 mV at 200 mA/cm² at 303.9 kPa (3 atm) and 75% fuel utilization at the onset of the test. Pressure was increased from 303.9 kPa to 605 kPa, which increased cell voltage by 35 to 40 mV; a fuel utilization increase from 75 to 80% reduced the voltage by 7-10 mV. Hence, cell voltage at the HEFPP conditions would be approximately 760 mV. M-C Power's fuel cell performance model also predicted that the performance required for MCFC stacks at the HEFPP is achievable.

Based on experimental data, model predictions, other published data, and M-C Power's performance improvement plan, this level of performance for MCFC stacks is obtainable.

The challenge is to achieve and maintain this level of performance at the full-area commercial-size stacks.

The following issues related to performance improvement are being addressed presently as part of the DOE funded PDI program:

1. Optimization of component pore matching and electrolyte distribution among components. Implementation of better pore matching will improve performance by reducing anode and cathode polarization. Adjusting the pore-size distributions and the wetting properties of the electrodes will improve pore matching.
2. Optimization of the cathode thickness and its porous structure. Because of the higher Li/Na ionic conductivity, the effective thickness of the cathode with Li/Na electrolyte is higher than that of a cathode operating with Li/K electrolyte. Increasing cathode thickness and increasing external agglomerate surface will improve cathode performance. Optimizing the cathode oxidation/lithiation reaction will result in 20-30 mV performance improvement.
3. Optimization of electrolyte composition. Comparison with the-state-of-the-art cell performance shows that the cell performance at 160 mA/cm² can be improved by 20-25 mV simply by the change of electrolyte composition. Another advantage of this alternate electrolyte is a reduced sensitivity of cell performance to the operating temperature, which results in a higher stack output and a lower stack maximum temperature.

4. Improvement of component manufacturing tolerances will result in: (a) more uniform flow distribution among cells and will allow the stacks to operate at 80+ % fuel utilization per pass and (b) uniform contact pressure distribution over the active area with reduced stack internal resistance.
5. An advanced plate design to reduce cost, ensure 100% sealing, lower electrolyte losses for corrosion, and improve uniformity of flow and pressure distribution.

The full-area commercial size stack performance will be verified via a comprehensive testing program that includes testing of 100-cm², 1000-cm², along with full-area cells and stacks.

3.A.2.b Gas Turbine Generator

Rolls-Royce Allison has investigated the critical design parameters associated with a 6.0 pressure ratio 3.8 MW class gas turbine-generator. Specific design parameters are addressed individually below:

Compressor

Up to a pressure ratio of approximately 6, axial and radial machines give similar performance between inlet and diffuser exit (at 0.15 Mach number). In the study analysis, a seven stage axial compressor was 0.6 percent more efficient than a single stage radial compressor. Above this pressure ratio, the radial compressor efficiency deteriorates when compared to the axial compressor due to rising Mach number effects in the single radial stage. The maximum radial stage pressure efficiency is normally limited by metal strength properties (maximum pressure ratio of 6.5 for steel and 8 to 9 for titanium). Radial compressors offer a large reduction in number of parts and cost for a given pressure ratio. At a pressure ratio of 6, the most cost effective compressor design results in a single steel stage radial compressor. It should be noted that the radial compressors have not been produced in aeroderivative gas turbines for many years. The radial compressor will have a large scroll diameter approaching 1.5 meters. In developing the radial compressor design, the following design parameters were used:

1. Inlet hub/tip radius ratio will be set to 0.4 to avoid extreme airfoil geometry (also allows for bearing if desired)
2. Inlet tip radius is designed to avoid exceeding a 1.0 inlet Mach number relative to the blades.
3. The rotor tip radius is adequate to generate the design pressure ratio at appropriate loading.

Turbine

In the HEFPP, turbine gas flow will be delivered from the fuel cell to the gas turbine generator by a circular pipe. Hot process gas from the fuel cell is transferred to the axial turbine inlet through a carefully shaped volute in which the design and fabrication processes have been used many times by Roll-Royce Allison. Delivery of gas from the volute is designed to be at a fixed swirl and axial velocity components to maintain a constant nozzle guide vane entry angle. The tangential component of the velocity serves to partially offload the vanes. By taking the principle too far, the resultant will lead to large volute velocity and potential wall scrubbing losses. It is therefore desirable and efficient to retain a row of guide vanes for the turbine section.

The turbine inlet temperature of 704°C is low enough to allow the use of low cost materials and allows elimination of turbine wall cooling temperatures.

The single shaft design turbine generator imposes a lower speed limit than preferred for the compressor. The reasons for this are as follows:

1. The volumetric flow leaving the turbine is 2 to 2.5 times larger than that entering the compressor because the exhaust temperature is higher. (If a stage hub/tip radius ratio of 0.5 is

used, the last turbine blade tip radius is approximately 1.5 times that of the first compressor blade which gives a correspondingly higher tip speed.)

2. The power turbine tips normally have to carry shrouds for maximum efficiency.
3. The turbine blades are hotter.

The axial turbine design criteria used are as follows:

1. Minimize the turbine stages required for each duty.
2. Keep the stages at moderate internal Mach number and loading to retain high efficiency.
3. The final stage exit area should yield an axial Mach number of 0.35. Diffuser losses should be kept at no greater than 40 percent of dynamic head, which yields a maximum 3 percent total pressure loss.
4. The highest rotational speed should be selected based on the selected material properties.
5. The exit hub/tip radius ratio should be kept between 0.5 and 0.6 to avoid local aerodynamic problems and provide sufficient space for structure. (Note: this now fixes the last stage hub and tip radius.)
6. Airfoils with high aspect ratio should be used to give good efficiency and low shroud stress. The aspect ratio should also be low enough to limit the blade count (cost) and avoid excessively thin sections or steep annulus angle. This criteria determines the stage axial lengths.

At a pressure ratio of 6, it was found that the preferred design RPM for a single stage radial compressor of 13,200 was acceptable when paired with the maximum desired value of the axial turbine. An axial compressor does not offer a significant efficiency advantage for this cycle.

For higher pressure ratio cycles, the tip speed of the radial turbine stage has to be increased (approximately 20% for a 10 pressure ratio machine) and this moves it into a region where it experiences shock losses. More efficient units can be created using either all axial or hybrid axi-centrifugal multistage designs. In order to keep the minimum number of stages, relatively high speed is still required. The speed limiting feature in a single shaft machine is normally the last stage turbine blade hub radial stress. The level of stress can be shown to be fixed by the blade swept area and the RPM. The swept area is set by the air flow amount, its exit temperature, and the requirement for an axial Mach number not exceeding 0.35.

For cycle options above 6 pressure ratio, it was found that single shaft configuration RPM is limited by turbine last stage stresses. This is just acceptable for a 10 pressure ratio radial compressor of lower efficiency (due to shock losses), but leads to the need for more stages in a more efficient axial or axi-centrifugal machine. Without RPM compromise, a (3 stage axial + 1 stage radial) hybrid machine is acceptable for 10 pressure ratio, but, when slowed to turbine-acceptable speed, (6 axial stages + 1 radial stage) is required. Although this represents a cost penalty to a single shaft configuration, the six axial stages/1 radial stage configuration remains the preferred scheme because of its merits of flow control for power turndown and resistance to load shed overspeed. In addition, pure axial compressors of appropriate speed are already in production, offering cost-effective solutions for a larger size fuel cell plant.

Given that the turbines are running near to the stress limited speed, the number of stages can be minimized. Two stages are necessary for the 6 pressure ratio cycle and three stages are required

for the 10 and 14 pressure ratio cycles respectively. It is highly unlikely that a low cost single stage radial turbine would give good results in the 6 pressure ratio cycle (at only 977 R inlet temperature, the loading would be greater than optimum) and would be unsatisfactory for the higher pressure ratio cases.

All the axial turbines proposed for the three pressure ratio levels are predicted to yield efficiency between 91 and 92 percent.

Alternator

Several options were considered for the alternator including the following:

1. Compressor, turbine, and asynchronous alternator on a single high speed shaft.
2. Compressor and turbine on one shaft with a gearbox driving an asynchronous alternator at 6000 RPM
3. Compressor and drive turbine on one shaft. Power turbine and asynchronous alternator on a second somewhat slower shaft.
4. The compressor and turbine on one shaft with a gearbox driving a conventional synchronous alternator at synchronous speed.
5. Turbine and compressor on separate shafts running at their individual best speeds geared to a synchronous alternator at synchronous speed.

In the first three options, the asynchronous generator would have to be connected to a power conditioner to produce the desired 60 Hz power. The power conditioner has losses associated with the conversion of high frequency power to 60 Hz power of 3 to 4 percent. In options three and four, the generator is synchronous and operates at 60 Hz power, but has the additional gearbox losses, which will reduce the overall turbine generator efficiency by 3 to 4 percent. The geared units also require a lubrication system to keep them operating properly, which may be eliminated with a single common shaft turbine/compressor/asynchronous alternator utilizing magnetic bearings.

The options were also evaluated economically in terms of first cost. The economic choice was the compressor, turbine, and asynchronous alternator on a common shaft option, as the assembly was the lightest of all the options. The geared options have additional weight and cost associated with the gearbox, lubrication system, and the generator. Reductions in speed, force the turbine, compressor, and generator to become larger.

Option 1 has some drawbacks in that the direct drive alternator in the 3.8 MW size range must be scaled up from existing technology presently at the 75 kW level or derived from the high speed motor technology. For more detailed information, please refer to the Design and Development Section 3.A.1.B.

Power Conditioner

With the direct drive alternator, a power conditioner or inverter must be used to convert the high frequency AC power to DC power and then back to 60Hz AC power. Several technologies exist or are in development such as insulated gate bipolar transistor (IGBT), gate turn-off thyristor (GTO), MOS turnoff thyristor (MTO), and integrated gate comutating thyristor (IGCT). The IGBTs are limited to approximately 250 kW single unit size but can have several units placed in parallel on each of three phases to meet the desired power output. The largest paralleled power conditioner in use today is 1.5 MW in capacity used in windmill applications. GTOs may also be used but suffer from slower switching speed and lower efficiency when compared to the IGBT technology. GTOs allow higher current and voltage to be used when compared to the IGBT technology. MTO and IGCT are still being developed but hold promise as the technology of the future. For further discussion, please refer to the Design and Development Issue Section 3.A.1.B.

Magnetic Bearings

Active magnetic bearings (AMBs) have been selected by Rolls-Royce Allison for use on the proposed gas turbine generator. The AMBs feature reduced friction losses, long bearing life, self correction to out of balance loads, noise reduction, and require no lubrication when compared to conventional lubricated bearings. AMBs are well understood, but will require some development work in the programming and controls required to electronically locate the rotor position and to generate the currents required for the magnetic bearings. For more information, please refer to the Design and Development Issue Section 3.A.1.B.

Combustor

In regular aero-derivative turbine generator practice, a 4 percent pressure loss from the compressor diffuser outlet to turbine inlet is required to serve the following purposes:

1. To drive the high velocity dilution air jet streams into the final combustion zones to ensure complete combustion and provide proper mixing within a short length.
2. To ensure that the cooling flow passed to the first turbine blade row is at a higher pressure than the very hot gases impinging on the turbine blade leading edge. This permits the formation of leading edge cooling films on the blade.

With the molten carbonate fuel cell power plant, the turbine generator combustor is different as follows:

1. There is more space for mixing.
2. There is no need to cool the turbine blades as operating temperatures of approximately 700 C are expected which is well below the much higher conventional temperature aeroderivative combustors.
3. The combustor is used only unit startups and transients but most of the pressure losses are always present when not in operation.
4. Weight and operating volume do not have to be minimized.
5. No requirement for very high wall temperatures or cooling air.

The preliminary sizing of the combustor components yields a low velocity unit of around 2 feet diameter burning zone and approximately 5 feet length. This combination gives an unfired pressure loss of 1.5 to 2 percent. The parameters mentioned above place the combustor well within the range of conventional duct burner technology.

3.A.2.c Balance of Plant

The next tasks are to establish the equipment design parameters and to select the process operating conditions. Equipment manufacturers provided the equipment design parameters. Table 1 lists the equipment design parameters and operating conditions for the major subsystems.

The selection of operating conditions is based on achieving the highest natural-gas-conversion efficiency for the plant. The following high-efficiency prerequisites directed the process-simulation search for the optimum operating conditions:

- high conversion of natural gas to H₂-rich fuel
- high conversion of H₂ fuel to electricity in the MCFC
- high conversion efficiency for H₂ chemical energy to electrical energy
- minimize fuel combustion
- high efficiency bottoming cycle

The key operating conditions are listed in the following Table:

Full-Load, Start of Run (SOR) Operation – 15°C Ambient Air, Sea Level	
System Pressure	610 kPa
Steam Temperature	536°K
Anode-Recycle Flow, weight ratio	3:1
Steam:HC-Carbon, mole ratio	4.7:1
Reformer-Inlet Temperature	799°K
Reformer Outlet Temperature	1,061°K
Anode Inlet Temperature	950°K
Cathode Inlet Temperature	978°K

Selection of Optimum Operating Conditions

The effects of key-equipment operating conditions on equipment performance and the natural-gas-conversion efficiency are discussed below.

System Pressure

The effect of system operating pressure on plant performance is discussed in Section 3.A.3.

Steam Temperature

Increasing the steam temperature results in a slight increase in power output from the gas-turbine generator - just over 1% increase for the steam temperature range 480-560°K. However, any increase in superheat is at the expense of steam flow (since steam is generated in a non-fired heat recovery steam generator (HRSG) unit), and reducing steam flow affects methane conversion in the reformer, leading to lower overall efficiency.

Steam Flow Rate

The steam flow rate to the reformer is an important determinant of the natural gas conversion. This variable plays an important role in determining the natural-gas-conversion efficiency.

The steam flow rate must be above the minimum needed to prevent carbon formation in the reformer and provide a safe margin to carbon formation in the anode. For this design, the steam flow rate is the maximum steam flow that can be generated, at 900 kPa and 536°K, from the gas turbine exhaust, without fuel firing.

As mentioned above, higher methane slip from the reformer results in more fuel combustion in the reformer combustor. When there is sufficient H₂ and CO in the anode exhaust to provide the reformer duty, the combustion of residual methane serves no other purpose than participating in the gas turbine power cycle. Therefore, under these circumstances, residual methane lowers the overall natural-gas-conversion efficiency, and must be minimized.

The steam flow rate also determines the quantity of anode recycle to the reformer.

Reforming Temperature

High reforming temperatures - i) increase methane conversion, ii) require more residual fuel in the anode exhaust to meet the reforming heat duty (lower plant efficiencies), and iii) requires more expensive materials of construction. The operating temperature is selected by careful consideration of the materials of construction and the values for other process variables affecting methane conversion (pressure, steam flow rate).

For this design, the low operating temperature is offset by high steam flow to the reformer and low operating pressure. The low reforming temperature means that the reformer can be fabricated from stainless steel.

Anode Recycle

Anode recycle improves flow distribution in the anode, facilitates thermal management, and lowers per-pass fuel utilization in the anode compartment. By introducing the recycle stream upstream of the reformer, the additional water content in the recycle gas helps with the conversion of natural gas to hydrogen-rich fuel.

Anode recycle improves the basic design concept in three important process areas. First, as mentioned, recycling exhaust gas from the anode compartment increases the amount of steam flowing to the reformer. Second, recycling anode exhaust results in an increase in the quantity of hydrogen entering the anode compartment. Therefore, for a fixed fuel conversion, the per-pass fuel utilization is lower. Third, anode recycle helps with the thermal management of the stack.

Higher concentrations of CO₂ and H₂O in the anode feed, as a result of recycling anode exhaust, should lower the stack operating voltage - through the Nernst term in the reversible cell potential equation. However, for the same high fuel conversion without anode recycle, the severely depleted H₂ concentration at the anode exhaust lowers the exit reversible potential significantly.

Since electrodes are iso-potential surfaces, the cell potential, for co-flow designs, is set by its exhaust value.

For the same fuel conversion, the per-pass fuel utilization increases from 80% with recycle, to 94% without recycle. To maintain 80% fuel utilization without anode recycle, the steam flow to reformer has to be increased. Raising additional steam means fuel has to be fired either in the turbine combustor or in the HRSG unit. Raising the additional steam lowers the natural-gas-conversion efficiency from 70% to 63%.

Increasing the reformer outlet temperature from 1,060°K to 1,117°K, in an attempt to increase the methane conversion at reduced steam addition, lowers the efficiency to 61%.

The amount of anode recycle is set by the pre-reformer temperature limitation. Increasing the recycle-to-steam flow weight ratio beyond 3:1 causes the temperature to exceed the pre-reformer catalyst maximum operating temperature.

Anode Inlet Temperature

The anode feed gas must be hot enough to keep the electrolyte in a molten state. The anode feed gas temperature must be above the Boudouard carbon formation temperature.

Cathode Inlet Temperature

Like anode feed gas, the cathode feed gas temperature must be sufficient to keep the electrolyte in a molten state.

Cathode Outlet Temperature

Stack cooling, by an external, gas recycle around the cathode, is responsible for the main auxiliary power demand. Recycle flow, and hence recycle blower horsepower, is minimized by operating between the minimum inlet temperature of 866°K and the maximum outlet temperature of 978°K.

High operating temperatures are needed to achieve lower electrolyte conductivity, which minimizes Ohmic resistance losses and consequent heat generation.

Vessel Heat Loss

The power module vessel is insulated to minimize heat loss. The argument for minimizing heat loss was made above.

TABLE 1
Major Subsystem
Design Parameters and Operating Conditions
Site Elevation-Sea Level

	SQR	EOR
Fuel Processor - Reformer	Type: External/Separate - IHI Gas-Heated Plate-Type	
Steam:Hydrocarbon ratio	4.68	5.3
Steam:Total Carbon ratio	1.43	1.5
Operating Pressure, kPa	606	606
Exit Temperature, K	1,061	1,061
Process Duty, kJ/h	1,018,806	1,026,361
Number of Units	36	36
Fuel Cell Stack Design	Type: MC-Power Molten Carbonate, Co-Flow	
Operating Pressure, kPa	605	605
Current Density, mA/cm2	200	200
Cell Voltage, mV	758	700
Power Density, kW/m2	1.52	1.40
Fuel Utilization - per pass, %	80.4	80.3
Fuel Utilization - overall, %	86.2	86.5
Number of Stacks Required	36	36
Inverter Design	Type: Solid-State	
Input Voltage, Volts	227	210
Input Current, Amps DC	2,180	2,180
Output Voltage, Volts	4,160	4,160
Output Electric Power, kW	479	443
Power Conversion Efficiency, %	97.5	97.5
Number of Units Required	9	9
Recycle Gas Blower		
Adiabatic Efficiency, %	75	75
Differential Pressure, kPa	6.2	6.2
Shaft Horse Power, kW	108	119
Motor Efficiency, %	92	92
Number of Units	3	3
Gas Turbine Design	Type: Rolls-Royce Allison	
Compression Ratio	6.16	6.16
Compressor Air Flow, kg/s	16.81	19.19
Compressor Efficiency, %	84	84
Turbine Efficiency, %	91.9	91.9
Generator Efficiency, %	98.5	98.5
Net Power Production, kW	20,115	19,159
Overall Electrical Efficiency	70.11	66.78

TABLE 2
Performance Summary
Site Elevation – Sea Level

~~Start-of-Run Operation~~

Ambient Temperature, °C	15	-29	49	49
Relative Humidity	60%	0%	0%	100%
AC Power from Power Modules, kW	17,256	17,256	17,256	17,256
AC Power from Gas turbine Generator, kW	3,429	3,603	3,246	3,242
Gross Power Production, kW	20,685	20,859	20,502	20,498
Auxiliary Loads, kW	<u>570</u>	<u>571</u>	<u>571</u>	<u>563</u>
Net Power Production, kW	20,115	20,288	19,931	19,935
Natural Gas Consumption, kJ/h (LHV)	1.0328E+08	1.0328E+08	1.0328E+08	1.0328E+08
Fuel Fired in Gas Turbine, kJ/h (LHV)	0	0	0	0
Fuel Fired in HRSG, kJ/h (LHV)	0	0	0	0
Overall Electrical Efficiency (LHV), %	<u>70.11</u>	<u>70.72</u>	<u>69.47</u>	<u>69.49</u>

TABLE 3
Performance Summary
Site Elevation – 5,000'

~~Start-of-Run Operation~~

Ambient Temperature, °C	15	-29	49	49
Relative Humidity	60%	0%	0%	100%
AC Power from Power Modules, kW	17,256	17,256	17,256	17,256
AC Power from Gas turbine Generator, kW	3,099	3,368	2,819	2,845
Gross Power Production, kW	20,355	20,624	20,075	20,101
Auxiliary Loads, kW	<u>570</u>	<u>571</u>	<u>571</u>	<u>563</u>
Net Power Production, kW	19,785	20,052	19,504	19,538
Natural Gas Consumption, kJ/h (LHV)	1.0328E+08	1.0328E+08	1.0328E+08	1.0328E+08
Fuel Fired in Gas Turbine, kJ/h (LHV)	0	0	0	0
Fuel Fired in HRSG, kJ/h (LHV)	0	0	0	0
Overall Electrical Efficiency (LHV), %	<u>68.96</u>	<u>69.90</u>	<u>67.98</u>	<u>68.10</u>

TABLE 4
Performance Summary
Site Elevation – Sea Level

~~End-of-Run Operation~~

Ambient Temperature, °C	15	49
Relative Humidity	60%	0%
AC Power from Power Modules, kW	15,946	15,946
AC Power from Gas turbine Generator, kW	3,821	3,595
Gross Power Production, kW	19,767	19,541
Auxiliary Loads, kW	<u>608</u>	<u>610</u>
Net Power Production, kW	19,159	18,931
Natural Gas Consumption, kJ/h (LHV)	1.0328E+08	1.0328E+08
Fuel Fired in Gas Turbine, kJ/h (LHV)	0	0
Fuel Fired in HRSG, kJ/h (LHV)	0	0
Overall Electrical Efficiency (LHV), %	<u>66.78</u>	<u>65.99</u>

3.A.3. Optimum Operating Pressure Study

After establishing the design concept, the BOP equipment configuration, and equipment design parameters, the next priority is to select the system operating pressure. In this study, pressure selection is based on process performance. i.e., the pressure, which results in the highest natural-gas-conversion efficiency, is selected.

Feed Stream Conditions

Per the study basis, process streams delivered to the battery limit are either at some nominal supply pressure, or else ambient pressure.

- Natural gas is supplied at 205 kPa, and has to be compressed to the system pressure.
- Ambient air, the oxidant and cooling streams in the fuel cells, is compressed to system pressure by the gas turbine compressor.
- Finally, steam is supplied to the power modules above system pressure, since steam is the motive fluid for the recycle gas ejector.

Three system pressures were considered: 600 kPa, 1,000 kPa and 1,400 kPa. These pressures cover the pressure-ratio ranges for aero-derivative gas turbines proposed by Rolls-Royce Allison.

Process simulations of the proposed design concept (power module, gas turbine, and HRSG unit) were performed at the different operating pressures, with equipment performance characteristics provided by equipment suppliers.

Effect of Pressure on Power Generation Efficiency

1) MCFC Stacks

The efficiency at which an MCFC stack converts chemical energy to electric power is directly proportional to the ratio of potential developed to the open circuit voltage. The Nernst equation defines the relationship between the cell's reversible potential and reactant partial pressures. An increase in system pressure results in an increase in cell potential. In addition, higher pressure enhances mass transfer and reaction kinetics at the electrodes, helping reduce the overpotentials - thereby increasing the operating voltage. The MCFC efficiency increases from 59% to 61% as the pressure is increased from 600 kPa to 1400 kPa.

Other Effects of Pressure

In addition to the above, system pressure affects the design and operation of MCFC stacks in several other ways.

Stack Pressure Drop

For a fixed cell configuration and stack design, higher pressure reduces the pressure loss for the fuel and oxidant streams. Low pressure drop is especially important for the flow of oxidant/cooling gas to the cathode compartments. A motor-driven blower delivers this stream to

the cathode compartments and recycles cathode exhaust back to the blower as part of the stack cooling process. The power requirements for the recycle blower are reduced significantly at higher operating pressures.

Cathode Dissolution

High CO₂ partial pressures can lead to nickel dissolution at the cathode – eventually leading to degradation of the cathode.

Carbon Formation

If carbon formation at the anode inlet occurs by the Boudouard reaction, then, since this reaction is favored by high pressure, the Boudouard carbon formation point moves closer to the operating condition of the anode feed gas. i.e., the safety margin between the anode inlet gas temperature and Boudouard carbon formation is reduced.

Electrolyte Vaporization

Since the carbonate evaporation rate is inversely proportional to the CO₂ partial pressure, loss of electrolyte through vaporization is reduced at higher pressures.

2) Gas-Turbine Generator Set

The cycle efficiency for the HEFPP gas turbine generator, relative to its maximum, is shown in Figure 1. The performance as a function of pressure ratio is based on a gas turbine operating as the “unfired” bottoming cycle to a MCFC power module (with fixed steam and natural gas flows) and for a fixed amount of waste heat available from the power modules.

FIGURE 1

From the figure above, the gas-turbine cycle maximum efficiency (ratio of net power produced to heat input¹) occurs in the pressure-ratio range 12-16².

As mentioned above, the HEFPP plant is expected to achieve maximum natural-gas-conversion efficiency when the MCFC stacks and the gas turbine are both operating at their maximum cycle efficiencies.

From the figures above, the optimum operating pressure appears to be somewhere in the range 1,000-1,600 kPa. However, this is an incomplete assessment, for, when the MCFC and the gas-turbine are integrated with BOP equipment a different picture emerges as discussed below.

Turbine Exhaust Temperature

For simple gas turbine cycles, the higher the pressure ratio, the lower the turbine exhaust temperature. For a 978 °K turbine inlet temperature, pressure ratios of 10 and above lead to turbine exhaust temperatures that are too low to preheat natural gas to the desulfurization temperature and too low to raise the process steam requirements. Simply put, as the system pressure increases above 600 kPa, the turbine exhaust has insufficient energy at the right temperature level to meet the process heating requirements.

¹ Power module waste heat is held constant.

² As the pressure ratio tends to 1 the net work tends to zero. At the maximum pressure ratio, where the compressor discharge tends to the turbine inlet temperature, the net work also tends to zero.

Turbine Exhaust Temperature, @ 978 °K inlet temperature			
Pressure Ratio	6	10	14
Exhaust Temperature, °K	670	593	547

Above 600 kPa it is necessary to fire fuel in either the turbine combustor or a HRSG duct burner. This effect is made worse by the steam saturation temperature, which increases as the system pressure (and therefore the steam pressure) increases.

Fuel combusted in either the reformer combustor or the gas-turbine combustor can only be converted to electric power through the gas-turbine cycle, at its cycle efficiency.

Consequently, for the concept proposed for the HEFPP demonstration plant, higher operating pressures result in lower overall electrical conversion efficiencies. Therefore, 610 kPa operating pressure was selected for the system design.

A summary of plant performances at different system operating pressures is shown in Table 5.

Effect of Pressure on BOP Systems

The following is a summary of the effects of system pressure on BOP equipment.

Natural Gas Compressor

Power consumption, design complexity, and cost for the natural gas compressor increase with increasing operating pressure.

Steam System

Increased system operating pressure means that steam for the ejector and fuel processor has to be raised at higher saturation temperature. However, increased system pressure results in lower turbine exhaust temperatures. Above 600 kPa, these two temperatures diverge from one another, making it necessary to combust fuel to redress the situation.

Fuel Processor

Increased operating pressure reduces methane conversion in the reformer. With increasing pressure, the target conversion is achieved either through higher reforming temperatures and/or increased steam flow. Otherwise, lower conversion has to be accepted, and the natural gas feed has to be increased. For 1,000 kPa and 1,400 kPa operating pressures, increasing the steam supply proved to be more efficient than accepting lower natural gas conversions by approximately 1% in natural-gas-conversion efficiency.

Any unconverted methane passes through the anode compartment to the reformer combustor, where it is combusted along with unconverted MCFC fuel (H_2 and CO). If there is sufficient H_2 and CO in the MCFC anode exhaust (flowing to the combustor) to supply the reformer heat duty at the correct temperature levels, then combustion of unconverted methane serves no other purpose than to generate power through the gas-turbine cycle, and should be avoided, or minimized.

Therefore, as far as the fuel processor is concerned, higher pressure operation tends to reduce natural-gas-conversion efficiency – either through increased methane slip, or from additional energy requirements to operate the reformer at higher temperature and/or to provide additional steam.

Recycle Blower

At higher system pressures the recycle blower has lower volumetric throughputs and has to develop less differential head. The net effect is to lower the plant auxiliary load.

Power Module Vessel

The operating pressure of the power module vessel is effectively the same as the system pressure. For the same diameter vessel (fixed equipment and piping sizes), higher pressures lead to thicker vessel walls. Higher pressures also lead to hotter purge air from the compressor. For a pressure ratio of 14:1 and an ambient temperature of 49°C, the vessel purge air temperature is close to 700 °K. These temperatures present additional design challenges for equipment instrument cables and vessel design temperatures.

Power Module Equipment Heat Loss

Power module equipment inside the vessel and the vessel itself are both thermally insulated to minimize heat loss – see discussion above.

TABLE 5
HEFPP Performance Summary
At Different Operating Pressures

Operating Pressure, kPa	610	1010	1420
Cell Voltage, mV	758	771	785
Per-pass Fuel Utilization, %	80	80	80
Cell Efficiency, %	59.1	60.2	61.2
Cathode-Inlet CO ₂ partial pressure, kPa	54	85	112
Reformer - Feed Methane Conversion, %	98.2	97.2	94.7
Steam Pressure, kPa	896	1,303	1,710
Saturation Temperature, K	448	465	478
BFW Consumption, kg/h	5,899	6,199	5,838
Compression Ratio	6.2	10.2	14.3
Compressor Discharge Temperature, K	518	603	666
Gas Turbine N.G. Fuel, kJ/h (LHV)	-	1.433E+07	2.072E+07
Gas Turbine Inlet Temperature, K	978	1,124	1,167
Gas Turbine Exhaust Temperature, K	671	698	674
Number of Power Modules	36	33	30
MCFC Stack Power Output, kW	17,157	15,380	14,487
G.T. Generator Power Output, kW	3,409	5,043	5,943
Auxiliary Power Consumption, kW	353	124	65
N.G. Compressor Power Consumption, kW	213	299	365
Net Power output, kW	20,000	20,000	20,000
N.G. Fuel to G.T. Combustor, kg/h	-	313	453
N.G. Feed to Steam Reformer, kg/h	2,244	1,988	1,895
Total N.G. Fuel Consumption, kJ/h (LHV)	1.027E+08	1.053E+08	1.074E+08
Overall Conversion Efficiency (LHV), %	70.1%	68.4%	67.0%

3.A.4. Optimum Number of Spare Trains for Each Major Process Area

To perform maintenance on any power module component, the power-module vessel has to be depressurized. The number of power modules contained in a single vessel is a compromise between vessel size, number of vessel penetrations, transportation limitations, and the power-loss burden to maintain, or replace, a single power module. With four power modules per vessel the power reduction incurred by taking a single vessel off-line is 11%. Similarly, loss of a single fuel cell inverter will reduce the plant output by 11% and force the power module vessel to be shutdown.

With four modules per vessel and thirty-six modules required for a 20 MW facility, there is a total of nine power-module vessels.

A single recycle blower serves three power-module vessels (12 power modules). This creates three power trains. Shutdown of a single blower results in the shutdown of its power train, and the loss of one-third of plant capacity.

Because of the modular nature of fuel cell power generation, there is appreciable manifolding and sub-manifolding required to distribute feed, exhaust, and utility streams to and from all the power modules.

An alternative arrangement is to locate all power modules inside a single, field-fabricated vessel. The advantages to this are: economy-of-scale cost reduction for the vessel, distribution to equipment is made inside the vessel (fewer vessel penetrations), most of the high-temperature piping is inside the vessel and therefore less insulation is needed, and piping design pressures and flange ratings can be lower. The main disadvantage is the need to shutdown the entire facility in order to maintain equipment inside the stack.

3.B System Definition

3.B.1 Process Description

Please refer to the Process Flow Diagram in Section 3.B.3 for better understanding of process.

Natural Gas Supply & Conditioning

Two, 100% capacity, motor-driven compressors (one operating, one standby) boost the natural gas from supply pressure to system operating pressure, 749 kPa. A heat-exchange coil in the HRSG unit preheats the pressurized natural gas to desulfurization temperature, 616 °K. Natural gas desulfurization takes place in a cylindrical vessel packed with zinc oxide. Following desulfurization, the natural gas splits into three equal streams - each stream flowing to one of the three MCFC power trains. Each power train contains three power module vessels, and each vessel contains four individual power modules. Natural gas is distributed to each of the power modules (thirty-six total).

Steam Supply & Conditioning

Demineralized boiler feed water (BFW) is supplied to the plant battery limit. An economizer coil (also inside the HRSG unit) preheats demineralized BFW before it enters the deaerator. The deaerator strips the BFW of dissolved oxygen by contacting it with steam, over trays or packing. Deaerated BFW has a residual oxygen concentration of 5ppb. BFW is then pumped through the BFW heater to an elevated steam drum, where steam is generated by natural circulation. Saturated steam leaves the steam drum and goes to a superheater where the steam is superheated. The superheated steam leaves the superheater and is distributed to the individual power modules.

Power Module

A single power module consists of a fuel processor (reformer), its supporting equipment, and a 500-kW MCFC stack. There are thirty-six power modules in total.

Power Module Vessel

Each power module vessel contains four separate power modules. The power module vessels are pressurized to the system operating pressure, i.e. the operating pressure of the fuel processors and the fuel-cell stacks. A small flow of compressed air simultaneously purges the vessel and maintains vessel pressure. The air stream flows from the vessel to the process side of the power module, and prevents combustible gases from accumulating inside the vessel in the event of an equipment leak. The fuel cells are connected electrically in series. A single power module vessel provides power for one inverter which is located externally, but adjacent to the vessel.

Fuel Processing

Natural gas is converted to a hydrogen-rich fuel stream by steam reforming. IHI will utilize a similar plate type reformer which has been used successfully on the 250 kW Test Facility at Miramar.

Anode Recycle (Ejector)

Recycling part of the anode exhaust through an ejector creates a recycle flow between the MCFC stack and the fuel processor. Steam flow to the individual power modules is the motive fluid for the ejectors.

The steam-anode exhaust stream from the ejector is mixed with natural gas and then preheated to 919 °K by heat exchange with the hydrogen-rich fuel stream leaving the reformer.

Pre-reformer

The pre-reformer starts the steam-reforming of the feed hydrocarbon, only under adiabatic conditions. The process stream temperature drops from 919 °K to 799 °K as 23% of the methane in the feed is reformed.

Reformer

The reformer is a plate-type design of alternating reforming and heat transfer sections.

The partially reformed feed stream from the pre-reformer enters the IHI plate-type steam reformer where it is fully reformed to the hydrogen-rich product stream. The operating conditions at the reformer exit are 606 kPa and 1,061 °K.

The source of energy needed to sustain the reforming reactions is provided from combustor exhaust gas as it flows through the heat transfer sections of the reformer. The anode exhaust gas not recycled to the reformer is fed to the catalytic combustor, where residual CH₄, CO, and H₂ are combusted to provide the temperature and flow conditions necessary to transfer heat to the reformer.

Hydrogen-rich fuel from the reformer (reformate) is cooled to the fuel cell anode feed temperature in the reformer feed/effluent heat exchanger. The reformate then is piped to the fuel cell anode inlet.

Molten Carbonate Fuel Cell (MCFC)

The fuel and oxidant are consumed by charge transfer reactions in the MCFC stack. The reactions cause a potential difference between electrodes.

Anode

The hydrogen-rich fuel enters the anode at 950 °K, undergoes oxidation charge transfer reactions at the gas/electrode/electrolyte interface, and leaves the stack at 978 °K. Carbonate ions from the electrolyte are converted to carbon dioxide at the reaction sites, and leave the stack in the anode exhaust stream.

The stack consists of a number of cells arranged in parallel. Anode feed gas is distributed uniformly to each cell by the inlet manifold. Uniform distribution across the plate surface is achieved by providing channels in the anode compartment for the gas to flow through. Gas exiting the cell is collected in the outlet manifold.

At start-of-run conditions the cell voltage is 757 mV and the average current density is 200 mA/cm². The cells are electrically connected in series, and therefore the stack develops practical voltages.

Cathode

The cathode feed stream comes from the three recycle blowers. Each cathode recycle blower feeds three power module vessels for a total of 12 fuel cells. The feed into each cathode blower is a mixture of three streams. The first is reformer flue gas which for a single recycle blower is collected from the 12 reformers (total) in the three power module vessels. The second stream added before the cathode recycle blower inlet is the cathode exhaust recycle gas. The third stream added before the cathode blower inlet is the air from the gas turbine compressor section. The air provides an oxidant supply and the fuel cell stack cooling. The first two streams, when mixed together, provide a source of carbon dioxide to replenish the electrolyte after carbonate ion oxidation at the anode. The cathode recycle blower provides homogeneous mixing of the three streams mentioned above and minimizes stratification of a single stream.

The cathode feed stream enters the stack at 866 °K. Oxygen and carbon dioxide undergo reduction charge transfer reactions at the gas/electrode/electrolyte interface. The gas leaves the stack at 978 °K.

Cathode feed gas is distributed uniformly to each cell by an inlet manifold. Uniform distribution across the plate surface is achieved by providing channels in the cathode compartment for the gas to flow through. Gas exiting the cell is collected in the outlet manifold.

Recycle Blower

The energy difference between the oxidant/fuel heat-of-reaction and the useful work produced by the MCFC stack is released as heat. This heat must be removed to maintain isothermal operating conditions inside the stack.

Stack cooling is provided to a small extent by gas flow through the anode compartment. However, most of the cooling is provided by the flow of gas through the cathode compartment – approximately twelve times the mass flow of anode gas.

The recycle blower creates the cathode recycle stream, which, when added to air from the compressor and the processed feed stream, provides the gas flow needed to cool the stack.

Inverter

For utility systems, such as this program, the d.c. output from four MCFC fuel cells must be converted to utility-quality a.c. power.

The inverter converts the combined power module fuel cell output voltage of 800-900V dc into 4160 VAC, 3 phase, power for export to the grid, and for use by the plant auxiliaries. The multiple inverter power is collected in a common bus and is joined with the turbine generator power for feed into the step up transformer. The stepup transformer high side bushings are the primary utility interface point.

Gas-Turbine Generator

Cathode-exhaust gas from the thirty-six power modules combines into a single stream which leads to the turbine (expander) section of the gas turbine generator. The high pressure exhaust and heat energy is used to drive the gas turbine compressor section and alternator section. Exhaust gas leaves the turbine section and is ducted to the HRSG. Atmospheric air is compressed in the gas turbine compressor section to approximately 610 kPa. The gas turbine generator system produces 4160VAC for use in plant auxiliaries and the electric utility grid.

HRSG

Waste heat from the turbine exhaust is recovered in the HRSG unit by convective heat transfer. Heat is transferred to preheat natural gas, raise and superheat steam, preheat BFW, and preheat demineralized make-up water. The exhaust gas is discharged to the atmosphere at 446°K.

Design Requirements

A design specification for the plant appears in Table 6.

Table 6
Design Specification

Application	Commercial distributed power generation utility – grid connected.
Fuel	<p>Natural Gas</p> <p>Composition: - Mol%</p> <p>CH₄ 96.0</p> <p>N₂ 2.0</p> <p>CO₂ 2.0</p> <p>S 4 ppmv – mercaptan</p> <p>Supply pressure 15 psig</p> <p>Supply temperature 59°F</p>
Oxidant	<p>Air composition: -Mol%</p> <p>O₂ 20.7</p> <p>N₂ 78.0</p> <p>CO₂ 0.3</p> <p>Water 1.0 (60% relative humidity)</p> <p>Ambient air temperature 59°F</p> <p>Ambient air pressure 14.7 psia</p>
Performance Requirements	20 MW net AC,
Power System Design-Point Capacity	+/- 2 MW
Design-Point Efficiency	> 70% (net AC/LHV)
Power System Dispatch Mode	Base loaded
Power System Normal-Operation Turndown Requirement	None
Power system Overpower Requirement	None
Output Power Conditions	60 Hz, Utility-grid quality
Noise Control	Consistent with typical gas turbine power plant practice
Ambient Air Parameter Ranges	
Pressure	Sea level to 5,000 ft.
Temperature	-20°F to 120°F
Relative Humidity	0 to 100%

Power System Physical Design Objectives	
Installation	Outdoors
Transportation Options	Truck, sea, air, rail
Transportation Basis	500 miles
Plant Design Lifetime	Conventional power plant equipment – 25 years
Utilities	All utilities needed to support the plant will be available at the site
Power Plant Operation	
Annual Operating Time	Fifty weeks
Annual Planned Shutdown	Two weeks
Instrumentation and Control The I&C system design concept and cost will be based on the projected needs of mature fuel cell/gas turbine technologies and commercial power plant operation	
Economic Evaluation Parameters	
Power system Cost Estimation Basis	Mature technologies and integration techniques, not first-of-a-kind.
Cost Basis	Mid-1998 US Dollars
Fuel Cost	\$3.00/MMBtu (HHV)
Capital Charge Rate	15%
Capacity Factor	92%
COE Evaluation Method	Constant Dollars
Power Plant Optimization Basis	Consistent with >70% (LHV) efficiency, achieve a design-point COE that is 10-20% below the COE achieved by today's conventional-plant technology
Conventional Power Plant Basis - for COE comparison purposes	Comparably sized gas turbine cycle

3.B.2. Plant Performance Summary

3.B.2.a Fuel Cell Stacks

The rated fuel cell performance based on an ambient temperature of 288 °K, relative humidity of 60 percent, sea level elevation, and 101.3 kPa absolute atmospheric pressure at beginning and end of fuel cell life appears below:

Parameter	Beginning of Life	End of Life
Number of Stacks	36	36
Current Density (mA/cm squared)	200	200
Millivolts/Cell	758	700
Power Density (kW/meter squared)	1.52	1.40
kW/Stack	496	458
kW Total from all Stacks	17,856	16,488
Operating Pressure (kPa)	605	605
Fuel Utilization-Per Pass (%)	80.4	80.3
Fuel Utilization-Overall (%)	86.2	86.5

3.B.2.b Turbine Generator

The gas turbine generator predicted performance at 288 °K, 101kPa absolute, and sea level conditions for the fuel cell beginning of life and end of life conditions appears below:

Parameter	Beginning of Fuel Cell Life	End of Fuel Cell Life
Compressor		
Compressor Inlet Temperature (°K)	288	288
Compressor Inlet Pressure (kPa)	101	101
Compressor Mass Flow (kg/sec)	16.8	19.2
Compressor Outlet Pressure (kPa)	621	621
Compressor Outlet Temperature (°K)	518	518
Compression Ratio	6.16	6.16
Compressor Efficiency (%)	84	84
Turbine		
Turbine Inlet Temperature (°K)	978	978
Turbine Inlet Pressure (kPa)	594	594
Turbine Mass Flow (kg/sec)	19.1	21.6
Turbine Outlet Pressure (kPa)	105	105
Turbine Outlet Temperature (°K)	671	669
Turbine Efficiency (%)	91.9	91.9
Generator		
Generator Efficiency (%)	98.5	98.5
Power Conditioner		
Power Conditioner Output Voltage (V)	4160	4160
Power Conditioner Output Current (A)	476	530
Power Conditioner Output Power (kW)	3429	3821
Power Conditioner Output Frequency (Hz)	60	60
Power Conditioner Output Phases	3	3
Power Conditioner Efficiency (%)	97.5	97.5

3.B.2.c Overall Plant

Plant performance summary tables are provided for the following cases:

Table 7 – Major Subsystem Design Parameters and Operating Conditions. This table represents expected operating conditions for major plant components at SOR and EOR fuel cell conditions for a plant at sea level.

Table 8 – Start of Run (SOR), sea level summary performance table. This table presents performance data for three different ambient air temperatures, at SOR, full-load operation and for a plant site at sea level.

Table 9 – Start of Run (SOR), 5,000 ft elevation summary performance table. This table presents performance data for three different ambient air temperatures, at SOR, full-load operation and for a plant site at 5,000 ft elevation.

Table 10 – End of Run (EOR), sea level summary performance table. This table presents performance data for two ambient air temperatures, at EOR, full-load operation and for a plant site at sea level.

TABLE 7
Major Subsystem
Design Parameters and Operating Conditions
Site Elevation – Sea Level

	SQR	EOR
Fuel Processor - Reformer	Type: External/Separate - IHI Gas-Heated Plate-Type	
Steam:Hydrocarbon ratio	4.68	5.3
Steam:Total Carbon ratio	1.43	1.5
Operating Pressure, kPa	606	606
Exit Temperature, K	1,061	1,061
Process Duty, kJ/h	1,018,806	1,026,361
Number of Units	36	36
Fuel Cell Stack Design	Type: MC-Power Molten Carbonate, Co-Flow	
Operating Pressure, kPa	605	605
Current Density, mA/cm2	200	200
Cell Voltage, mV	758	700
Power Density, kW/m2	1.52	1.40
Fuel Utilization - per pass, %	80.4	80.3
Fuel Utilization - overall, %	86.2	86.5
Number of Stacks Required	36	36
Inverter Design	Type: Solid-State	
Input Voltage, Volts	227	210
Input Current, Amps DC	2,180	2,180
Output Voltage, Volts	4,160	4,160
Output Electric Power, kW	479	443
Power Conversion Efficiency, %	97.5	97.5
Number of Units Required	9	9
Recycle Gas Blower		
Adiabatic Efficiency, %	75	75
Differential Pressure, kPa	6.2	6.2
Shaft Horse Power, kW	108	119
Motor Efficiency, %	92	92
Number of Units	3	3
Gas Turbine Design	Type: Rolls-Royce Allison	
Compression Ratio	6.16	6.16
Compressor Air Flow, kg/s	16.81	19.19
Compressor Efficiency, %	84	84
Turbine Efficiency, %	91.9	91.9
Generator Efficiency, %	98.5	98.5
Net Power Production, kW	20,115	19,159
Overall Electrical Efficiency	70.11	66.78

TABLE 8
Performance Summary
Site Elevation – Sea Level

~~Start-of-Run Operation~~

Ambient Temperature, °K	288	244	322	322
Relative Humidity	60%	0%	0%	100%
AC Power from Power Modules, kW	17,256	17,256	17,256	17,256
AC Power from Gas turbine Generator, kW	3,429	3,603	3,246	3,242
Gross Power Production, kW	20,685	20,859	20,502	20,498
Auxiliary Loads, kW	<u>570</u>	<u>571</u>	<u>571</u>	<u>563</u>
Net Power Production, kW	20,115	20,288	19,931	19,935
Natural Gas Consumption, kJ/h (LHV)	1.0328E+08	1.0328E+08	1.0328E+08	1.0328E+08
Fuel Fired in Gas Turbine, kJ/h (LHV)	0	0	0	0
Fuel Fired in HRSG, kJ/h (LHV)	0	0	0	0
Overall Electrical Efficiency (LHV), %	<u>70.11</u>	<u>70.72</u>	<u>69.47</u>	<u>69.49</u>

TABLE 9
Performance Summary
Site Elevation – 5,000'

~~Start-of-Run Operation~~

Ambient Temperature, °K	288	244	322	322
Relative Humidity	60%	0%	0%	100%
AC Power from Power Modules, kW	17,256	17,256	17,256	17,256
AC Power from Gas turbine Generator, kW	3,099	3,368	2,819	2,845
Gross Power Production, kW	20,355	20,624	20,075	20,101
Auxiliary Loads, kW	<u>570</u>	<u>571</u>	<u>571</u>	<u>563</u>
Net Power Production, kW	19,785	20,052	19,504	19,538
Natural Gas Consumption, kJ/h (LHV)	1.0328E+08	1.0328E+08	1.0328E+08	1.0328E+08
Fuel Fired in Gas Turbine, kJ/h (LHV)	0	0	0	0
Fuel Fired in HRSG, kJ/h (LHV)	0	0	0	0
Overall Electrical Efficiency (LHV), %	<u>68.96</u>	<u>69.90</u>	<u>67.98</u>	<u>68.10</u>

TABLE 10
Performance Summary
Site Elevation – Sea Level

~~End-of-Run Operation~~

Ambient Temperature, °K	288	322
Relative Humidity	60%	0%
AC Power from Power Modules, kW	15,946	15,946
AC Power from Gas turbine Generator, kW	3,821	3,595
Gross Power Production, kW	19,767	19,541
Auxiliary Loads, kW	<u>608</u>	<u>610</u>
Net Power Production, kW	19,159	18,931
Natural Gas Consumption, kJ/h (LHV)	1.0328E+08	1.0328E+08
Fuel Fired in Gas Turbine, kJ/h (LHV)	0	0
Fuel Fired in HRSG, kJ/h (LHV)	0	0
Overall Electrical Efficiency (LHV), %	<u>66.78</u>	<u>65.99</u>

3.B.3. Process Flow Diagram

Start-of-Run (SOR) Operation

DWG B-001 is the process flow diagram (PFD) for the HEFPP overall facility. It depicts the design and control concepts, process equipment, and presents the flows and conditions for all streams entering and exiting the facility, including electric power.

DWG B-002 is the process flow diagram (PFD) for a single power module. It depicts the design and control concepts, process equipment, and presents the flows and conditions for all streams entering and exiting a single power module contained within a power-module vessel.

Operating conditions shown on these drawings are for full-load, SOR operation at 15°C ambient air temperature and at sea level.

The process stream flows and conditions for the following other operating cases are shown on the PFDs.

DRAWING B-001

TABLE 11
Material Balance – Full-Load, SOR – Sea Level 15°C/-29°C Ambient Air (Page 1 of 2)

FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR																	
Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
Stream Description:	Total Natural Gas Feed	Demin. Feedwater	Total Air Feed	Process Air	Flue Gas Vent to Atmosphere	Desulfurized Natural Gas	Superheated Steam	Nat. Gas to Power Train	Shr'd. Steam to Power Train	Process Air to Power Train	Process Air to Power Train	Recycle Blower Discharge	Cathode Exhaust from Power Train	Turbine Inlet	Turbine Exhaust	Turbine Nat. Gas Fuel	Steam to Deaerator
Temperature, K	426	289	288	518	446	616	536	616	536	518	518	866	978	978	671	426	536
Pressure, kPa	749	276	101	621	101.4	715	886	715	881	616	616	605	603	594	105	749	886
Mass Flow, kg/h	2,257	5,933	60,507	60,507	68,688	2,257	6,124	752	1,975	341	19,828	105,577	22,896	68,688	68,688	-	200
Std. Volume Flow, m3/h (0°C, 101.3 kPa)	3,004	-	47,113	47,113	57,487	3,004	7,619	1,001	2,457	266	15,439	85,949	19,162	57,487	57,487	-	248
Molar Flow, kg-mole/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CH4	128.68	-	-	-	-	128.68	-	42.89	-	-	-	-	-	-	-	-	-
CO	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CO2	2.68	-	6.31	6.31	137.66	2.68	-	0.89	-	0.04	2.07	339.41	45.89	137.66	137.66	-	-
H2	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H2O	-	329.36	21.02	21.02	607.22	-	339.93	-	109.62	0.12	6.89	856.26	202.41	607.22	607.22	-	11.08
N2	2.68	-	1,639.52	1,639.52	1,642.17	2.68	-	0.89	-	9.25	537.26	2,315.66	547.39	1,642.17	1,642.17	-	-
O2	-	-	435.10	435.10	177.75	-	-	-	-	2.45	142.58	323.29	59.25	177.75	177.75	-	-
Total, kg-mole/hr	134.04	329.36	2,101.94	2,101.94	2,564.80	134.04	339.93	44.68	109.62	11.86	688.79	3,834.63	854.93	2,564.80	2,564.80	-	11.08
Density, kg/m3	3.58	1,014.84	1.21	4.14	0.73	2.35	3.69	2.35	3.67	4.11	2.31	1.98	1.98	1.95	0.50	3.58	3.69
Mole %	16.84	18.02	28.79	28.79	26.78	16.84	18.02	16.84	18.02	28.79	28.79	27.53	26.78	26.78	16.84	18.02	18.02
Z Factor	0.996	-	0.999	1.002	0.999	1.001	0.972	1.001	0.972	1.002	1.002	1.001	1.001	1.001	1.000	0.996	0.972
FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / -29°C AMBIENT AIR																	
Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
Stream Description:	Total Natural Gas Feed	Demin. Feedwater	Total Air Feed	Process Air	Flue Gas Vent to Atmosphere	Desulfurized Natural Gas	Superheated Steam	Nat. Gas to Power Train	Shr'd. Steam to Power Train	Process Air to Power Train	Process Air to Power Train	Recycle Blower Discharge	Cathode Exhaust from Power Train	Turbine Inlet	Turbine Exhaust	Turbine Nat. Gas Fuel	Steam to Deaerator
Temperature, K	426	289	244	441	450	616	536	616	536	441	441	866	978	978	671	426	536
Pressure, kPa	749	276	101	621	101.4	715	886	715	881	616	616	605	603	594	105	749	886
Mass Flow, kg/h	2,257	5,190	53,802	53,802	61,239	2,257	5,356	752	1,727	343	17,591	106,199	20,413	61,239	61,239	-	175
Std. Volume Flow, m3/h (0°C, 101.3 kPa)	3,004	-	41,734	41,734	51,184	3,004	6,663	1,001	2,149	266	13,646	86,359	17,061	51,184	51,184	-	217
Molar Flow, kg-mole/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CH4	128.68	-	-	-	-	128.68	-	42.89	-	-	-	-	-	-	-	-	-
CO	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CO2	2.68	-	5.64	5.64	137.00	2.68	-	0.89	-	0.04	1.84	363.37	45.67	137.00	137.00	-	-
H2	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H2O	-	288.08	-	-	544.93	-	297.28	-	95.86	-	-	867.42	181.64	544.93	544.93	-	9.70
N2	2.68	-	1,467.00	1,467.00	1,469.68	2.68	-	0.89	-	9.34	479.66	2,339.41	489.89	1,469.68	1,469.68	-	-
O2	-	-	389.32	389.32	131.97	-	-	-	-	2.48	127.29	282.71	43.99	131.97	131.97	-	-
Total, kg-mole/hr	134.04	288.08	1,861.96	1,861.96	2,283.58	134.04	297.28	44.68	95.86	11.86	608.80	3,852.91	761.19	2,283.58	2,283.58	-	9.70
Density, kg/m3	3.58	1,014.84	1.43	4.88	0.73	2.35	3.69	2.35	3.67	4.84	4.84	2.31	1.99	1.96	0.50	3.58	3.69
Mole %	16.84	18.02	28.90	28.90	26.82	16.84	18.02	16.84	18.02	28.90	28.90	27.56	26.82	26.82	16.84	18.02	18.02
Z Factor	0.996	-	0.999	1.001	0.999	1.001	0.972	1.001	0.972	1.001	1.001	1.001	1.001	1.001	1.000	0.996	0.972
MATERIAL BALANCE - HEPP FACILITY										MC-POWER CORPORATION				JOB NO.: 24024-000			
										20 MW HEPP PROGRAM				DOCUMENT NO.			
														REV.			
														SHEET 1 OF 2

TABLE 11 cont.
Material Balance - Full Load, SOR – Sea Level 49°C (0% /100% Rel. Humity) (Page 2 of 2)

FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR																		
Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	
Stream Description:	Total Natural Gas Feed	Demin. Feedwater	Total Air Feed	Process Air	Flue Gas Vent to Atmosphere	Desulfurized Natural Gas	Superheated Steam	Nat. Gas to Power Train	Sh't'd. Steam to Power Train	Process Air to Power Train	Process Air to Power Train	Recycle Blower Discharge	Cathode Exhaust from Power Train	Turbine Inlet	Turbine Exhaust	Turbine Nat. Gas Fuel	Steam to Dearator	
Temperature, K	426	289	322	577	449	616	536	616	536	577	577	866	978	978	669	426	536	
Pressure, kPa	749	276	101	621	101.4	715	886	715	881	616	616	605	603	594	105	749	886	
Mass Flow, kg/h	2,257	6,465	68,293	68,293	77,006	2,257	6,673	752	2,152	343	22,422	106,497	25,669	77,006	77,006	-	217	
Std. Volume Flow, m3/h (0°C, 101.3 kPa)	3,004	-	52,975	52,975	64,012	3,004	8,302	1,001	2,677	266	17,393	86,163	21,337	64,012	64,012	-	270	
Molar Flow, kg-mole/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	
CH4	128.68	-	-	-	-	128.68	-	42.89	-	-	-	-	-	-	-	-	-	
CO	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	
CO2	2.68	-	7.16	7.16	138.52	2.68	-	0.89	-	0.04	2.35	321.18	46.17	138.52	138.52	-	-	
H2	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	
H2O	-	358.85	-	-	615.68	-	370.40	-	119.45	-	-	781.75	205.23	615.68	615.68	-	12.06	
N2	2.68	-	1,862.13	1,862.13	1,864.84	2.68	-	0.89	-	9.34	611.37	2,367.87	621.61	1,864.84	1,864.84	-	-	
O2	-	-	494.18	494.18	236.83	-	-	-	-	2.48	162.25	373.36	78.94	236.83	236.83	-	-	
Total, kg-mole/hr	134.04	358.85	2,363.47	2,363.47	2,855.88	134.04	370.40	44.68	119.45	11.86	775.97	3,844.17	951.96	2,855.88	2,855.88	-	12.06	
Density, kg/m3	3.58	1,014.84	1.09	3.73	0.73	2.35	3.69	2.35	3.67	3.70	3.70	2.32	2.00	1.97	0.51	3.58	3.69	
Mole Wt.	16.84	18.02	28.90	28.90	26.96	16.84	18.02	16.84	18.02	28.90	28.90	27.70	26.96	26.96	26.96	16.84	18.02	
Z Factor	0.996	-	1.000	1.002	0.999	1.001	0.972	1.001	0.972	1.002	1.002	1.001	1.001	1.001	1.000	0.996	0.972	
FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (100% REL. HUMIDITY) AMBIENT AIR																		
Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	
Stream Description:	Total Natural Gas Feed	Demin. Feedwater	Total Air Feed	Process Air	Flue Gas Vent to Atmosphere	Desulfurized Natural Gas	Superheated Steam	Nat. Gas to Power Train	Sh't'd. Steam to Power Train	Process Air to Power Train	Process Air to Power Train	Recycle Blower Discharge	Cathode Exhaust from Power Train	Turbine Inlet	Turbine Exhaust	Turbine Nat. Gas Fuel	Steam to Dearator	
Temperature, K	426	289	322	571	449	616	536	616	536	571	571	866	978	978	674	426	536	
Pressure, kPa	749	276	101	621	101.4	715	886	715	881	616	616	605	603	594	105	749	886	
Mass Flow, kg/h	2,257	6,546	62,595	62,595	71,379	2,257	6,757	752	2,179	328	20,534	100,244	23,793	71,379	71,379	-	220	
Std. Volume Flow, m3/h (0°C, 101.3 kPa)	3,004	-	50,748	50,748	61,886	3,004	8,407	1,001	2,711	266	16,650	84,238	20,629	61,886	61,886	-	274	
Molar Flow, kg-mole/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	
CH4	128.68	-	-	-	-	128.68	-	42.89	-	-	-	-	-	-	-	-	-	
CO	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	
CO2	2.68	-	6.07	6.07	137.43	2.68	-	0.89	-	0.03	1.99	321.51	45.81	137.43	137.43	-	-	
H2	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	
H2O	-	363.38	260.81	260.81	881.05	-	375.09	-	120.96	1.37	85.57	1,129.70	293.68	881.05	881.05	-	12.21	
N2	2.68	-	1,578.38	1,578.38	1,581.06	2.68	-	0.89	-	8.27	517.86	2,027.29	527.02	1,581.06	1,581.06	-	-	
O2	-	-	418.88	418.88	161.52	-	-	-	-	2.19	137.43	279.76	53.84	161.52	161.52	-	-	
Total, kg-mole/hr	134.04	363.38	2,264.14	2,264.14	2,761.06	134.04	375.09	44.68	120.96	11.86	742.86	3,758.26	920.35	2,761.06	2,761.06	-	12.21	
Density, kg/m3	3.58	1,014.84	1.04	3.61	0.70	2.35	3.69	2.35	3.67	3.58	3.58	2.24	1.91	1.89	0.48	3.58	3.69	
Mole Wt.	16.84	18.02	27.64	27.64	25.85	16.84	18.02	16.84	18.02	27.64	27.64	26.67	25.85	25.85	25.85	16.84	18.02	
Z Factor	0.996	-	0.999	1.001	0.999	1.001	0.972	1.001	0.972	1.001	1.001	1.001	1.001	1.001	1.000	0.996	0.972
MATERIAL BALANCE - HEFPP FACILITY										MC-POWER CORPORATION				JOB NO.: 24024-000				
										20 MW HEFPP PROGRAM				DOCUMENT NO.				REV.
														SHEET 2 OF 2				

TABLE 12
Material Balance – Full Load SOR – 5,000' Elev./15°C/-29°C Ambient Air (Page 1 of 2)

FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 15°C AMBIENT AIR																	
Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
Stream Description:	Total Natural Gas Feed	Demin. Feedwater	Total Air Feed	Process Air	Flue Gas Vent to Atmosphere	Desulfurized Natural Gas	Superheated Steam	Nat. Gas to Power Train	Sh'd. Steam to Power Train	Process Air to Power Train	Process Air to Power Train	Recycle Blower Discharge	Cathode Exhaust from Power Train	Turbine Inlet	Turbine Exhaust	Turbine Nat. Gas Fuel	Steam to Dearator
Temperature, K	426	289	288	543	447	616	536	616	536	543	543	867	978	978	670	426	536
Pressure, kPa	749	276	86	621	101.4	715	886	715	881	616	616	605	603	594	105	749	886
Mass Flow, kg/h	2,257	6,146	63,408	63,408	71,803	2,257	6,343	752	2,046	341	20,795	105,767	23,934	71,803	71,803	-	207
Std. Volume Flow, m ³ /h (0°C, 101.3 kPa)	3,004	-	49,372	49,372	60,011	3,004	7,892	1,001	2,545	266	16,132	85,996	20,004	60,011	60,011	-	257
Molar Flow, kg-mole/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CH ₄	128.68	-	-	-	-	128.68	-	42.89	-	-	-	-	-	-	-	-	-
CO	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CO ₂	2.68	-	6.61	6.61	137.96	2.68	-	0.89	-	0.04	2.17	331.76	45.99	137.96	137.96	-	-
H ₂	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H ₂ O	-	341.15	22.03	22.03	619.98	-	352.12	-	113.55	0.12	7.22	837.96	206.66	619.98	619.98	-	11.47
N ₂	2.68	-	1,718.13	1,718.13	1,720.83	2.68	-	0.89	-	9.25	563.46	2,325.86	573.61	1,720.83	1,720.83	-	-
O ₂	-	-	455.96	455.96	198.63	-	-	-	-	2.45	149.53	341.12	66.21	198.63	198.63	-	-
Total, kg-mole/hr	134.04	341.15	2,202.73	2,202.73	2,677.41	134.04	352.12	44.68	113.55	11.86	722.39	3,636.70	892.47	2,677.41	2,677.41	-	11.47
Density, kg/m ³	3.58	1,014.84	1.04	3.95	0.73	2.35	3.69	2.35	3.67	3.92	3.92	2.31	1.99	1.96	0.50	3.58	3.69
Mole %	16.84	18.02	28.79	28.79	26.82	16.84	18.02	16.84	18.02	28.79	28.79	27.57	26.82	26.82	26.82	16.84	18.02
Z Factor	0.996	-	0.999	1.002	0.999	1.001	0.972	1.001	0.972	1.002	1.002	1.001	1.001	1.001	1.000	0.996	0.972
FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / -29°C AMBIENT AIR																	
Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
Stream Description:	Total Natural Gas Feed	Demin. Feedwater	Total Air Feed	Process Air	Flue Gas Vent to Atmosphere	Desulfurized Natural Gas	Superheated Steam	Nat. Gas to Power Train	Sh'd. Steam to Power Train	Process Air to Power Train	Process Air to Power Train	Recycle Blower Discharge	Cathode Exhaust from Power Train	Turbine Inlet	Turbine Exhaust	Turbine Nat. Gas Fuel	Steam to Dearator
Temperature, K	426	289	244	463	449	616	536	616	536	463	463	866	978	978	671	426	536
Pressure, kPa	749	276	86	621	101.4	715	886	715	881	616	616	605	603	594	105	749	886
Mass Flow, kg/h	2,257	5,361	55,659	55,659	63,269	2,257	5,533	752	1,784	343	18,210	106,273	21,090	63,269	63,269	-	181
Std. Volume Flow, m ³ /h (0°C, 101.3 kPa)	3,004	-	43,175	43,175	52,838	3,004	6,884	1,001	2,220	266	14,126	86,357	17,613	52,838	52,838	-	225
Molar Flow, kg-mole/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CH ₄	128.68	-	-	-	-	128.68	-	42.89	-	-	-	-	-	-	-	-	-
CO	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CO ₂	2.68	-	5.84	5.84	137.19	2.68	-	0.89	-	0.04	1.91	356.83	45.73	137.19	137.19	-	-
H ₂	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H ₂ O	-	297.61	-	-	554.45	-	307.13	-	99.04	-	-	854.91	184.82	554.45	554.45	-	10.02
N ₂	2.68	-	1,517.65	1,517.65	1,520.33	2.68	-	0.89	-	9.34	496.54	2,344.20	506.78	1,520.33	1,520.33	-	-
O ₂	-	-	402.76	402.76	145.41	-	-	-	-	2.48	131.77	296.86	48.47	145.41	145.41	-	-
Total, kg-mole/hr	134.04	297.61	1,926.24	1,926.24	2,357.39	134.04	307.13	44.68	99.04	11.86	630.23	3,852.79	785.80	2,357.39	2,357.39	-	10.02
Density, kg/m ³	3.58	1,014.84	1.23	4.65	0.73	2.35	3.69	2.35	3.67	4.61	4.61	2.31	1.99	1.96	0.50	3.58	3.69
Mole %	16.84	18.02	28.90	28.90	26.84	16.84	18.02	16.84	18.02	28.90	28.90	27.58	26.84	26.84	26.84	16.84	18.02
Z Factor	0.996	-	0.999	1.001	0.999	1.001	0.972	1.001	0.972	1.001	1.001	1.001	1.001	1.001	1.000	0.996	0.972
MATERIAL BALANCE - HEPPP FACILITY											MC-POWER CORPORATION						
											20 MW HEPPP PROGRAM						
											JOB NO.: 24024-000						
											DOCUMENT NO.						
											REV.						
											SHEET 1 OF 2						

TABLE 12
Material Balance – Full Load SOR – 5,000' Elev./49°(0%/100% Rel. Humidity) (Page 2 of 2)

FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (0% REL. HUMIDITY) AMBIENT AIR																	
Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
Stream Description:	Total Natural Gas Feed	Demin. Feedwater	Total Air Feed	Process Air	Flue Gas Vent to Atmosphere	Desulfurized Natural Gas	Superheated Steam	Nat. Gas to Power Train	Sh't'd. Steam to Power Train	Process Air to Power Train	Process Air to Power Train	Recycle Blower Discharge	Cathode Exhaust from Power Train	Turbine Inlet	Turbine Exhaust	Turbine Nat. Gas Fuel	Steam to Dearthor
Temperature, K	426	289	322	605	450	616	536	616	536	605	605	866	978	978	668	426	536
Pressure, kPa	749	276	86	621	101.4	715	886	715	881	616	616	605	603	594	105	749	886
Mass Flow, kg/h	2,257	6,791	72,373	72,373	81,410	2,257	7,011	752	2,261	343	23,782	106,589	27,137	81,410	81,410	-	228
Std. Volume Flow, m3/h (0°C, 101.3 kPa)	3,004	-	56,140	56,140	67,581	3,004	8,722	1,001	2,813	266	18,447	86,130	22,527	67,581	67,581	-	284
Molar Flow, kg-mole/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CH4	128.68	-	-	-	-	128.68	-	42.89	-	-	-	-	-	-	-	-	-
CO	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CO2	2.68	-	7.59	7.59	138.95	2.68	-	0.89	-	0.04	2.49	312.34	46.32	138.95	138.95	-	-
H2	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H2O	-	376.99	-	-	633.85	-	389.15	-	125.50	-	-	762.01	211.28	633.85	633.85	-	12.66
N2	2.68	-	1,973.38	1,973.38	1,975.98	2.68	-	0.89	-	9.34	648.45	2,375.51	658.66	1,975.98	1,975.98	-	-
O2	-	-	523.70	523.70	266.33	-	-	-	-	2.48	172.09	392.82	88.78	266.33	266.33	-	-
Total, kg-mole/hr	134.04	376.99	2,504.67	2,504.67	3,015.11	134.04	389.15	44.68	125.50	11.86	823.03	3,842.67	1,005.04	3,015.11	3,015.11	-	12.66
Density, kg/m3	3.58	1,014.84	0.93	3.56	0.73	2.35	3.69	2.35	3.67	3.53	3.53	2.33	2.00	1.97	0.51	3.58	3.69
Mole Wt.	16.84	18.02	28.90	28.90	27.00	16.84	18.02	16.84	18.02	28.90	28.90	27.74	27.00	27.00	27.00	16.84	18.02
Z Factor	0.996	-	1.000	1.002	0.999	1.001	0.972	1.001	0.972	1.002	1.002	1.001	1.001	1.001	1.000	0.996	0.972
FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (100% REL. HUMIDITY) AMBIENT AIR																	
Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
Stream Description:	Total Natural Gas Feed	Demin. Feedwater	Total Air Feed	Process Air	Flue Gas Vent to Atmosphere	Desulfurized Natural Gas	Superheated Steam	Nat. Gas to Power Train	Sh't'd. Steam to Power Train	Process Air to Power Train	Process Air to Power Train	Recycle Blower Discharge	Cathode Exhaust from Power Train	Turbine Inlet	Turbine Exhaust	Turbine Nat. Gas Fuel	Steam to Dearthor
Temperature, K	426	289	322	598	448	616	536	616	536	598	598	866	978	978	674	426	536
Pressure, kPa	749	276	86	621	101.4	715	886	715	881	616	616	605	603	594	105	749	886
Mass Flow, kg/h	2,257	6,914	66,028	66,028	75,191	2,257	7,137	752	2,302	328	21,682	100,225	25,064	75,191	75,191	-	232
Std. Volume Flow, m3/h (0°C, 101.3 kPa)	3,004	-	53,541	53,541	65,136	3,004	8,880	1,001	2,864	266	17,581	84,155	21,712	65,136	65,136	-	289
Molar Flow, kg-mole/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CH4	128.68	-	-	-	-	128.68	-	42.89	-	-	-	-	-	-	-	-	-
CO	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CO2	2.68	-	6.40	6.40	137.77	2.68	-	0.89	-	0.03	2.10	312.96	45.92	137.77	137.77	-	-
H2	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H2O	-	383.79	275.16	275.16	915.80	-	396.18	-	127.76	1.37	90.35	1,114.51	305.27	915.80	915.80	-	12.89
N2	2.68	-	1,665.22	1,665.22	1,667.91	2.68	-	0.89	-	8.27	546.81	2,029.82	556.97	1,667.91	1,667.91	-	-
O2	-	-	441.92	441.92	184.57	-	-	-	-	2.19	145.11	297.26	61.52	184.57	184.57	-	-
Total, kg-mole/hr	134.04	383.79	2,388.71	2,388.71	2,906.04	134.04	396.18	44.68	127.76	11.86	784.38	3,754.55	968.68	2,906.04	2,906.04	-	12.89
Density, kg/m3	3.58	1,014.84	0.89	3.45	0.70	2.35	3.69	2.35	3.67	3.42	3.42	2.24	1.92	1.89	0.48	3.58	3.69
Mole Wt.	16.84	18.02	27.64	27.64	25.87	16.84	18.02	16.84	18.02	27.64	27.64	26.69	25.87	25.87	25.87	16.84	18.02
Z Factor	0.996	-	0.999	1.001	0.999	1.001	0.972	1.001	0.972	1.001	1.001	1.001	1.001	1.001	1.000	0.996	0.972
<div> <div> <div>MATERIAL BALANCE - HEFPP FACILITY</div> <div>MC-POWER CORPORATION</div> <div>20 MW HEFPP PROGRAM</div> </div> <div> <div>JOB NO.: 24024-000</div> <div>DOCUMENT NO.</div> <div>SHEET 2 OF 2</div> </div> <div>REV.</div> </div>																	

TABLE 13
Material Balance – Full Load EOR – Sea Level 15°C/49°C (0% Rel. Humidity) Ambient Air

FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR																		
Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	
Stream Description:	Total Natural Gas Feed	Demin. Feedwater	Total Air Feed	Process Air	Flue Gas Vent to Atmosphere	Desulfurized Natural Gas	Superheated Steam	Nat. Gas to Power Train	Sh't'd. Steam to Power Train	Process Air to Power Train	Process Air to Power Train	Recycle Blower Discharge	Cathode Exhaust from Power Train	Turbine Inlet	Turbine Exhaust	Turbine Nat. Gas Fuel	Steam to Dearator	
Temperature, K	426	289	289	518	449	616	536	616	536	518	518	866	978	978	669	426	536	
Pressure, kPa	749	276	101	621	101.4	715	886	715	881	616	616	605	603	594	105	749	886	
Mass Flow, kg/h	2,257	6,587	69,074	69,074	77,909	2,257	6,799	752	2,193	341	22,683	117,055	25,970	77,909	77,909	-	221	
Std. Volume Flow, m3/h (0°C, 101.3 kPa)	3,004		53,783	53,783	64,972	3,004	8,460	1,001	2,728	266	17,662	95,232	21,657	64,972	64,972	-	275	
Molar Flow, kg-mole/hr																		
CH4	128.68	-	-	-	-	128.68	-	42.89						-	-	-		
CO	-	-	-	-	-	-	-	-						-	-	-		
CO2	2.68	-	7.20	7.20	138.57	2.68	-	0.89		0.04	2.36	337.98	46.19	138.57	138.57	-		
H2	-	-	-	-	-	-	-	-						-	-	-		
H2O	-	365.65	24.00	24.00	646.52	-	377.43		121.72	0.12	7.88	899.03	215.51	646.52	646.52	-	12.28	
N2	2.68	-	1,871.63	1,871.63	1,874.29	2.68	-	0.89		9.25	614.63	2,606.32	624.76	1,874.29	1,874.29	-		
O2	-	-	496.70	496.70	239.33	-	-	-		2.45	163.11	405.44	79.78	239.33	239.33	-		
Total, kg-mole/hr	134.04	365.65	2,399.53	2,399.53	2,898.71	134.04	377.43	44.68	121.72	11.86	787.99	4,248.77	966.24	2,898.71	2,898.71	-	12.28	
Density, kg/m3	3.58	1,014.84	1.21	4.14	0.73	2.35	3.69	2.35	3.67	4.11	4.11	2.31	1.99	1.96	0.51	3.58	3.69	
Mole Vt.	16.84	18.02	28.79	28.79	26.88	16.84	18.02	16.84	18.02	28.79	27.55	27.55	26.88	26.88	26.88	16.84	18.02	
Z Factor	0.996		0.999	1.002	0.999	1.001	0.972	1.001	0.972	1.002	1.002	1.001	1.001	1.001	1.000	0.996	0.972	
FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR																		
Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	
Stream Description:	Total Natural Gas Feed	Demin. Feedwater	Total Air Feed	Process Air	Flue Gas Vent to Atmosphere	Desulfurized Natural Gas	Superheated Steam	Nat. Gas to Power Train	Sh't'd. Steam to Power Train	Process Air to Power Train	Process Air to Power Train	Recycle Blower Discharge	Cathode Exhaust from Power Train	Turbine Inlet	Turbine Exhaust	Turbine Nat. Gas Fuel	Steam to Dearator	
Temperature, K	426	289	322	577	456	616	536	616	536	577	577	866	978	978	667	426	536	
Pressure, kPa	749	276	101	621	101.4	715	886	715	881	616	616	605	603	594	105	749	886	
Mass Flow, kg/h	2,257	7,035	78,300	78,300	87,582	2,257	7,262	752	2,342	343	25,757	118,310	29,194	87,582	87,582	-	236	
Std. Volume Flow, m3/h (0°C, 101.3 kPa)	3,004		60,737	60,737	72,483	3,004	9,035	1,001	2,914	266	19,980	95,582	24,161	72,483	72,483	-	294	
Molar Flow, kg-mole/hr																		
CH4	128.68	-	-	-	-	128.68	-	42.89						-	-	-		
CO	-	-	-	-	-	-	-	-						-	-	-		
CO2	2.68	-	8.21	8.21	139.57	2.68	-	0.89		0.04	2.70	319.94	46.52	139.57	139.57	-		
H2	-	-	-	-	-	-	-	-						-	-	-		
H2O	-	390.50	-	-	647.35	-	403.11		130.00	-	-	810.02	215.78	647.35	647.35	-	13.11	
N2	2.68	-	2,134.98	2,134.98	2,137.66	2.68	-	0.89		9.34	702.32	2,674.82	712.55	2,137.66	2,137.66	-		
O2	-	-	566.59	566.59	309.24	-	-	-		2.48	186.38	459.59	103.08	309.24	309.24	-		
Total, kg-mole/hr	134.04	390.50	2,709.78	2,709.78	3,233.82	134.04	403.11	44.68	130.00	11.86	891.41	4,264.37	1,077.94	3,233.82	3,233.82	-	13.11	
Density, kg/m3	3.58	1,014.84	1.09	3.73	0.72	2.35	3.69	2.35	3.67	3.70	3.70	2.33	2.01	1.98	0.51	3.58	3.69	
Mole Vt.	16.84	18.02	28.90	28.90	27.08	16.84	18.02	16.84	18.02	28.90	28.90	27.74	27.08	27.08	27.08	16.84	18.02	
Z Factor	0.996		1.000	1.002	1.000	1.001	0.972	1.001	0.972	1.002	1.002	1.001	1.001	1.001	1.001	1.000	0.996	0.972
<div> <div> MATERIAL BALANCE - HEFPP FACILITY </div> <div> MC-POWER CORPORATION 20 MW HEFPP PROGRAM </div> <div> JOB NO.: 24024-000 DOCUMENT NO. SHEET 1 OF 2 </div> <div> REV. </div> </div>																		

DRAWING B-002

TABLE 14
Material Balance – Full Load SOR – Sea Level/15°C/-29°C Ambient Air (Page 1 of 2)

FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR

Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
Stream Description:	Natural Gas to PM #1	Steam to PM #1	Anode Recycle	Mixed Feed to Exchanger	Reformer Feed	Reformer Exhaust (Reformate)	Anode Feed	Anode Exhaust	Anode-Ehst. Fuel to Combustor	Air to Combustor	Cathode Recycle to Combustor	Combustor Exhaust	Reformer Flue Gas	Cathode Recycle (Bypass to Blower)	PM #1 Gas to Blower	Cathode Feed	Cathode Exhaust	Cathode Recycle	PM #1 Gas to Gas Turbine
Temperature, K	616	536	978	801	799	1,061	950	978	978	518	978	1,072	893	978	329	866	978	978	978
Pressure, kPa	698	872	604.1	609.3	608.6	606.3	604.8	604.4	604.1	602.5	602.5	601.9	600.7	602.5	598.6	603.8	603.0	602.5	602.5
Mass Flow, kg/h	63	165	494	721	721	721	721	1,448	954	28	3,114	4,096	4,096	3,050	7,146	8,798	8,072	6,164	1,908
Std. Volume Flow, m3/h (0°C, 101.3 kPa)	83	205	371	659	698	817	817	1,088	717	22	2,606	3,323	3,323	2,553	5,876	7,162	6,755	5,158	1,597
Molar Flow, kg-mole/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CH4	3.57	-	0.03	3.61	2.75	0.10	0.10	0.10	0.06	-	-	-	-	-	-	1.45	-	-	-
CO	-	-	0.40	0.40	0.54	4.71	4.71	1.17	0.77	-	-	-	-	-	-	-	-	-	-
CO2	0.07	-	7.72	7.80	8.52	7.00	7.00	22.64	14.92	0.00	6.24	22.00	22.00	6.11	28.11	28.28	16.18	12.35	3.82
H2	-	-	0.60	0.60	3.90	10.34	10.34	1.77	1.16	-	-	-	-	-	-	-	-	-	-
H2O	-	9.13	7.76	16.90	15.32	14.19	14.19	22.75	14.99	0.01	27.53	43.82	43.82	26.96	70.78	71.36	71.36	54.49	16.87
N2	0.07	-	0.04	0.11	0.11	0.11	0.11	0.11	0.07	0.77	74.44	75.29	75.29	72.91	148.20	192.97	192.97	147.36	45.62
O2	-	-	-	-	-	-	-	-	-	0.20	8.06	7.17	7.17	7.89	15.06	26.94	20.89	15.95	4.94
Total, kg-mole/hr	3.72	9.13	16.56	29.42	31.13	36.44	36.44	48.95	31.99	0.99	116.26	148.27	148.27	113.88	262.15	319.55	301.39	230.15	71.24
Density, kg/m3	2.29	3.63	2.22	2.25	2.12	1.36	1.51	2.22	2.22	4.02	1.98	1.86	2.23	1.98	2.11	2.30	1.98	1.98	1.98
Mole Wt.	16.84	18.02	29.82	24.51	23.16	19.79	19.79	29.82	29.82	28.79	26.78	27.62	27.62	26.78	27.26	27.53	26.78	26.78	26.78
Z Factor	1.001	-	1.000	0.999	0.999	1.001	1.001	1.000	1.000	1.002	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001

FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / -29°C AMBIENT AIR

Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
Stream Description:	Natural Gas to PM #1	Steam to PM #1	Anode Recycle	Mixed Feed to Exchanger	Reformer Feed	Reformer Exhaust (Reformate)	Anode Feed	Anode Exhaust	Anode-Ehst. Fuel to Combustor	Air to Combustor	Cathode Recycle to Combustor	Combustor Exhaust	Reformer Flue Gas	Cathode Recycle (Bypass to Blower)	PM #1 Gas to Blower	Cathode Feed	Cathode Exhaust	Cathode Recycle	PM #1 Gas to Gas Turbine
Temperature, K	616	536	978	811	804	1,061	950	978	978	441	978	1,072	894	978	331	866	978	978	978
Pressure, kPa	698	872	604.1	609.2	608.5	606.3	604.8	604.4	604.1	602.5	602.5	601.9	600.7	602.5	598.6	603.8	603.0	602.5	602.5
Mass Flow, kg/h	63	144	504	710	710	710	710	1,437	933	29	3,122	4,083	4,083	3,301	7,384	8,850	8,123	6,422	1,701
Std. Volume Flow, m3/h (0°C, 101.3 kPa)	83	179	373	635	674	793	793	1,064	691	22	2,609	3,301	3,301	2,759	6,059	7,197	6,790	5,368	1,422
Molar Flow, kg-mole/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CH4	3.57	-	0.04	3.61	2.76	0.11	0.11	0.11	0.07	-	-	-	-	-	-	4.16	-	-	-
CO	-	-	0.43	0.43	0.62	5.01	5.01	1.23	0.80	-	-	-	-	-	-	-	-	-	-
CO2	0.07	-	8.04	8.11	8.77	7.04	7.04	22.93	14.89	0.00	6.98	22.74	22.74	7.38	30.13	30.28	18.17	14.37	3.81
H2	-	-	0.60	0.60	3.82	10.04	10.04	1.72	1.11	-	-	-	-	-	-	-	-	-	-
H2O	-	7.99	7.49	15.48	13.97	13.05	13.05	21.38	13.88	-	27.78	42.92	42.92	29.37	72.28	72.28	72.28	57.15	15.14
N2	0.07	-	0.04	0.11	0.11	0.11	0.11	0.11	0.07	0.78	74.92	75.77	75.77	79.21	154.98	194.95	194.95	154.13	40.82
O2	-	-	-	-	-	-	-	-	-	0.21	8.73	5.84	5.84	7.11	12.95	23.56	17.51	13.84	3.67
Total, kg-mole/hr	3.72	7.99	16.84	28.35	30.05	35.36	35.36	47.47	30.83	0.99	116.41	147.27	147.27	123.08	270.34	321.08	302.91	239.48	63.43
Density, kg/m3	2.29	3.63	2.25	2.27	2.15	1.38	1.54	2.25	2.25	4.74	1.99	1.87	2.24	1.99	2.11	2.31	1.99	1.99	1.99
Mole Wt.	16.84	18.02	30.27	25.05	23.63	20.09	20.09	30.27	30.27	28.90	26.82	27.73	27.73	26.82	27.31	27.56	26.82	26.82	26.82
Z Factor	1.001	-	1.000	0.999	0.999	1.001	1.001	1.000	1.000	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001

MATERIAL BALANCE - POWER MODULE #1

MC-POWER CORPORATION
20 MW HEPP PROGRAM

JOB NO.: 24024-000

DOCUMENT NO.

REV.

SHEET 1 OF 2

TABLE 14
Material Balance – Full Load SOR – Sea Level/49°C (0%/100% Rel. Humidity) Ambient Air (Page 2 of 2)

FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR																				
Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	
Stream Description:	Natural Gas to PM #1	Steam to PM #1	Anode Recycle	Mixed Feed to Exchanger	Reformer Feed	Reformer Exhaust (Reformate)	Anode Feed	Anode Exhaust	Anode-Exhst. Fuel to Combustor	Air to Combustor	Cathode Recycle to Combustor	Combustor Exhaust	Reformer Flue Gas	Cathode Recycle (Bypass to Blower)	PM #1 Gas to Blower	Cathode Feed	Cathode Exhaust	Cathode Recycle	PM #1 Gas to Gas Turbine	
Temperature, K	616	536	978	804	801	1,061	950	978	978	577	978	1,072	892	978	926	866	978	978	978	
Pressure, kPa	698	872	604.2	610.3	609.4	606.8	605.0	604.6	604.2	602.5	602.5	602.0	600.7	602.5	598.6	603.8	603.0	602.5	602.5	
Mass Flow, kg/h	63	179	538	780	780	780	780	1,507	969	29	3,165	4,163	4,163	2,844	7,006	8,875	8,148	6,009	2,139	
Std. Volume Flow, m ³ /h (0°C, 101.3 kPa)	83	223	409	715	757	873	873	1,145	736	22	2,631	3,367	3,367	2,364	5,731	7,180	6,773	4,995	1,778	
Molar Flow, kg-mole/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	
CH ₄	3.57	-	0.03	3.60	2.67	0.08	0.08	0.08	0.05	-	-	-	-	-	-	-	-	-	-	
CO	-	-	0.43	0.43	0.58	4.68	4.68	1.20	0.77	-	-	-	-	-	-	-	-	-	-	
CO ₂	0.07	-	8.30	8.37	9.15	7.65	7.65	23.23	14.94	0.00	5.69	21.45	21.45	5.12	26.57	26.77	14.66	10.81	3.95	
H ₂	-	-	0.68	0.68	4.25	10.53	10.53	1.90	1.22	-	-	-	-	-	-	-	-	-	-	
H ₂ O	-	9.95	8.77	18.72	17.01	15.92	15.92	24.55	15.78	-	25.31	42.41	42.41	22.74	65.15	65.15	65.15	48.04	17.10	
N ₂	0.07	-	0.04	0.12	0.12	0.12	0.12	0.12	0.07	0.78	76.66	77.51	77.51	68.86	146.37	197.32	197.32	145.52	51.80	
O ₂	-	-	-	-	-	-	-	-	-	0.21	9.74	8.85	8.85	8.75	17.59	31.11	25.06	18.48	6.58	
Total, kg-mole/hr	3.72	9.95	18.24	31.91	33.78	38.96	38.96	51.07	32.83	0.99	117.40	150.22	150.22	105.46	255.68	320.35	302.19	222.86	79.33	
Density, kg/m ³	2.29	3.63	2.19	2.23	2.12	1.38	1.53	2.19	2.19	3.62	2.00	1.87	2.24	2.00	2.13	2.32	2.00	2.00	2.00	
Mole Wt.	16.84	18.02	29.50	24.44	23.09	20.02	20.02	29.50	29.50	28.90	26.96	27.71	27.71	26.96	27.40	27.70	26.96	26.96	26.96	
Z Factor	1.001	-	1.000	0.998	0.999	1.001	1.000	1.000	1.000	1.002	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001	
FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (100% REL. HUMIDITY) AMBIENT AIR																				
Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	
Stream Description:	Natural Gas to PM #1	Steam to PM #1	Anode Recycle	Mixed Feed to Exchanger	Reformer Feed	Reformer Exhaust (Reformate)	Anode Feed	Anode Exhaust	Anode-Exhst. Fuel to Combustor	Air to Combustor	Cathode Recycle to Combustor	Combustor Exhaust	Reformer Flue Gas	Cathode Recycle (Bypass to Blower)	PM #1 Gas to Blower	Cathode Feed	Cathode Exhaust	Cathode Recycle	PM #1 Gas to Gas Turbine	
Temperature, K	616	536	978	804	801	1,061	950	978	978	571	978	1,072	892	978	926	866	978	978	978	
Pressure, kPa	698	872	603.6	609.3	609.0	606.3	604.4	604.0	603.6	602.5	602.5	601.6	600.4	602.5	598.5	603.7	603.0	602.5	602.5	
Mass Flow, kg/h	63	182	545	789	789	789	789	1,516	971	27	2,956	3,954	3,954	2,688	6,642	8,354	7,627	5,644	1,983	
Std. Volume Flow, m ³ /h (0°C, 101.3 kPa)	83	226	415	724	766	882	882	1,153	739	22	2,563	3,302	3,302	2,331	5,632	7,020	6,613	4,894	1,719	
Molar Flow, kg-mole/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	
CH ₄	3.57	-	0.03	3.60	2.66	0.08	0.08	0.08	0.05	-	-	-	-	-	-	-	-	-	-	
CO	-	-	0.43	0.43	0.58	4.67	4.67	1.20	0.77	-	-	-	-	-	-	-	-	-	-	
CO ₂	0.07	-	8.38	8.46	9.25	7.75	7.75	23.32	14.94	0.00	5.69	21.45	21.45	5.18	26.63	26.79	14.68	10.87	3.82	
H ₂	-	-	0.69	0.69	4.30	10.56	10.56	1.92	1.23	-	-	-	-	-	-	-	-	-	-	
H ₂ O	-	10.08	8.92	19.00	17.27	16.19	16.19	24.83	15.91	0.11	36.49	53.83	53.83	33.18	87.01	94.14	94.14	69.67	24.47	
N ₂	0.07	-	0.04	0.12	0.12	0.12	0.12	0.12	0.07	0.69	65.48	66.24	66.24	59.54	125.79	168.94	168.94	125.02	43.92	
O ₂	-	-	-	-	-	-	-	-	-	0.18	6.69	5.78	5.78	6.08	11.86	23.31	17.26	12.77	4.49	
Total, kg-mole/hr	3.72	10.08	18.50	32.30	34.18	39.35	39.35	51.46	32.96	0.99	114.35	147.30	147.30	103.98	251.28	313.19	295.03	218.33	76.70	
Density, kg/m ³	2.29	3.63	2.19	2.23	2.11	1.38	1.53	2.19	2.19	3.51	1.91	1.81	2.17	1.91	2.05	2.23	1.92	1.91	1.91	
Mole Wt.	16.84	18.02	29.45	24.43	23.08	20.05	20.05	29.45	29.45	27.64	25.85	26.85	26.85	25.85	26.43	26.67	25.85	25.85	25.85	
Z Factor	1.001	-	1.000	0.998	0.999	1.001	1.000	1.000	1.000	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001	
MATERIAL BALANCE - POWER MODULE #1															JOB NO.: 24024-000					
															MC-POWER CORPORATION					
															20 MW HEFPP PROGRAM					
															DOCUMENT NO.					
															REV.					
															SHEET 2 OF 2					

TABLE 15
Material Balance – Full Load SOR – -5,000' Elev./15°C/-29°C Ambient Air (Page 1 of 2)

FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 15°C AMBIENT AIR

Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
Stream Description:	Natural Gas to PM #1	Steam to PM #1	Anode Recycle	Mixed Feed to Exchanger	Reformer Feed	Reformer Exhaust (Reformate)	Anode Feed	Anode Exhaust	Anode-Exhst. Fuel to Combustor	Air to Combustor	Cathode Recycle to Combustor	Combustor Exhaust	Reformer Flue Gas	Cathode Recycle (Bypass to Blower)	PM #1 Gas to Blower	Cathode Feed	Cathode Exhaust	Cathode Recycle	PM #1 Gas to Gas Turbine
Temperature, K	616	536	978	802	800	1,061	950	978	978	543	978	1,072	893	978	928	867	978	978	978
Pressure, kPa	698	872	604.1	609.7	608.9	606.5	604.9	604.4	604.1	602.5	602.5	601.9	600.7	602.5	598.6	603.8	603.0	602.5	602.5
Mass Flow, kg/h	63	170	511	745	745	745	745	1,471	960	28	3,130	4,119	4,119	2,962	7,081	8,814	8,087	6,093	1,995
Std. Volume Flow, m ³ /h (0°C, 101.3 kPa)	83	212	386	682	721	839	839	1,111	725	22	2,616	3,341	3,341	2,476	5,817	7,166	6,759	5,092	1,667
Molar Flow, kg-mole/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CH ₄	3.57	-	0.03	3.60	2.72	0.09	0.09	0.09	0.06	-	-	-	-	-	-	1.45	-	-	-
CO	-	-	0.41	0.41	0.56	4.70	4.70	1.19	0.77	-	-	-	-	-	-	-	-	-	-
CO ₂	0.07	-	7.95	8.03	8.77	7.26	7.26	22.88	14.93	0.00	6.01	21.77	21.77	5.69	27.47	27.65	15.54	11.71	3.83
H ₂	-	-	0.63	0.63	4.04	10.41	10.41	1.82	1.19	-	-	-	-	-	-	-	-	-	-
H ₂ O	-	9.46	8.16	17.62	15.99	14.87	14.87	23.47	15.31	0.01	27.03	43.65	43.65	25.58	69.23	69.83	69.83	52.61	17.22
N ₂	0.07	-	0.04	0.11	0.11	0.11	0.11	0.11	0.07	0.77	75.03	75.87	75.87	71.00	146.87	193.82	193.82	146.02	47.80
O ₂	-	-	-	-	-	-	-	-	-	0.20	8.66	7.77	7.77	8.20	15.97	28.43	22.37	16.86	5.52
Total, kg-mole/hr	3.72	9.46	17.23	30.41	32.19	37.45	37.45	49.55	32.33	0.99	116.73	149.07	149.07	110.46	259.53	319.72	301.56	227.19	74.37
Density, kg/m ³	2.29	3.63	2.21	2.24	2.12	1.37	1.52	2.21	2.21	3.84	1.99	1.86	2.23	1.99	2.12	2.31	1.99	1.99	1.99
Mole Wt.	16.84	18.02	29.69	24.48	23.13	19.88	19.88	29.69	29.69	28.79	26.82	27.63	27.63	26.82	27.28	27.57	26.82	26.82	26.82
Z Factor	1.001	-	1.000	0.999	0.999	1.001	1.001	1.000	1.000	1.002	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001

FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / -29°C AMBIENT AIR

Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
Stream Description:	Natural Gas to PM #1	Steam to PM #1	Anode Recycle	Mixed Feed to Exchanger	Reformer Feed	Reformer Exhaust (Reformate)	Anode Feed	Anode Exhaust	Anode-Exhst. Fuel to Combustor	Air to Combustor	Cathode Recycle to Combustor	Combustor Exhaust	Reformer Flue Gas	Cathode Recycle (Bypass to Blower)	PM #1 Gas to Blower	Cathode Feed	Cathode Exhaust	Cathode Recycle	PM #1 Gas to Gas Turbine
Temperature, K	616	536	978	812	805	1,061	950	978	978	463	978	1,072	894	978	930	866	978	978	978
Pressure, kPa	698	872	604.1	609.5	608.8	606.4	604.8	604.4	604.1	602.5	602.5	601.9	600.7	602.5	598.6	603.8	603.0	602.5	602.5
Mass Flow, kg/h	63	149	520	732	732	732	732	1,458	938	29	3,134	4,101	4,101	3,238	7,339	8,856	8,129	6,372	1,757
Std. Volume Flow, m ³ /h (0°C, 101.3 kPa)	83	185	387	655	695	813	813	1,084	697	22	2,618	3,315	3,315	2,704	6,019	7,196	6,789	5,322	1,468
Molar Flow, kg-mole/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CH ₄	3.57	-	0.04	3.61	2.73	0.10	0.10	0.10	0.06	-	-	-	-	-	-	4.16	-	-	-
CO	-	-	0.44	0.44	0.64	5.00	5.00	1.24	0.80	-	-	-	-	-	-	-	-	-	-
CO ₂	0.07	-	8.26	8.34	9.02	7.29	7.29	23.16	14.89	0.00	6.80	22.56	22.56	7.02	29.58	29.74	17.63	13.82	3.81
H ₂	-	-	0.63	0.63	3.95	10.11	10.11	1.77	1.14	-	-	-	-	-	-	-	-	-	-
H ₂ O	-	8.25	7.84	16.10	14.54	13.63	13.63	21.98	14.14	-	27.47	42.87	42.87	28.37	71.24	71.24	71.24	55.84	15.40
N ₂	0.07	-	0.04	0.12	0.12	0.12	0.12	0.12	0.07	0.78	75.31	76.17	76.17	77.80	153.97	195.35	195.35	153.12	42.23
O ₂	-	-	-	-	-	-	-	-	-	0.21	7.20	6.32	6.32	7.44	13.76	24.74	18.68	14.65	4.04
Total, kg-mole/hr	3.72	8.25	17.26	29.23	30.99	36.25	36.25	48.36	31.10	0.99	116.78	147.91	147.91	120.64	268.55	321.07	302.90	237.42	65.48
Density, kg/m ³	2.29	3.63	2.24	2.26	2.15	1.39	1.55	2.24	2.24	4.52	1.99	1.87	2.24	1.99	2.11	2.31	1.99	1.99	1.99
Mole Wt.	16.84	18.02	30.15	25.03	23.61	20.18	20.18	30.15	30.15	28.90	26.84	27.73	27.73	26.84	27.33	27.58	26.84	26.84	26.84
Z Factor	1.001	-	1.000	0.999	0.999	1.001	1.001	1.000	1.000	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001

MATERIAL BALANCE - POWER MODULE #1

MC-POWER CORPORATION
20 MW HEPP PROGRAM

JOB NO.: 24024-000

DOCUMENT NO.

REV.

SHEET 1 OF 2

TABLE 15
Material Balance – Full Load SOR – -5,000' Elev./49°C (0%/100% Rel. Humidity) Ambient Air (Page 2 of 2)

FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (0% REL. HUMIDITY) AMBIENT AIR																				
Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	
Stream Description:	Natural Gas to PM #1	Steam to PM #1	Anode Recycle	Mixed Feed to Exchanger	Reformer Feed	Reformer Exhaust (Reformate)	Anode Feed	Anode Exhaust	Anode-Exhst. Fuel to Combustor	Air to Combustor	Cathode Recycle to Combustor	Combustor Exhaust	Reformer Flue Gas	Cathode Recycle (Bypass to Blower)	PM #1 Gas to Blower	Cathode Feed	Cathode Exhaust	Cathode Recycle	PM #1 Gas to Gas Turbine	
Temperature, K	616	536	978	806	802	1,061	950	978	978	605	978	1,072	892	978	925	866	978	978	978	
Pressure, kPa	698	872	604.3	610.9	609.9	607.0	605.1	604.6	604.3	602.5	602.5	602.0	600.7	602.5	598.6	603.8	603.0	602.5	602.5	
Mass Flow, kg/h	63	188	565	816	816	816	816	1,543	978	29	3,176	4,182	4,182	2,719	6,901	8,882	8,156	5,894	2,261	
Std. Volume Flow, m3/h (0°C, 101.3 kPa)	83	234	432	750	794	908	908	1,180	748	22	2,636	3,383	3,383	2,257	5,640	7,177	6,770	4,893	1,877	
Molar Flow, kg-mole/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	
CH4	3.57	-	0.03	3.60	2.62	0.07	0.07	0.07	0.04	-	-	-	-	-	-	-	-	-	-	
CO	-	-	0.44	0.44	0.60	4.65	4.65	1.21	0.77	-	-	-	-	-	-	-	-	-	-	
CO2	0.07	-	8.64	8.72	9.53	8.04	8.04	23.59	14.95	0.00	5.42	21.18	21.18	4.64	25.82	26.03	13.92	10.06	3.86	
H2	-	-	0.72	0.72	4.47	10.64	10.64	1.97	1.25	-	-	-	-	-	-	-	-	-	-	
H2O	-	10.46	9.41	19.86	18.07	17.01	17.01	25.68	16.27	-	24.73	42.33	42.33	21.17	63.50	63.50	63.50	45.89	17.61	
N2	0.07	-	0.04	0.12	0.12	0.12	0.12	0.12	0.07	0.78	77.08	77.93	77.93	65.99	143.92	197.96	197.96	143.07	54.89	
O2	-	-	-	-	-	-	-	-	-	0.21	10.39	9.50	9.50	8.89	18.39	32.74	26.68	19.28	7.40	
Total, kg-mole/hr	3.72	10.46	19.28	33.46	35.41	40.52	40.52	52.63	33.35	0.99	117.61	150.94	150.94	100.69	251.64	320.22	302.06	218.31	83.75	
Density, kg/m3	2.29	3.63	2.18	2.23	2.11	1.39	1.54	2.18	2.18	3.46	2.00	1.87	2.24	2.00	2.13	2.32	2.00	2.00	2.00	
Mole Wt.	16.84	18.02	29.32	24.40	23.05	20.14	20.14	29.32	29.32	28.90	27.00	27.70	27.70	27.00	27.42	27.74	27.00	27.00	27.00	
Z Factor	1.001	-	1.000	0.998	0.999	1.001	1.000	1.000	1.000	1.002	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001	
FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (100% REL. HUMIDITY) AMBIENT AIR																				
Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	
Stream Description:	Natural Gas to PM #1	Steam to PM #1	Anode Recycle	Mixed Feed to Exchanger	Reformer Feed	Reformer Exhaust (Reformate)	Anode Feed	Anode Exhaust	Anode-Exhst. Fuel to Combustor	Air to Combustor	Cathode Recycle to Combustor	Combustor Exhaust	Reformer Flue Gas	Cathode Recycle (Bypass to Blower)	PM #1 Gas to Blower	Cathode Feed	Cathode Exhaust	Cathode Recycle	PM #1 Gas to Gas Turbine	
Temperature, K	616	536	978	807	803	1,061	950	978	978	598	978	1,072	891	978	925	866	978	978	978	
Pressure, kPa	698	872	603.7	610.5	609.5	606.5	604.5	604.0	603.7	602.5	602.5	601.6	600.4	602.5	598.5	603.7	603.0	602.5	602.5	
Mass Flow, kg/h	63	192	575	830	830	830	830	1,557	981	27	2,963	3,971	3,971	2,574	6,545	8,352	7,625	5,537	2,089	
Std. Volume Flow, m3/h (0°C, 101.3 kPa)	83	239	441	763	808	921	921	1,193	752	22	2,567	3,318	3,318	2,230	5,548	7,013	6,606	4,796	1,809	
Molar Flow, kg-mole/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	
CH4	3.57	-	0.02	3.60	2.61	0.07	0.07	0.07	0.04	-	-	-	-	-	-	-	-	-	-	
CO	-	-	0.45	0.45	0.61	4.64	4.64	1.21	0.76	-	-	-	-	-	-	-	-	-	-	
CO2	0.07	-	8.77	8.84	9.67	8.19	8.19	23.72	14.95	0.00	5.43	21.19	21.19	4.72	25.90	26.08	13.97	10.15	3.83	
H2	-	-	0.74	0.74	4.55	10.68	10.68	2.00	1.26	-	-	-	-	-	-	-	-	-	-	
H2O	-	10.65	9.65	20.30	18.47	17.42	17.42	26.10	16.45	0.11	36.09	54.00	54.00	31.35	85.35	92.88	92.88	67.44	25.44	
N2	0.07	-	0.04	0.12	0.12	0.12	0.12	0.12	0.07	0.69	65.72	66.49	66.49	57.10	123.58	169.15	169.15	122.82	46.33	
O2	-	-	-	-	-	-	-	-	-	0.18	7.27	6.36	6.36	6.32	12.68	24.77	18.72	13.59	5.13	
Total, kg-mole/hr	3.72	10.65	19.67	34.04	36.03	41.11	41.11	53.22	33.54	0.99	114.51	148.03	148.03	99.48	247.51	312.88	294.72	213.99	80.72	
Density, kg/m3	2.29	3.63	2.17	2.22	2.11	1.39	1.54	2.17	2.17	3.35	1.92	1.81	2.17	1.92	2.06	2.23	1.92	1.92	1.92	
Mole Wt.	16.84	18.02	29.25	24.38	23.03	20.19	20.19	29.25	29.25	27.64	25.87	26.83	26.83	25.87	26.44	26.69	25.87	25.87	25.87	
Z Factor	1.001	-	1.000	0.998	0.999	1.001	1.000	1.000	1.000	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001	
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MATERIAL BALANCE - POWER MODULE #1

MC-POWER CORPORATION

20 MW HEPP PROGRAM

TABLE 16
Material Balance – Full Load EOR – Sea Level /15°C/49°C (0% Rel. Humidity) Ambient Air

FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR

Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
Stream Description:	Natural Gas to PM #1	Steam to PM #1	Anode Recycle	Mixed Feed to Exchanger	Reformer Feed	Reformer Exhaust (Reformate)	Anode Feed	Anode Exhaust	Anode-Exhst. Fuel to Combustor	Air to Combustor	Cathode Recycle to Combustor	Combustor Exhaust	Reformer Flue Gas	Cathode Recycle (Bypass to Blower)	PM #1 Gas to Blower	Cathode Feed	Cathode Exhaust	Cathode Recycle	PM #1 Gas to Gas Turbine
Temperature, K	616	536	978	805	801	1,061	950	978	978	518	978	1,072	890	978	932	866	978	978	978
Pressure, kPa	698	872	604.6	610.9	609.9	607.2	605.3	604.9	604.6	602.5	602.5	602.4	601.2	602.5	598.8	604.0	603.0	602.5	602.5
Mass Flow, kg/h	63	183	548	794	794	794	794	1,520	972	28	3,083	4,083	4,083	3,781	7,864	9,755	9,028	6,864	2,164
Std. Volume Flow, m ³ /h (0°C, 101.3 kPa)	83	227	417	728	770	885	885	1,156	739	22	2,571	3,311	3,311	3,153	6,464	7,936	7,529	5,724	1,805
Molar Flow, kg-mole/hr																			
CH ₄	3.57	-	0.04	3.62	2.66	0.12	0.12	0.12	0.07					-	-	1.45			
CO	-	-	0.41	0.41	0.58	4.61	4.61	1.14	0.73					-	-	-			
CO ₂	0.07	-	8.43	8.51	9.29	7.81	7.81	23.39	14.95	0.00	5.48	21.24	21.24	6.73	27.97	28.16	16.06	12.21	3.85
H ₂	-	-	0.66	0.66	4.30	10.46	10.46	1.82	1.17					-	-	-			
H ₂ O	-	10.14	9.01	19.15	17.42	16.36	16.36	24.39	15.98	0.01	25.58	42.88	42.88	31.38	74.26	74.92	74.92	56.96	17.96
N ₂	0.07	-	0.04	0.12	0.12	0.12	0.12	0.12	0.07	0.77	74.17	75.01	75.01	90.96	165.97	217.19	217.19	165.13	52.06
O ₂	-	-	-	-	-	-	-	-	-	0.20	9.47	8.58	8.58	11.61	20.19	33.79	27.73	21.09	6.65
Total, kg-mole/hr	3.72	10.14	18.60	32.46	34.37	39.46	39.46	51.57	32.97	0.99	114.70	147.72	147.72	140.68	288.40	354.06	335.90	255.38	80.52
Density, kg/m ³	2.29	3.63	2.19	2.24	2.12	1.38	1.54	2.19	2.19	4.02	1.99	1.87	2.24	1.99	2.11	2.31	1.99	1.99	1.99
Mole %	16.84	18.02	29.48	24.45	23.09	20.11	20.11	29.48	29.48	28.79	26.88	27.64	27.64	26.88	27.27	27.55	26.88	26.88	26.88
Z Factor	1.001		1.000	0.998	0.999	1.001	1.000	1.000	1.000	1.002	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001

FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
Stream Description:	Natural Gas to PM #1	Steam to PM #1	Anode Recycle	Mixed Feed to Exchanger	Reformer Feed	Reformer Exhaust (Reformate)	Anode Feed	Anode Exhaust	Anode-Exhst. Fuel to Combustor	Air to Combustor	Cathode Recycle to Combustor	Combustor Exhaust	Reformer Flue Gas	Cathode Recycle (Bypass to Blower)	PM #1 Gas to Blower	Cathode Feed	Cathode Exhaust	Cathode Recycle	PM #1 Gas to Gas Turbine
Temperature, K	616	536	978	807	802	1,061	950	978	978	577	978	1,072	890	978	930	866	978	978	978
Pressure, kPa	698	872	604.7	611.8	610.8	607.7	605.6	605.1	604.7	602.5	602.5	602.5	601.2	602.5	598.8	604.0	603.0	602.5	602.5
Mass Flow, kg/h	63	195	585	843	843	843	843	1,570	984	29	3,140	4,153	4,153	3,559	7,713	9,859	9,133	6,700	2,433
Std. Volume Flow, m ³ /h (0°C, 101.3 kPa)	83	243	449	775	821	933	933	1,204	755	22	2,599	3,354	3,354	2,946	6,300	7,965	7,558	5,545	2,013
Molar Flow, kg-mole/hr																			
CH ₄	3.57	-	0.04	3.61	2.60	0.10	0.10	0.10	0.06					-	-	4.16			
CO	-	-	0.43	0.43	0.61	4.58	4.58	1.16	0.73					-	-	-			
CO ₂	0.07	-	8.90	8.98	9.81	8.34	8.34	23.87	14.97	0.00	5.00	20.76	20.76	5.67	26.44	26.66	14.55	10.68	3.88
H ₂	-	-	0.72	0.72	4.60	10.62	10.62	1.93	1.21					-	-	-			
H ₂ O	-	10.83	9.90	20.73	18.88	17.86	17.86	26.54	16.64	-	23.21	41.19	41.19	26.31	67.50	67.50	67.50	49.52	17.98
N ₂	0.07	-	0.04	0.12	0.12	0.12	0.12	0.12	0.07	0.78	76.65	77.50	77.50	86.88	164.38	222.90	222.90	163.52	59.38
O ₂	-	-	-	-	-	-	-	-	-	0.21	11.09	10.20	10.20	12.57	22.77	38.30	32.25	23.66	8.59
Total, kg-mole/hr	3.72	10.83	20.04	34.59	36.62	41.62	41.62	53.72	33.69	0.99	115.95	149.65	149.65	131.43	281.08	355.36	337.20	247.37	89.83
Density, kg/m ³	2.29	3.63	2.17	2.23	2.11	1.40	1.55	2.18	2.17	3.62	2.01	1.87	2.25	2.01	2.12	2.32	2.01	2.01	2.01
Mole %	16.84	18.02	29.22	24.38	23.03	20.26	20.26	29.22	29.22	28.90	27.08	27.75	27.75	27.08	27.44	27.74	27.08	27.08	27.08
Z Factor	1.001		1.000	0.998	0.999	1.001	1.000	1.000	1.000	1.002	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001	1.001

MATERIAL BALANCE - POWER MODULE #1

MC-POWER CORPORATION
20 MW HEFPP PROGRAM

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TABLE 17
Stream Summary Information
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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR

Stream Description	Natural Gas Supply	Discharge from Natural Gas Compressor	Natural Gas Compressor GHP	Natural Gas Fuel to Combustor	Natural Gas to Process Unit	Pre-heated Natural Gas	Natural Gas Pre-heater Duty
Temperature K	288	426		426	426	616	
Pressure kPa	200	749		749	749	715	
Molar Flow kgmole/h	134.04	134.04		0.00	134.04	134.04	
Mass Flow kg/h	2257.4	2257.4		0.0	2257.4	2257.4	
Enthalpy kJ/h	1.287E+06	1.990E+06	7.032E+05	0.000E+00	1.990E+06	3.203E+06	1.212E+06
Density kg/m3	1.4125	3.5809		3.5809	3.5809	2.3479	
Mole Wt.	16.8416	16.8416		16.8416	16.8416	16.8416	
Spec. Heat kJ/kg-C	2.1213	2.499		2.499	2.499	3.1371	
Therm Cond W/m-K	0.0317	0.053		0.053	0.053	0.0862	
Viscosity cP	0.0111	0.0154		0.0154	0.0154	0.0201	
Z Factor	0.9951	0.9962		0.9962	0.9962	1.0006	
Component Flow, kg-mol/h							
CH4	128.676	128.676		0.000	128.676	128.676	
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000		0.000	0.000	0.000	
CO2	2.681	2.681		0.000	2.681	2.681	
H2	0.000	0.000		0.000	0.000	0.000	
H2O	0.000	0.000		0.000	0.000	0.000	
N2	2.681	2.681		0.000	2.681	2.681	
O2	0.000	0.000		0.000	0.000	0.000	
TOTAL, kg-mol/h	134.038	134.038		0.000	134.038	134.038	

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(Page 2 of 9)

FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR

Stream Description	Demineralized Feedwater	Pre-heated Feedwater	Demin. Pre-heater Duty	Steam to Deaerator	Boiler Feedwater	Deaerator Vent	BFW Pump Suction
Temperature K	289	358		536	378	378	378
Pressure kPa	276	241		896	119	119	148
Molar Flow kgmole/h	329.36	329.36		11.08	340.31	0.13	340.31
Mass Flow kg/h	5933.5	5933.5		199.5	6130.8	2.3	6130.8
Enthalpy kJ/h	-1.161E+07	-9.823E+06	1.783E+06	1.965E+05	-9.629E+06	1.577E+03	-9.628E+06
Density kg/m3	1014.4741	960.4306		3.7298	944.1871	0.6873	944.2008
Mole Wt.	18.0151	18.0151		18.0151	18.0151	18.0151	18.0151
Spec. Heat kJ/kg-C	4.3154	4.3654		2.0458	4.4086	1.9114	4.4085
Therm Cond W/m-K	0.5963	0.673		0.0404	0.6824	0.0248	0.6824
Viscosity cP	1.1197	0.3301		0.0186	0.2666	0.0122	0.2666
Z Factor	0.002	0.0015		0.9716	0.0007	0.9902	0.0009
Component Flow, kg-mol/h							
CH4	0.000	0.000		0.000	0.000	0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000		0.000	0.000	0.000	0.000
CO2	0.000	0.000		0.000	0.000	0.000	0.000
H2	0.000	0.000		0.000	0.000	0.000	0.000
H2O	329.361	329.361		11.077	340.312	0.126	340.312
N2	0.000	0.000		0.000	0.000	0.000	0.000
O2	0.000	0.000		0.000	0.000	0.000	0.000
TOTAL, kg-mol/h	329.361	329.361		11.077	340.312	0.126	340.312

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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR

Stream Description	BFW Pump Discharge	BFW Pump HP	BFW to Steam Drum	BFW Heater Duty	Boiler Duty	Boiler Blowdown	Boiler Steam
Temperature K	378		416			448	448
Pressure kPa	1448		1276			896	896
Molar Flow kgmole/h	340.31		340.31			0.38	339.93
Mass Flow kg/h	6130.8		6130.8			6.8	6123.9
Enthalpy kJ/h	-9.616E+06	1.206E+04	-8.557E+06	1.059E+06	1.349E+07	-8.494E+03	4.940E+06
Density kg/m3	944.6235		910.4422			879.3838	4.5488
Mole Wt.	18.0151		18.0151			18.0151	18.0151
Spec. Heat kJ/kg-C	4.4058		4.5373			4.7092	2.0286
Therm Cond W/m-K	0.6825		0.6875			0.6788	0.0308
Viscosity cP	0.266		0.1901			0.1535	0.0148
Z Factor	0.0088		0.0073			0.0049	0.952
Component Flow, kg-mol/h							
CH4	0.000		0.000			0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000		0.000			0.000	0.000
CO2	0.000		0.000			0.000	0.000
H2	0.000		0.000			0.000	0.000
H2O	340.312		340.312			0.378	339.934
N2	0.000		0.000			0.000	0.000
O2	0.000		0.000			0.000	0.000
TOTAL, kg-mol/h	340.312		340.312			0.378	339.934

TABLE 17
(Page 4 of 9)

FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR

Stream Description	Superheated Steam	Superheater Duty	Steam to Process Unit	Ambient Air	Air Compressor Discharge	Air Compressor GHP	Process Air to Combustor
Temperature K	536		536	288	518		518
Pressure kPa	896		896	101	621		621
Molar Flow kgmole/h	339.93		328.86	2101.94	2101.94		0.99
Mass Flow kg/h	6123.9		5924.4	60507.2	60507.2		28.4
Enthalpy kJ/h	6.030E+06	1.090E+06	5.834E+06	1.761E+07	3.189E+07	1.427E+07	1.499E+04
Density kg/m3	3.7298		3.7298	1.2105	4.1419		4.1419
Mole Wt.	18.0151		18.0151	28.7863	28.7863		28.7863
Spec. Heat kJ/kg-C	2.0458		2.0458	1.015	1.0499		1.0499
Therm Cond W/m-K	0.0404		0.0404	0.0251	0.0399		0.0399
Viscosity cP	0.0186		0.0186	0.0182	0.0279		0.0279
Z Factor	0.9716		0.9716	0.9992	1.0016		1.0016
Component Flow, kg-mol/h							
CH4	0.000		0.000	0.000	0.000		0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000		0.000	0.000	0.000		0.000
CO2	0.000		0.000	6.306	6.306		0.003
H2	0.000		0.000	0.000	0.000		0.000
H2O	339.934		328.858	21.019	21.019		0.010
N2	0.000		0.000	1639.516	1639.516		0.771
O2	0.000		0.000	435.102	435.102		0.205
TOTAL, kg-mol/h	339.934		328.858	2101.944	2101.944		0.988

TABLE 17
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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR

Stream Description	Process Air to Recycle Blower	Turbine Combustor Feed	Turbine Inlet	Turbine Exhaust	Turbine GHP	Flue Gas Exit N.G. Pre-heater	Flue Gas Exit Steam Superheater
Temperature K	518	978	978	671		656	643
Pressure kPa	621	603	594	105		104	103
Molar Flow kgmole/h	688.79	2564.80	2564.80	2564.80		2564.80	2564.80
Mass Flow kg/h	19827.8	68687.6	68687.6	68687.6		68687.6	68687.6
Enthalpy kJ/h	1.045E+07	8.072E+07	8.072E+07	5.350E+07	2.722E+07	5.229E+07	5.120E+07
Density kg/m3	4.1419	1.9831	1.9534	0.5034		0.5109	0.5177
Mole Wt.	28.7863	26.7809	26.7809	26.7809		26.7809	26.7809
Spec. Heat kJ/kg-C	1.0499	1.3403	1.3403	1.2421		1.2376	1.2335
Therm Cond W/m-K	0.0399	0.0694	0.0693	0.0493		0.0483	0.0474
Viscosity cP	0.0279	0.0386	0.0386	0.0292		0.0288	0.0284
Z Factor	1.0016	1.0012	1.0012	1.0001		1.0001	1.0001
Component Flow, kg-mol/h							
CH4	0.000	0.000	0.000	0.000		0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000	0.000	0.000		0.000	0.000
CO2	2.066	137.662	137.662	137.662		137.662	137.662
H2	0.000	0.000	0.000	0.000		0.000	0.000
H2O	6.888	607.225	607.225	607.225		607.225	607.225
N2	537.258	1642.171	1642.171	1642.171		1642.171	1642.171
O2	142.580	177.745	177.745	177.745		177.745	177.745
TOTAL, kg-mol/h	688.792	2564.803	2564.803	2564.803		2564.803	2564.803

TABLE 17
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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR

Stream Description	Flue Gas Exit Steam Boiler	Flue Gas Exit BFW Heater	Flue Gas Exit Demin Feedwater Heater	Anode Recycle Gas	Anode Recycle Gas	Ejector Exit Stream	Exchanger Feed Stream
Temperature K	481	468	446	978	978	837	801
Pressure kPa	103	102	101	598	598	607	607
Molar Flow kgmole/h	2564.80	2564.80	2564.80	16.56	16.56	25.69	29.42
Mass Flow kg/h	68687.6	68687.6	68687.6	493.7	493.7	658.3	721.0
Enthalpy kJ/h	3.771E+07	3.665E+07	3.487E+07	6.246E+05	6.246E+05	7.867E+05	8.757E+05
Density kg/m ³	0.6882	0.7026	0.7324	2.1927	2.1927	2.2372	2.237
Mole Wt.	26.7809	26.7809	26.7809	29.8177	29.8177	25.6212	24.51
Spec. Heat kJ/kg-C	1.1857	1.1823	1.1768	1.5567	1.5567	1.6725	1.8334
Therm Cond W/m-K	0.0362	0.0354	0.0338	0.0795	0.0795	0.0696	0.0727
Viscosity cP	0.023	0.0225	0.0217	0.0368	0.0368	0.0292	0.0279
Z Factor	0.9995	0.9995	0.9993	1.0001	1.0001	0.9982	0.9985
Component Flow, kg-mol/h							
CH ₄	0.000	0.000	0.000	0.032	0.032	0.032	3.607
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.000	0.000	0.000	0.401	0.401	0.401	0.401
CO ₂	137.662	137.662	137.662	7.722	7.722	7.722	7.797
H ₂	0.000	0.000	0.000	0.603	0.603	0.603	0.603
H ₂ O	607.225	607.225	607.225	7.760	7.760	16.895	16.895
N ₂	1642.171	1642.171	1642.171	0.039	0.039	0.039	0.113
O ₂	177.745	177.745	177.745	0.000	0.000	0.000	0.000
TOTAL, kg-mol/h	2564.803	2564.803	2564.803	16.557	16.557	25.692	29.416

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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR

Stream Description	Pre-reformer Feed Stream	Reformer Feed Stream	Reformer Exit	Reformer Process Duty	Anode Feed	MCFC Stack Energy Recovery	Anode Exhaust
Temperature K	920	799	1061		950		978
Pressure kPa	603	603	603		599		598
Molar Flow kgmole/h	29.42	31.13	36.44		36.44		48.55
Mass Flow kg/h	721.0	721.0	721.0		721.0		1447.5
Enthalpy kJ/h	1.036E+06	8.990E+05	1.340E+06	1.019E+06	1.180E+06	1.797E+06	1.831E+06
Density kg/m3	1.9349	2.1031	1.3507		1.5004		2.1927
Mole Wt.	24.51	23.159	19.7859		19.7859		29.8177
Spec. Heat kJ/kg-C	1.914	1.8619	2.0299		1.979		1.5567
Therm Cond W/m-K	0.0845	0.0848	0.1324		0.1211		0.0795
Viscosity cP	0.0319	0.0285	0.0338		0.0312		0.0368
Z Factor	0.9995	0.9991	1.0007		1.0005		1.0001
Component Flow, kg-mol/h							
CH4	3.607	2.749	0.095		0.095		0.095
C2H6							
C3H8							
n-C4H10							
CO	0.401	0.540	4.714		4.714		1.175
CO2	7.797	8.515	6.995		6.995		22.642
H2	0.603	3.895	10.337		10.337		1.768
H2O	16.895	15.319	14.185		14.185		22.754
N2	0.113	0.113	0.113		0.113		0.113
O2	0.000	0.000	0.000		0.000		0.000
TOTAL, kg-mol/h	29.416	31.131	36.439		36.439		48.547

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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR

Stream Description	Anode Exhaust to Combustor	Cathode Recycle Gas	Cathode Recycle Gas	Cathode Recycle Bypass	Cathode Recycle to Combustor	Combined Feed to Combustor	Combustor Effluent
Temperature K	978	978	978	978	978	975	1072
Pressure kPa	598	603	603	603	603	598	597
Molar Flow kgmole/h	31.99	230.14	230.15	113.88	116.26	149.24	148.27
Mass Flow kg/h	953.8	6163.5	6163.5	3049.8	3113.6	4095.9	4095.9
Enthalpy kJ/h	1.207E+06	7.244E+06	7.244E+06	3.584E+06	3.659E+06	4.881E+06	5.427E+06
Density kg/m ³	2.1927	1.9831	1.9831	1.9831	1.9831	2.0217	1.8486
Mole Wt.	29.8177	26.7809	26.7809	26.7809	26.7809	27.4451	27.6245
Spec. Heat kJ/kg-C	1.5567	1.3403	1.3403	1.3403	1.3403	1.3885	1.4168
Therm Cond W/m-K	0.0795	0.0694	0.0694	0.0694	0.0694	0.0714	0.076
Viscosity cP	0.0368	0.0386	0.0386	0.0386	0.0386	0.0377	0.0404
Z Factor	1.0001	1.0012	1.0012	1.0012	1.0012	1.001	1.001
Component Flow, kg-mol/h							
CH ₄	0.063	0.000	0.000	0.000	0.000	0.063	0.000
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.774	0.000	0.000	0.000	0.000	0.774	0.000
CO ₂	14.920	12.353	12.353	6.112	6.240	21.163	22.000
H ₂	1.165	0.000	0.000	0.000	0.000	1.165	0.000
H ₂ O	14.994	54.488	54.488	26.962	27.526	42.529	43.819
N ₂	0.074	147.355	147.356	72.915	74.440	75.285	75.285
O ₂	0.000	15.949	15.950	7.892	8.057	8.262	7.167
TOTAL, kg-mol/h	31.989	230.145	230.146	113.881	116.264	149.241	148.272

TABLE 17
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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR

Stream Description	Reformer Flue Gas	Power Module Gas to Blower	Recycle Blower Suction	Recycle Blower Discharge	Recycle Blower GHP	Cathode Exhaust Gas	Cathode Exhaust to Expander
Temperature K	893	929	864	866		978	978
Pressure kPa	596	596	596	603		603	603
Molar Flow kgmole/h	148.27	262.15	3834.63	3834.63		301.39	71.24
Mass Flow kg/h	4095.9	7145.8	105577.0	105577.0		8071.5	1908.0
Enthalpy kJ/h	4.408E+06	7.992E+06	1.064E+08	1.067E+08	3.841E+05	9.486E+06	2.242E+06
Density kg/m3	2.2173	2.1035	2.2841	2.3003		1.9831	1.9831
Mole Wt.	27.6245	27.258	27.5325	27.5325		26.7809	26.7809
Spec. Heat kJ/kg-C	1.3602	1.3523	1.2931	1.294		1.3403	1.3403
Therm Cond W/m-K	0.065	0.0669	0.0622	0.0624		0.0694	0.0694
Viscosity cP	0.0353	0.0366	0.0355	0.0356		0.0386	0.0386
Z Factor	1.0009	1.001	1.0011	1.0011		1.0012	1.0012
Component Flow, kg-mol/h							
CH4	0.000	0.000	0.000	0.000		0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000	0.000	0.000		0.000	0.000
CO2	22.000	28.112	339.413	339.413		16.177	3.824
H2	0.000	0.000	0.000	0.000		0.000	0.000
H2O	43.819	70.781	856.262	856.262		71.355	16.867
N2	75.285	148.200	2315.662	2315.662		192.972	45.616
O2	7.167	15.059	323.289	323.289		20.887	4.937
TOTAL, kg-mol/h	148.272	262.153	3834.627	3834.627		301.391	71.245

TABLE 18
Stream Summary Information
(Page 1 of 9)

FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / -20°C AMBIENT AIR

Stream Description	Natural Gas Supply	Discharge from Natural Gas Compressor	Natural Gas Compressor GHP	Natural Gas Fuel to Combustor	Natural Gas to Process Unit	Pre-heated Natural Gas	Natural Gas Pre-heater Duty
Temperature K	288	426		426	426	616	
Pressure kPa	200	749		749	749	715	
Molar Flow kgmole/h	134.04	134.04		0.00	134.04	134.04	
Mass Flow kg/h	2257.4	2257.4		0.0	2257.4	2257.4	
Enthalpy kJ/h	1.287E+06	1.990E+06	7.032E+05	0.000E+00	1.990E+06	3.203E+06	1.212E+06
Density kg/m3	1.4125	3.5809		3.5809	3.5809	2.3479	
Mole Wt.	16.8416	16.8416		16.8416	16.8416	16.8416	
Spec. Heat kJ/kg-C	2.1213	2.499		2.499	2.499	3.1371	
Therm Cond W/m-K	0.0317	0.053		0.053	0.053	0.0862	
Viscosity cP	0.0111	0.0154		0.0154	0.0154	0.0201	
Z Factor	0.9951	0.9962		0.9962	0.9962	1.0006	
Component Flow, kg-mol/h							
CH4	128.676	128.676		0.000	128.676	128.676	
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000		0.000	0.000	0.000	
CO2	2.681	2.681		0.000	2.681	2.681	
H2	0.000	0.000		0.000	0.000	0.000	
H2O	0.000	0.000		0.000	0.000	0.000	
N2	2.681	2.681		0.000	2.681	2.681	
O2	0.000	0.000		0.000	0.000	0.000	
TOTAL, kg-mol/h	134.038	134.038		0.000	134.038	134.038	

TABLE 18
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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / -20°C AMBIENT AIR

Stream Description	Demineralized Feedwater	Pre-heated Feedwater	Demin. Pre-heater Duty	Steam to Deaerator	Boiler Feedwater	Deaerator Vent	BFW Pump Suction
Temperature K	289	358		536	378	378	378
Pressure kPa	276	241		896	119	119	148
Molar Flow kgmole/h	288.08	288.08		9.70	297.66	0.13	297.66
Mass Flow kg/h	5189.9	5189.9		174.8	5362.4	2.3	5362.4
Enthalpy kJ/h	-1.015E+07	-8.592E+06	1.560E+06	1.721E+05	-8.422E+06	1.580E+03	-8.422E+06
Density kg/m3	1014.4741	960.4306		3.7298	944.1871	0.6873	944.2016
Mole Wt.	18.0151	18.0151		18.0151	18.0151	18.0151	18.0151
Spec. Heat kJ/kg-C	4.3154	4.3654		2.0458	4.4086	1.9114	4.4085
Therm Cond W/m-K	0.5963	0.673		0.0404	0.6824	0.0248	0.6824
Viscosity cP	1.1197	0.3301		0.0186	0.2666	0.0122	0.2666
Z Factor	0.002	0.0015		0.9716	0.0007	0.9902	0.0009
Component Flow, kg-mol/h							
CH4	0.000	0.000		0.000	0.000	0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000		0.000	0.000	0.000	0.000
CO2	0.000	0.000		0.000	0.000	0.000	0.000
H2	0.000	0.000		0.000	0.000	0.000	0.000
H2O	288.084	288.084		9.702	297.660	0.126	297.660
N2	0.000	0.000		0.000	0.000	0.000	0.000
O2	0.000	0.000		0.000	0.000	0.000	0.000
TOTAL, kg-mol/h	288.084	288.084		9.702	297.660	0.126	297.660

TABLE 18
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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / -20°C AMBIENT AIR

Stream Description	BFW Pump Discharge	BFW Pump HP	BFW to Steam Drum	BFW Heater Duty	Boiler Duty	Boiler Blowdown	Boiler Steam
Temperature K	378		416			448	448
Pressure kPa	1448		1276			896	896
Molar Flow kgmole/h	297.66		297.66			0.38	297.28
Mass Flow kg/h	5362.4		5362.4			6.8	5355.6
Enthalpy kJ/h	-8.411E+06	1.054E+04	-7.485E+06	9.263E+05	1.180E+07	-8.494E+03	4.320E+06
Density kg/m3	944.6243		910.4422			879.3838	4.5488
Mole Wt.	18.0151		18.0151			18.0151	18.0151
Spec. Heat kJ/kg-C	4.4058		4.5373			4.7092	2.0286
Therm Cond W/m-K	0.6825		0.6875			0.6788	0.0308
Viscosity cP	0.266		0.1901			0.1535	0.0148
Z Factor	0.0088		0.0073			0.0049	0.952
Component Flow, kg-mol/h							
CH4	0.000		0.000			0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000		0.000			0.000	0.000
CO2	0.000		0.000			0.000	0.000
H2	0.000		0.000			0.000	0.000
H2O	297.660		297.660			0.378	297.282
N2	0.000		0.000			0.000	0.000
O2	0.000		0.000			0.000	0.000
TOTAL, kg-mol/h	297.660		297.660			0.378	297.282

TABLE 18
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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / -20°C AMBIENT AIR

Stream Description	Superheated Steam	Superheater Duty	Steam to Process Unit	Ambient Air	Air Compressor Discharge	Air Compressor GHP	Process Air to Combustor
Temperature K	536		536	244	441		441
Pressure kPa	896		896	101	621		621
Molar Flow kgmole/h	297.28		287.58	1861.96	1861.96		0.99
Mass Flow kg/h	5355.6		5180.8	53801.6	53801.6		28.5
Enthalpy kJ/h	5.273E+06	9.531E+05	5.101E+06	1.320E+07	2.393E+07	1.073E+07	1.270E+04
Density kg/m3	3.7298		3.7298	1.4343	4.8815		4.8815
Mole Wt.	18.0151		18.0151	28.8951	28.8951		28.8951
Spec. Heat kJ/kg-C	2.0458		2.0458	1.0083	1.0311		1.0311
Therm Cond W/m-K	0.0404		0.0404	0.022	0.0353		0.0353
Viscosity cP	0.0186		0.0186	0.0161	0.025		0.025
Z Factor	0.9716		0.9716	0.9985	1.0011		1.0011
Component Flow, kg-mol/h							
CH4	0.000		0.000	0.000	0.000		0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000		0.000	0.000	0.000		0.000
CO2	0.000		0.000	5.642	5.642		0.003
H2	0.000		0.000	0.000	0.000		0.000
H2O	297.282		287.580	0.000	0.000		0.000
N2	0.000		0.000	1466.999	1466.999		0.778
O2	0.000		0.000	389.319	389.319		0.207
TOTAL, kg-mol/h	297.282		287.580	1861.960	1861.960		0.988

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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / -20°C AMBIENT AIR

Stream Description	Process Air to Recycle Blower	Turbine Combustor Feed	Turbine Inlet	Turbine Exhaust	Turbine GHP	Flue Gas Exit N.G. Pre-heater	Flue Gas Exit Steam Superheater
Temperature K	441	978	978	671		655	643
Pressure kPa	621	603	594	105		104	103
Molar Flow kmole/h	608.80	2283.58	2283.58	2283.58		2283.58	2283.58
Mass Flow kg/h	17591.3	61239.5	61239.5	61239.5		61239.5	61239.5
Enthalpy kJ/h	7.824E+06	7.204E+07	7.204E+07	4.779E+07	2.425E+07	4.658E+07	4.563E+07
Density kg/m ³	4.8815	1.9858	1.9561	0.5034		0.5123	0.5189
Mole Wt.	28.8951	26.8173	26.8173	26.8173		26.8173	26.8173
Spec. Heat kJ/kg-C	1.0311	1.3431	1.343	1.2444		1.2393	1.2352
Therm Cond W/m-K	0.0353	0.0693	0.0693	0.0493		0.0482	0.0473
Viscosity cP	0.025	0.0386	0.0386	0.0292		0.0287	0.0283
Z Factor	1.0011	1.0012	1.0012	1.0001		1.0001	1
Component Flow, kg-mol/h							
CH ₄	0.000	0.000	0.000	0.000		0.000	0.000
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.000	0.000	0.000	0.000		0.000	0.000
CO ₂	1.845	136.999	136.999	136.999		136.999	136.999
H ₂	0.000	0.000	0.000	0.000		0.000	0.000
H ₂ O	0.000	544.933	544.933	544.933		544.933	544.933
N ₂	479.659	1469.681	1469.681	1469.681		1469.681	1469.681
O ₂	127.294	131.968	131.968	131.968		131.968	131.968
TOTAL, kg-mol/h	608.797	2283.581	2283.581	2283.581		2283.581	2283.581

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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / -20°C AMBIENT AIR

Stream Description	Flue Gas Exit Steam Boiler	Flue Gas Exit BFW Heater	Flue Gas Exit Demin Feedwater Heater	Anode Recycle Gas	Anode Recycle Gas	Ejector Exit Stream	Exchanger Feed Stream
Temperature K	484	471	450	978	978	851	811
Pressure kPa	103	102	101	598	598	607	607
Molar Flow kgmole/h	2283.58	2283.58	2283.58	16.64	16.64	24.63	28.35
Mass Flow kg/h	61239.5	61239.5	61239.5	503.7	503.7	647.6	710.3
Enthalpy kJ/h	3.383E+07	3.290E+07	3.134E+07	6.295E+05	6.295E+05	7.712E+05	8.602E+05
Density kg/m ³	0.6852	0.6991	0.7277	2.2258	2.2258	2.2591	2.2578
Mole Wt.	26.8173	26.8173	26.8173	30.2698	30.2698	26.2949	25.0535
Spec. Heat kJ/kg-C	1.1879	1.1846	1.1791	1.5402	1.5402	1.647	1.8142
Therm Cond W/m-K	0.0364	0.0355	0.034	0.0791	0.0791	0.0705	0.0736
Viscosity cP	0.0231	0.0226	0.0218	0.037	0.037	0.0303	0.0287
Z Factor	0.9995	0.9995	0.9993	1.0002	1.0002	0.9985	0.9988
Component Flow, kg-mol/h							
CH ₄	0.000	0.000	0.000	0.037	0.037	0.037	3.612
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.000	0.000	0.000	0.431	0.431	0.431	0.431
CO ₂	136.999	136.999	136.999	8.037	8.036	8.037	8.111
H ₂	0.000	0.000	0.000	0.601	0.601	0.601	0.601
H ₂ O	544.933	544.933	544.933	7.494	7.494	15.482	15.482
N ₂	1469.681	1469.681	1469.681	0.040	0.040	0.040	0.115
O ₂	131.968	131.968	131.968	0.000	0.000	0.000	0.000
TOTAL, kg-mol/h	2283.581	2283.581	2283.581	16.640	16.640	24.628	28.352

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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / -20°C AMBIENT AIR

Stream Description	Pre-reformer Feed Stream	Reformer Feed Stream	Reformer Exit	Reformer Process Duty	Anode Feed	MCFC Stack Energy Recovery	Anode Exhaust
Temperature K	929	804	1061		950		978
Pressure kPa	603	603	603		599		598
Molar Flow kgmole/h	28.35	30.05	35.36		35.36		47.47
Mass Flow kg/h	710.3	710.3	710.3		710.3		1436.9
Enthalpy kJ/h	1.016E+06	8.777E+05	1.300E+06	1.009E+06	1.144E+06	1.797E+06	1.796E+06
Density kg/m3	1.9573	2.1308	1.3712		1.5232		2.2258
Mole Wt.	25.0535	23.6345	20.0872		20.0872		30.2698
Spec. Heat kJ/kg-C	1.8932	1.8389	1.9991		1.9492		1.5402
Therm Cond W/m-K	0.0853	0.0853	0.1318		0.1206		0.0791
Viscosity cP	0.0324	0.0289	0.0341		0.0314		0.037
Z Factor	0.9997	0.9993	1.0007		1.0006		1.0002
Component Flow, kg-mol/h							
CH4	3.612	2.761	0.107		0.107		0.107
C2H6							
C3H8							
n-C4H10							
CO	0.431	0.621	5.008		5.008		1.228
CO2	8.111	8.772	7.038		7.038		22.926
H2	0.601	3.815	10.043		10.043		1.715
H2O	15.482	13.971	13.050		13.050		21.378
N2	0.115	0.115	0.115		0.115		0.115
O2	0.000	0.000	0.000		0.000		0.000
TOTAL, kg-mol/h	28.352	30.054	35.361		35.361		47.469

TABLE 18
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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / -20°C AMBIENT AIR

Stream Description	Anode Exhaust to Combustor	Cathode Recycle Gas	Cathode Recycle Gas	Cathode Recycle Bypass	Cathode Recycle to Combustor	Combined Feed to Combustor	Combustor Effluent
Temperature K	978	978	978	978	978	975	1072
Pressure kPa	598	603	603	603	603	598	597
Molar Flow kmole/h	30.83	239.48	239.48	123.08	116.41	148.22	147.27
Mass Flow kg/h	933.2	6422.2	6422.2	3300.6	3121.7	4083.4	4083.4
Enthalpy kJ/h	1.166E+06	7.555E+06	7.555E+06	3.883E+06	3.672E+06	4.851E+06	5.397E+06
Density kg/m3	2.2258	1.9858	1.9858	1.9858	1.9858	2.0302	1.8555
Mole Wt.	30.2698	26.8173	26.8173	26.8173	26.8173	27.5492	27.728
Spec. Heat kJ/kg-C	1.5402	1.3431	1.3431	1.3431	1.3431	1.3857	1.4143
Therm Cond W/m-K	0.0791	0.0693	0.0693	0.0693	0.0693	0.0712	0.0759
Viscosity cP	0.037	0.0386	0.0386	0.0386	0.0386	0.0377	0.0404
Z Factor	1.0002	1.0012	1.0012	1.0012	1.0012	1.001	1.001
Component Flow, kg-mol/h							
CH4	0.069	0.000	0.000	0.000	0.000	0.069	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.798	0.000	0.000	0.000	0.000	0.798	0.000
CO2	14.889	14.367	14.367	7.384	6.984	21.876	22.743
H2	1.114	0.000	0.000	0.000	0.000	1.114	0.000
H2O	13.884	57.148	57.148	29.370	27.778	41.662	42.915
N2	0.074	154.127	154.127	79.210	74.917	75.770	75.770
O2	0.000	13.840	13.840	7.113	6.727	6.934	5.839
TOTAL, kg-mol/h	30.829	239.481	239.481	123.075	116.406	148.223	147.267

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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / -20°C AMBIENT AIR

Stream Description	Reformer Flue Gas	Power Module Gas to Blower	Recycle Blower Suction	Recycle Blower Discharge	Recycle Blower GHP	Cathode Exhaust Gas	Cathode Exhaust to Expander
Temperature K	894	931	864	866		978	978
Pressure kPa	596	596	596	603		603	603
Molar Flow kgmole/h	147.27	270.34	3852.91	3852.91		302.91	63.43
Mass Flow kg/h	4083.4	7384.0	106199.0	106199.0		8123.3	1701.1
Enthalpy kJ/h	4.388E+06	8.271E+06	1.071E+08	1.075E+08	3.859E+05	9.556E+06	2.001E+06
Density kg/m3	2.2232	2.1025	2.2866	2.3029		1.9858	1.9858
Mole Wt.	27.728	27.3134	27.5633	27.5633		26.8173	26.8173
Spec. Heat kJ/kg-C	1.3579	1.3519	1.2954	1.2963		1.3431	1.3431
Therm Cond W/m-K	0.065	0.067	0.0622	0.0623		0.0693	0.0693
Viscosity cP	0.0353	0.0367	0.0354	0.0355		0.0386	0.0386
Z Factor	1.0009	1.001	1.0011	1.0011		1.0012	1.0012
Component Flow, kg-mol/h							
CH4	0.000	0.000	0.000	0.000		0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000	0.000	0.000		0.000	0.000
CO2	22.743	30.127	363.367	363.367		18.173	3.806
H2	0.000	0.000	0.000	0.000		0.000	0.000
H2O	42.915	72.285	867.415	867.415		72.285	15.137
N2	75.770	154.980	2339.415	2339.415		194.951	40.824
O2	5.839	12.951	282.711	282.711		17.505	3.666
TOTAL, kg-mol/h	147.267	270.343	3852.908	3852.908		302.914	63.433

TABLE 19
Stream Summary Information (Page 1 of 9)

FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Natural Gas Supply	Discharge from Natural Gas Compressor	Natural Gas Compressor GHP	Natural Gas Fuel to Combustor	Natural Gas to Process Unit	Pre-heated Natural Gas	Natural Gas Pre-heater Duty
Temperature K	288	426		426	426	616	
Pressure kPa	200	749		749	749	715	
Molar Flow kgmole/h	134.04	134.04		0.00	134.04	134.04	
Mass Flow kg/h	2257.4	2257.4		0.0	2257.4	2257.4	
Enthalpy kJ/h	1.287E+06	1.990E+06	7.032E+05	0.000E+00	1.990E+06	3.203E+06	1.212E+06
Density kg/m3	1.4125	3.5809		3.5809	3.5809	2.3479	
Mole Wt.	16.8416	16.8416		16.8416	16.8416	16.8416	
Spec. Heat kJ/kg-C	2.1213	2.499		2.499	2.499	3.1371	
Therm Cond W/m-K	0.0317	0.053		0.053	0.053	0.0862	
Viscosity cP	0.0111	0.0154		0.0154	0.0154	0.0201	
Z Factor	0.9951	0.9962		0.9962	0.9962	1.0006	
Component Flow, kg-mol/h							
CH4	128.676	128.676		0.000	128.676	128.676	
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000		0.000	0.000	0.000	
CO2	2.681	2.681		0.000	2.681	2.681	
H2	0.000	0.000		0.000	0.000	0.000	
H2O	0.000	0.000		0.000	0.000	0.000	
N2	2.681	2.681		0.000	2.681	2.681	
O2	0.000	0.000		0.000	0.000	0.000	
TOTAL, kg-mol/h	134.038	134.038		0.000	134.038	134.038	

TABLE 19
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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Demineralized Feedwater	Pre-heated Feedwater	Demin. Pre-heater Duty	Steam to Deaerator	Boiler Feedwater	Deaerator Vent	BFW Pump Suction
Temperature K	289	358		536	378	378	378
Pressure kPa	276	241		896	119	119	148
Molar Flow kgmole/h	358.85	358.85		12.06	370.78	0.13	370.78
Mass Flow kg/h	6464.6	6464.6		217.2	6679.6	2.3	6679.6
Enthalpy kJ/h	-1.265E+07	-1.070E+07	1.943E+06	2.139E+05	-1.049E+07	1.568E+03	-1.049E+07
Density kg/m3	1014.4741	960.4306		3.7298	944.1871	0.6873	944.2007
Mole Wt.	18.0151	18.0151		18.0151	18.0151	18.0151	18.0151
Spec. Heat kJ/kg-C	4.3154	4.3654		2.0458	4.4086	1.9114	4.4085
Therm Cond W/m-K	0.5963	0.673		0.0404	0.6824	0.0248	0.6824
Viscosity cP	1.1197	0.3301		0.0186	0.2666	0.0122	0.2666
Z Factor	0.002	0.0015		0.9716	0.0007	0.9902	0.0009
Component Flow, kg-mol/h							
CH4	0.000	0.000		0.000	0.000	0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000		0.000	0.000	0.000	0.000
CO2	0.000	0.000		0.000	0.000	0.000	0.000
H2	0.000	0.000		0.000	0.000	0.000	0.000
H2O	358.845	358.845		12.057	370.777	0.125	370.777
N2	0.000	0.000		0.000	0.000	0.000	0.000
O2	0.000	0.000		0.000	0.000	0.000	0.000
TOTAL, kg-mol/h	358.845	358.845		12.057	370.777	0.125	370.777

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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	BFW Pump Discharge	BFW Pump HP	BFW to Steam Drum	BFW Heater Duty	Boiler Duty	Boiler Blowdown	Boiler Steam
Temperature K	378		416			448	448
Pressure kPa	1448		1276			896	896
Molar Flow kgmole/h	370.78		370.78			0.38	370.40
Mass Flow kg/h	6679.6		6679.6			6.8	6672.8
Enthalpy kJ/h	-1.048E+07	1.313E+04	-9.323E+06	1.154E+06	1.470E+07	-8.494E+03	5.383E+06
Density kg/m3	944.6234		910.4422			879.3838	4.5488
Mole Wt.	18.0151		18.0151			18.0151	18.0151
Spec. Heat kJ/kg-C	4.4058		4.5373			4.7092	2.0286
Therm Cond W/m-K	0.6825		0.6875			0.6788	0.0308
Viscosity cP	0.266		0.1901			0.1535	0.0148
Z Factor	0.0088		0.0073			0.0049	0.952
Component Flow, kg-mol/h							
CH4	0.000		0.000			0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000		0.000			0.000	0.000
CO2	0.000		0.000			0.000	0.000
H2	0.000		0.000			0.000	0.000
H2O	370.777		370.777			0.378	370.400
N2	0.000		0.000			0.000	0.000
O2	0.000		0.000			0.000	0.000
TOTAL, kg-mol/h	370.777		370.777			0.378	370.400

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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Superheated Steam	Superheater Duty	Steam to Process Unit	Ambient Air	Air Compressor Discharge	Air Compressor GHP	Process Air to Combustor
Temperature K	536		536	322	577		577
Pressure kPa	896		896	101	621		621
Molar Flow kgmole/h	370.40		358.34	2363.47	2363.47		0.99
Mass Flow kg/h	6672.8		6455.6	68292.7	68292.7		28.5
Enthalpy kJ/h	6.570E+06	1.188E+06	6.357E+06	2.212E+07	4.004E+07	1.792E+07	1.674E+04
Density kg/m3	3.7298		3.7298	1.0867	3.7303		3.7303
Mole Wt.	18.0151		18.0151	28.8951	28.8951		28.8951
Spec. Heat kJ/kg-C	2.0458		2.0458	1.0118	1.0557		1.0557
Therm Cond W/m-K	0.0404		0.0404	0.0275	0.0434		0.0434
Viscosity cP	0.0186		0.0186	0.0199	0.0304		0.0304
Z Factor	0.9716		0.9716	0.9997	1.0019		1.0019
Component Flow, kg-mol/h							
CH4	0.000		0.000	0.000	0.000		0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000		0.000	0.000	0.000		0.000
CO2	0.000		0.000	7.162	7.162		0.003
H2	0.000		0.000	0.000	0.000		0.000
H2O	370.400		358.342	0.000	0.000		0.000
N2	0.000		0.000	1862.127	1862.127		0.778
O2	0.000		0.000	494.180	494.180		0.207
TOTAL, kg-mol/h	370.400		358.342	2363.469	2363.469		0.988

TABLE 19
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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Process Air to Recycle Blower	Turbine Combustor Feed	Turbine Inlet	Turbine Exhaust	Turbine GHP	Flue Gas Exit N.G. Pre-heater	Flue Gas Exit Steam Superheater
Temperature K	577	978	978	669		656	643
Pressure kPa	621	603	594	105		104	103
Molar Flow kgmole/h	775.97	2855.88	2855.88	2855.88		2855.88	2855.88
Mass Flow kg/h	22421.7	77006.5	77006.5	77006.5		77006.5	77006.5
Enthalpy kJ/h	1.315E+07	8.941E+07	8.941E+07	5.914E+07	3.027E+07	5.793E+07	5.674E+07
Density kg/m3	3.7303	1.9966	1.9667	0.5081		0.5146	0.5213
Mole Wt.	28.8951	26.9642	26.9642	26.9642		26.9642	26.9642
Spec. Heat kJ/kg-C	1.0557	1.3215	1.3215	1.2255		1.2215	1.2176
Therm Cond W/m-K	0.0434	0.069	0.069	0.0491		0.0482	0.0474
Viscosity cP	0.0304	0.0392	0.0392	0.0295		0.0291	0.0287
Z Factor	1.0019	1.0012	1.0012	1.0001		1.0001	1.0001
Component Flow, kg-mol/h							
CH4	0.000	0.000	0.000	0.000		0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000	0.000	0.000		0.000	0.000
CO2	2.351	138.525	138.525	138.525		138.525	138.525
H2	0.000	0.000	0.000	0.000		0.000	0.000
H2O	0.000	615.679	615.680	615.680		615.680	615.680
N2	611.368	1864.842	1864.842	1864.842		1864.842	1864.842
O2	162.248	236.834	236.834	236.834		236.834	236.834
TOTAL, kg-mol/h	775.967	2855.880	2855.880	2855.880		2855.880	2855.880

TABLE 19
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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Flue Gas Exit Steam Boiler	Flue Gas Exit BFW Heater	Flue Gas Exit Demin Feedwater Heater	Anode Recycle Gas	Anode Recycle Gas	Ejector Exit Stream	Exchanger Feed Stream
Temperature K	484	471	449	978	978	838	804
Pressure kPa	103	102	101	598	598	607	607
Molar Flow kgmole/h	2855.88	2855.88	2855.88	18.24	18.24	28.19	31.91
Mass Flow kg/h	77006.5	77006.5	77006.5	538.0	538.0	717.3	780.0
Enthalpy kJ/h	4.204E+07	4.089E+07	3.895E+07	6.865E+05	6.865E+05	8.630E+05	9.520E+05
Density kg/m ³	0.6893	0.7033	0.7323	2.1694	2.1694	2.2199	2.2218
Mole Wt.	26.9642	26.9642	26.9642	29.5002	29.5002	25.4448	24.441
Spec. Heat kJ/kg-C	1.1717	1.1685	1.1632	1.5681	1.5681	1.6812	1.8297
Therm Cond W/m-K	0.0365	0.0357	0.0342	0.0798	0.0798	0.0699	0.0726
Viscosity cP	0.0234	0.0229	0.0221	0.0366	0.0366	0.0291	0.028
Z Factor	0.9996	0.9996	0.9994	1.0001	1.0001	0.9982	0.9984
Component Flow, kg-mol/h							
CH ₄	0.000	0.000	0.000	0.028	0.028	0.028	3.602
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.000	0.000	0.000	0.428	0.428	0.428	0.428
CO ₂	138.525	138.525	138.525	8.296	8.296	8.296	8.370
H ₂	0.000	0.000	0.000	0.677	0.677	0.677	0.677
H ₂ O	615.680	615.680	615.680	8.766	8.766	18.720	18.720
N ₂	1864.842	1864.842	1864.842	0.041	0.041	0.041	0.116
O ₂	236.834	236.834	236.834	0.000	0.000	0.000	0.000
TOTAL, kg-mol/h	2855.880	2855.880	2855.880	18.236	18.236	28.190	31.913

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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Pre-reformer Feed Stream	Reformer Feed Stream	Reformer Exit	Reformer Process Duty	Anode Feed	MCFC Stack Energy Recovery	Anode Exhaust
Temperature K	923	801	1061		950		978
Pressure kPa	603	603	603		599		598
Molar Flow kgmole/h	31.91	33.78	38.96		38.96		51.07
Mass Flow kg/h	780.0	780.0	780.0		780.0		1506.6
Enthalpy kJ/h	1.125E+06	9.759E+05	1.441E+06	1.030E+06	1.268E+06	1.797E+06	1.922E+06
Density kg/m ³	1.9233	2.0914	1.3667		1.5182		2.1694
Mole Wt.	24.441	23.0936	20.0194		20.0194		29.5002
Spec. Heat kJ/kg-C	1.9083	1.8574	2.0197		1.9688		1.5681
Therm Cond W/m-K	0.0844	0.0847	0.1298		0.1187		0.0798
Viscosity cP	0.032	0.0285	0.0339		0.0313		0.0366
Z Factor	0.9995	0.999	1.0006		1.0004		1.0001
Component Flow, kg-mol/h							
CH ₄	3.602	2.671	0.078		0.078		0.078
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.428	0.579	4.676		4.676		1.199
CO ₂	8.370	9.150	7.647		7.647		23.232
H ₂	0.677	4.250	10.527		10.527		1.897
H ₂ O	18.720	17.009	15.919		15.919		24.549
N ₂	0.116	0.116	0.116		0.116		0.116
O ₂	0.000	0.000	0.000		0.000		0.000
TOTAL, kg-mol/h	31.913	33.775	38.962		38.962		51.070

TABLE 19
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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Anode Exhaust to Combustor	Cathode Recycle Gas	Cathode Recycle Gas	Cathode Recycle Bypass	Cathode Recycle to Combustor	Combined Feed to Combustor	Combustor Effluent
Temperature K	978	978	978	978	978	975	1072
Pressure kPa	598	603	603	603	603	598	597
Molar Flow kgmole/h	32.83	222.86	222.86	105.46	117.40	151.22	150.22
Mass Flow kg/h	968.6	6009.1	6009.1	2843.7	3165.5	4162.6	4162.6
Enthalpy kJ/h	1.236E+06	6.977E+06	6.977E+06	3.302E+06	3.675E+06	4.928E+06	5.476E+06
Density kg/m ³	2.1694	1.9966	1.9966	1.9966	1.9966	2.027	1.8543
Mole Wt.	29.5002	26.9642	26.9642	26.9642	26.9642	27.5274	27.7098
Spec. Heat kJ/kg-C	1.5681	1.3215	1.3215	1.3215	1.3215	1.3769	1.4046
Therm Cond W/m-K	0.0798	0.069	0.069	0.069	0.069	0.0712	0.0758
Viscosity cP	0.0366	0.0392	0.0392	0.0392	0.0392	0.038	0.0407
Z Factor	1.0001	1.0012	1.0012	1.0012	1.0012	1.001	1.0011
Component Flow, kg-mol/h							
CH ₄	0.050	0.000	0.000	0.000	0.000	0.050	0.000
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.771	0.000	0.000	0.000	0.000	0.771	0.000
CO ₂	14.936	10.810	10.810	5.115	5.694	20.633	21.454
H ₂	1.219	0.000	0.000	0.000	0.000	1.219	0.000
H ₂ O	15.783	48.044	48.044	22.735	25.308	41.091	42.411
N ₂	0.074	145.522	145.521	68.864	76.658	77.511	77.511
O ₂	0.000	18.481	18.481	8.746	9.736	9.942	8.847
TOTAL, kg-mol/h	32.834	222.857	222.855	105.461	117.396	151.218	150.223

TABLE 19
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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Reformer Flue Gas	Power Module Gas to Blower	Recycle Blower Suction	Recycle Blower Discharge	Recycle Blower GHP	Cathode Exhaust Gas	Cathode Exhaust to Expander
Temperature K	892	926	864	866		978	978
Pressure kPa	596	596	596	603		603	603
Molar Flow kgmole/h	150.22	255.68	3844.17	3844.17		302.19	79.33
Mass Flow kg/h	4162.6	7006.3	106497.2	106497.2		8148.2	2139.1
Enthalpy kJ/h	4.446E+06	7.747E+06	1.061E+08	1.065E+08	3.850E+05	9.461E+06	2.484E+06
Density kg/m ³	2.2255	2.1197	2.2982	2.3145		1.9966	1.9966
Mole Wt.	27.7098	27.4023	27.7036	27.7036		26.9642	26.9642
Spec. Heat kJ/kg-C	1.3487	1.3383	1.2768	1.2777		1.3215	1.3215
Therm Cond W/m-K	0.0648	0.0666	0.062	0.0622		0.069	0.069
Viscosity cP	0.0355	0.0368	0.0359	0.036		0.0392	0.0392
Z Factor	1.0009	1.0011	1.0012	1.0012		1.0012	1.0012
Component Flow, kg-mol/h							
CH ₄	0.000	0.000	0.000	0.000		0.000	0.000
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.000	0.000	0.000	0.000		0.000	0.000
CO ₂	21.454	26.569	321.184	321.184		14.658	3.848
H ₂	0.000	0.000	0.000	0.000		0.000	0.000
H ₂ O	42.411	65.146	781.753	781.753		65.146	17.102
N ₂	77.511	146.375	2367.865	2367.865		197.322	51.801
O ₂	8.847	17.593	373.364	373.364		25.060	6.579
TOTAL, kg-mol/h	150.223	255.683	3844.166	3844.166		302.185	79.330

TABLE 20
Stream Summary Information
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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (100% REL. HUMIDITY) AMBIENT AIR

Stream Description	Natural Gas Supply	Discharge from Natural Gas Compressor	Natural Gas Compressor GHP	Natural Gas Fuel to Combustor	Natural Gas to Process Unit	Pre-heated Natural Gas	Natural Gas Pre-heater Duty
Temperature K	288	426		426	426	616	
Pressure kPa	200	749		749	749	715	
Molar Flow kgmole/h	134.04	134.04		0.00	134.04	134.04	
Mass Flow kg/h	2257.4	2257.4		0.0	2257.4	2257.4	
Enthalpy kJ/h	1.287E+06	1.990E+06	7.032E+05	0.000E+00	1.990E+06	3.203E+06	1.212E+06
Density kg/m3	1.4125	3.5809		3.5809	3.5809	2.3479	
Mole Wt.	16.8416	16.8416		16.8416	16.8416	16.8416	
Spec. Heat kJ/kg-C	2.1213	2.499		2.499	2.499	3.1371	
Therm Cond W/m-K	0.0317	0.053		0.053	0.053	0.0862	
Viscosity cP	0.0111	0.0154		0.0154	0.0154	0.0201	
Z Factor	0.9951	0.9962		0.9962	0.9962	1.0006	
Component Flow, kg-mol/h							
CH4	128.676	128.676		0.000	128.676	128.676	
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000		0.000	0.000	0.000	
CO2	2.681	2.681		0.000	2.681	2.681	
H2	0.000	0.000		0.000	0.000	0.000	
H2O	0.000	0.000		0.000	0.000	0.000	
N2	2.681	2.681		0.000	2.681	2.681	
O2	0.000	0.000		0.000	0.000	0.000	
TOTAL, kg-mol/h	134.038	134.038		0.000	134.038	134.038	

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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL /49°C (100%REL. HUMIDITY) AMBIENT AIR

Stream Description	Demineralized Feedwater	Pre-heated Feedwater	Demin. Pre-heater Duty	Steam to Deaerator	Boiler Feedwater	Deaerator Vent	BFW Pump Suction
Temperature K	289	358		536	378	378	378
Pressure kPa	276	241		896	119	119	148
Molar Flow kgmole/h	363.38	363.38		12.21	375.46	0.13	375.46
Mass Flow kg/h	6546.3	6546.3		219.9	6764.0	2.3	6764.0
Enthalpy kJ/h	-1.281E+07	-1.084E+07	1.967E+06	2.166E+05	-1.062E+07	1.576E+03	-1.062E+07
Density kg/m3	1014.4741	960.4306		3.7298	944.1871	0.6873	944.2007
Mole Wt.	18.0151	18.0151		18.0151	18.0151	18.0151	18.0151
Spec. Heat kJ/kg-C	4.3154	4.3654		2.0458	4.4086	1.9114	4.4085
Therm Cond W/m-K	0.5963	0.673		0.0404	0.6824	0.0248	0.6824
Viscosity cP	1.1197	0.3301		0.0186	0.2666	0.0122	0.2666
Z Factor	0.002	0.0015		0.9716	0.0007	0.9902	0.0009
Component Flow, kg-mol/h							
CH4	0.000	0.000		0.000	0.000	0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000		0.000	0.000	0.000	0.000
CO2	0.000	0.000		0.000	0.000	0.000	0.000
H2	0.000	0.000		0.000	0.000	0.000	0.000
H2O	363.381	363.381		12.209	375.464	0.126	375.464
N2	0.000	0.000		0.000	0.000	0.000	0.000
O2	0.000	0.000		0.000	0.000	0.000	0.000
TOTAL, kg-mol/h	363.381	363.381		12.209	375.464	0.126	375.464

TABLE 20
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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (100% REL. HUMIDITY) AMBIENT AIR

Stream Description	BFW Pump Discharge	BFW Pump HP	BFW to Steam Drum	BFW Heater Duty	Boiler Duty	Boiler Blowdown	Boiler Steam
Temperature K	378		416			448	448
Pressure kPa	1448		1276			896	896
Molar Flow kgmole/h	375.46		375.46			0.38	375.09
Mass Flow kg/h	6764.0		6764.0			6.8	6757.2
Enthalpy kJ/h	-1.061E+07	1.330E+04	-9.441E+06	1.168E+06	1.488E+07	-8.494E+03	5.451E+06
Density kg/m3	944.6234		910.4422			879.3838	4.5488
Mole Wt.	18.0151		18.0151			18.0151	18.0151
Spec. Heat kJ/kg-C	4.4058		4.5373			4.7092	2.0286
Therm Cond W/m-K	0.6825		0.6875			0.6788	0.0308
Viscosity cP	0.266		0.1901			0.1535	0.0148
Z Factor	0.0088		0.0073			0.0049	0.952
Component Flow, kg-mol/h							
CH4	0.000		0.000			0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000		0.000			0.000	0.000
CO2	0.000		0.000			0.000	0.000
H2	0.000		0.000			0.000	0.000
H2O	375.464		375.464			0.378	375.086
N2	0.000		0.000			0.000	0.000
O2	0.000		0.000			0.000	0.000
TOTAL, kg-mol/h	375.464		375.464			0.378	375.086

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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (100% REL. HUMIDITY) AMBIENT AIR

Stream Description	Superheated Steam	Superheater Duty	Steam to Process Unit	Ambient Air	Air Compressor Discharge	Air Compressor GHP	Process Air to Combustor
Temperature K	536		536	322	571		571
Pressure kPa	896		896	101	621		621
Molar Flow kgmole/h	375.09		362.88	2264.14	2264.14		0.99
Mass Flow kg/h	6757.2		6537.3	62585.1	62585.1		27.3
Enthalpy kJ/h	6.654E+06	1.203E+06	6.437E+06	2.152E+07	3.859E+07	1.707E+07	1.684E+04
Density kg/m3	3.7298		3.7298	1.0407	3.6101		3.6101
Mole Wt.	18.0151		18.0151	27.6418	27.6418		27.6418
Spec. Heat kJ/kg-C	2.0458		2.0458	1.0777	1.1275		1.1275
Therm Cond W/m-K	0.0404		0.0404	0.0267	0.0431		0.0431
Viscosity cP	0.0186		0.0186	0.0182	0.028		0.028
Z Factor	0.9716		0.9716	0.9986	1.0009		1.0009
Component Flow, kg-mol/h							
CH4	0.000		0.000	0.000	0.000		0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000		0.000	0.000	0.000		0.000
CO2	0.000		0.000	6.071	6.071		0.003
H2	0.000		0.000	0.000	0.000		0.000
H2O	375.086		362.878	260.811	260.811		0.114
N2	0.000		0.000	1578.383	1578.383		0.689
O2	0.000		0.000	418.879	418.879		0.183
TOTAL, kg-mol/h	375.086		362.878	2264.143	2264.143		0.988

TABLE 20
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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (100% REL. HUMIDITY) AMBIENT AIR

Stream Description	Process Air to Recycle Blower	Turbine Combustor Feed	Turbine Inlet	Turbine Exhaust	Turbine GHP	Flue Gas Exit N.G. Pre-heater	Flue Gas Exit Steam Superheater
Temperature K	571	978	978	674		661	648
Pressure kPa	621	603	594	105		104	103
Molar Flow kgmole/h	742.86	2761.06	2761.06	2761.06		2761.06	2761.06
Mass Flow kg/h	20534.0	71379.3	71379.3	71379.3		71379.3	71379.3
Enthalpy kJ/h	1.266E+07	8.801E+07	8.801E+07	5.862E+07	2.939E+07	5.741E+07	5.621E+07
Density kg/m ³	3.6101	1.9148	1.8861	0.4832		0.4895	0.496
Mole Wt.	27.6418	25.8522	25.8522	25.8522		25.8522	25.8522
Spec. Heat kJ/kg-C	1.1275	1.4117	1.4117	1.3064		1.3019	1.2974
Therm Cond W/m-K	0.0431	0.0708	0.0708	0.05		0.049	0.0481
Viscosity cP	0.028	0.0371	0.0371	0.0284		0.028	0.0276
Z Factor	1.0009	1.001	1.0009	1		1	0.9999
Component Flow, kg-mol/h							
CH ₄	0.000	0.000	0.000	0.000		0.000	0.000
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.000	0.000	0.000	0.000		0.000	0.000
CO ₂	1.992	137.429	137.429	137.429		137.429	137.429
H ₂	0.000	0.000	0.000	0.000		0.000	0.000
H ₂ O	85.571	881.045	881.045	881.045		881.045	881.045
N ₂	517.863	1581.061	1581.061	1581.061		1581.061	1581.061
O ₂	137.433	161.523	161.523	161.523		161.523	161.523
TOTAL, kg-mol/h	742.858	2761.058	2761.058	2761.058		2761.058	2761.058

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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (100% REL. HUMIDITY) AMBIENT AIR

Stream Description	Flue Gas Exit Steam Boiler	Flue Gas Exit BFW Heater	Flue Gas Exit Demin Feedwater Heater	Anode Recycle Gas	Anode Recycle Gas	Ejector Exit Stream	Exchanger Feed Stream
Temperature K	484	471	449	978	978	838	804
Pressure kPa	103	102	101	598	598	607	607
Molar Flow kgmole/h	2761.06	2761.06	2761.06	18.50	18.50	28.58	32.30
Mass Flow kg/h	71379.3	71379.3	71379.3	544.8	544.8	726.4	789.1
Enthalpy kJ/h	4.132E+07	4.016E+07	3.819E+07	6.960E+05	6.960E+05	8.748E+05	9.638E+05
Density kg/m ³	0.6601	0.674	0.7028	2.166	2.166	2.2174	2.2195
Mole Wt.	25.8522	25.8522	25.8522	29.4533	29.4533	25.4186	24.4299
Spec. Heat kJ/kg-C	1.2446	1.2409	1.2348	1.5699	1.5699	1.6825	1.8293
Therm Cond W/m-K	0.0363	0.0353	0.0337	0.0798	0.0798	0.0699	0.0726
Viscosity cP	0.022	0.0214	0.0204	0.0366	0.0366	0.0291	0.028
Z Factor	0.9993	0.9992	0.999	1.0001	1.0001	0.9982	0.9984
Component Flow, kg-mol/h							
CH ₄	0.000	0.000	0.000	0.027	0.027	0.027	3.601
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.000	0.000	0.000	0.432	0.432	0.432	0.432
CO ₂	137.429	137.429	137.429	8.382	8.382	8.382	8.457
H ₂	0.000	0.000	0.000	0.688	0.688	0.688	0.688
H ₂ O	881.045	881.045	881.045	8.925	8.925	19.005	19.005
N ₂	1581.061	1581.061	1581.061	0.042	0.042	0.042	0.116
O ₂	161.523	161.523	161.523	0.000	0.000	0.000	0.000
TOTAL, kg-mol/h	2761.058	2761.058	2761.058	18.496	18.496	28.576	32.299

TABLE 20
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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (100% REL. HUMIDITY) AMBIENT AIR

Stream Description	Pre-reformer Feed Stream	Reformer Feed Stream	Reformer Exit	Reformer Process Duty	Anode Feed	MCFC Stack Energy Recovery	Anode Exhaust
Temperature K	923	801	1061		950		978
Pressure kPa	603	603	603		599		598
Molar Flow kgmole/h	32.30	34.18	39.35		39.35		51.46
Mass Flow kg/h	789.1	789.1	789.1		789.1		1515.6
Enthalpy kJ/h	1.139E+06	9.878E+05	1.456E+06	1.032E+06	1.281E+06	1.797E+06	1.936E+06
Density kg/m ³	1.9216	2.0897	1.369		1.5207		2.166
Mole Wt.	24.4299	23.083	20.0519		20.0519		29.4533
Spec. Heat kJ/kg-C	1.9077	1.8569	2.0184		1.9675		1.5699
Therm Cond W/m-K	0.0844	0.0847	0.1294		0.1184		0.0798
Viscosity cP	0.032	0.0286	0.0339		0.0313		0.0366
Z Factor	0.9995	0.999	1.0006		1.0004		1.0001
Component Flow, kg-mol/h							
CH ₄	3.601	2.659	0.075		0.075		0.075
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.432	0.585	4.669		4.669		1.202
CO ₂	8.457	9.246	7.745		7.745		23.321
H ₂	0.688	4.305	10.555		10.555		1.915
H ₂ O	19.005	17.273	16.190		16.190		24.830
N ₂	0.116	0.116	0.116		0.116		0.116
O ₂	0.000	0.000	0.000		0.000		0.000
TOTAL, kg-mol/h	32.299	34.184	39.352		39.352		51.459

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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (100% REL. HUMIDITY) AMBIENT AIR

Stream Description	Anode Exhaust to Combustor	Cathode Recycle Gas	Cathode Recycle Gas	Cathode Recycle Bypass	Cathode Recycle to Combustor	Combined Feed to Combustor	Combustor Effluent
Temperature K	978	978	978	978	978	975	1072
Pressure kPa	598	603	603	603	603	598	597
Molar Flow kgmole/h	32.96	218.33	218.33	103.98	114.35	148.30	147.30
Mass Flow kg/h	970.9	5644.3	5644.3	2688.2	2956.1	3954.3	3954.3
Enthalpy kJ/h	1.240E+06	6.959E+06	6.959E+06	3.314E+06	3.645E+06	4.902E+06	5.450E+06
Density kg/m3	2.166	1.9148	1.9148	1.9148	1.9148	1.9641	1.7967
Mole Wt.	29.4533	25.8522	25.8522	25.8522	25.8522	26.6645	26.8452
Spec. Heat kJ/kg-C	1.5699	1.4117	1.4117	1.4117	1.4117	1.4485	1.4785
Therm Cond W/m-K	0.0798	0.0708	0.0708	0.0708	0.0708	0.0727	0.0774
Viscosity cP	0.0366	0.0371	0.0371	0.0371	0.0371	0.0367	0.0392
Z Factor	1.0001	1.001	1.001	1.001	1.001	1.0008	1.0009
Component Flow, kg-mol/h							
CH4	0.048	0.000	0.000	0.000	0.000	0.048	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.770	0.000	0.000	0.000	0.000	0.770	0.000
CO2	14.939	10.867	10.867	5.176	5.692	20.633	21.451
H2	1.227	0.000	0.000	0.000	0.000	1.227	0.000
H2O	15.905	69.669	69.669	33.181	36.488	52.507	53.830
N2	0.074	125.022	125.022	59.544	65.479	66.242	66.242
O2	0.000	12.772	12.772	6.083	6.689	6.872	5.777
TOTAL, kg-mol/h	32.963	218.330	218.330	103.983	114.347	148.299	147.300

TABLE 20
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FULL-LOAD, START-OF-RUN OPERATION - SEA LEVEL / 49°C (100% REL. HUMIDITY) AMBIENT AIR

Stream Description	Reformer Flue Gas	Power Module Gas to Blower	Recycle Blower Suction	Recycle Blower Discharge	Recycle Blower GHP	Cathode Exhaust Gas	Cathode Exhaust to Expander
Temperature K	892	926	864	866		978	978
Pressure kPa	596	596	596	603		603	603
Molar Flow kgmole/h	147.30	251.28	3758.26	3758.26		295.03	76.70
Mass Flow kg/h	3954.3	6642.5	100243.9	100243.9		7627.1	1982.8
Enthalpy kJ/h	4.418E+06	7.733E+06	1.055E+08	1.058E+08	3.763E+05	9.404E+06	2.445E+06
Density kg/m3	2.1577	2.0455	2.2133	2.2292		1.9148	1.9148
Mole Wt.	26.8452	26.4343	26.673	26.673		25.8522	25.8522
Spec. Heat kJ/kg-C	1.4178	1.416	1.355	1.356		1.4117	1.4117
Therm Cond W/m-K	0.066	0.068	0.0633	0.0635		0.0708	0.0708
Viscosity cP	0.0344	0.0354	0.0343	0.0344		0.0371	0.0371
Z Factor	1.0006	1.0008	1.0008	1.0008		1.001	1.001
Component Flow, kg-mol/h							
CH4	0.000	0.000	0.000	0.000		0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000	0.000	0.000		0.000	0.000
CO2	21.451	26.626	321.509	321.509		14.685	3.817
H2	0.000	0.000	0.000	0.000		0.000	0.000
H2O	53.830	87.011	1129.705	1129.705		94.142	24.473
N2	66.242	125.785	2027.287	2027.287		168.941	43.918
O2	5.777	11.860	279.757	279.757		17.259	4.487
TOTAL, kg-mol/h	147.300	251.283	3758.258	3758.258		295.027	76.696

TABLE 21
Stream Summary Information
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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 15°C AMBIENT AIR

Stream Description	Natural Gas Supply	Discharge from Natural Gas Compressor	Natural Gas Compressor GHP	Natural Gas Fuel to Combustor	Natural Gas to Process Unit	Pre-heated Natural Gas	Natural Gas Pre-heater Duty
Temperature K	288	426		426	426	616	
Pressure kPa	200	749		749	749	715	
Molar Flow kgmole/h	134.04	134.04		0.00	134.04	134.04	
Mass Flow kg/h	2257.4	2257.4		0.0	2257.4	2257.4	
Enthalpy kJ/h	1.287E+06	1.990E+06	7.032E+05	0.000E+00	1.990E+06	3.203E+06	1.212E+06
Density kg/m ³	1.4125	3.5809		3.5809	3.5809	2.3479	
Mole Wt.	16.8416	16.8416		16.8416	16.8416	16.8416	
Spec. Heat kJ/kg-C	2.1213	2.499		2.499	2.499	3.1371	
Therm Cond W/m-K	0.0317	0.053		0.053	0.053	0.0862	
Viscosity cP	0.0111	0.0154		0.0154	0.0154	0.0201	
Z Factor	0.9951	0.9962		0.9962	0.9962	1.0006	
Component Flow, kg-mol/h							
CH ₄	128.676	128.676		0.000	128.676	128.676	
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.000	0.000		0.000	0.000	0.000	
CO ₂	2.681	2.681		0.000	2.681	2.681	
H ₂	0.000	0.000		0.000	0.000	0.000	
H ₂ O	0.000	0.000		0.000	0.000	0.000	
N ₂	2.681	2.681		0.000	2.681	2.681	
O ₂	0.000	0.000		0.000	0.000	0.000	
TOTAL, kg-mol/h	134.038	134.038		0.000	134.038	134.038	

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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 15°C AMBIENT AIR

Stream Description	Demineralized Feedwater	Pre-heated Feedwater	Demin. Pre-heater Duty	Steam to Deaerator	Boiler Feedwater	Deaerator Vent	BFW Pump Suction
Temperature K	289	358		536	378	378	378
Pressure kPa	276	241		896	119	119	148
Molar Flow kgmole/h	341.15	341.15		11.47	352.50	0.13	352.50
Mass Flow kg/h	6145.9	6145.9		206.6	6350.3	2.3	6350.3
Enthalpy kJ/h	-1.202E+07	-1.018E+07	1.847E+06	2.034E+05	-9.973E+06	1.574E+03	-9.973E+06
Density kg/m3	1014.4741	960.4306		3.7298	944.1871	0.6873	944.2007
Mole Wt.	18.0151	18.0151		18.0151	18.0151	18.0151	18.0151
Spec. Heat kJ/kg-C	4.3154	4.3654		2.0458	4.4086	1.9114	4.4085
Therm Cond W/m-K	0.5963	0.673		0.0404	0.6824	0.0248	0.6824
Viscosity cP	1.1197	0.3301		0.0186	0.2666	0.0122	0.2666
Z Factor	0.002	0.0015		0.9716	0.0007	0.9902	0.0009
Component Flow, kg-mol/h							
CH4	0.000	0.000		0.000	0.000	0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000		0.000	0.000	0.000	0.000
CO2	0.000	0.000		0.000	0.000	0.000	0.000
H2	0.000	0.000		0.000	0.000	0.000	0.000
H2O	341.155	341.155		11.469	352.498	0.126	352.498
N2	0.000	0.000		0.000	0.000	0.000	0.000
O2	0.000	0.000		0.000	0.000	0.000	0.000
TOTAL, kg-mol/h	341.155	341.155		11.469	352.498	0.126	352.498

TABLE 21
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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 15°C AMBIENT AIR

Stream Description	BFW Pump Discharge	BFW Pump HP	BFW to Steam Drum	BFW Heater Duty	Boiler Duty	Boiler Blowdown	Boiler Steam
Temperature K	378		416			448	448
Pressure kPa	1448		1276			896	896
Molar Flow kgmole/h	352.50		352.50			0.38	352.12
Mass Flow kg/h	6350.3		6350.3			6.8	6343.5
Enthalpy kJ/h	-9.961E+06	1.249E+04	-8.864E+06	1.097E+06	1.397E+07	-8.494E+03	5.117E+06
Density kg/m3	944.6234		910.4422			879.3838	4.5488
Mole Wt.	18.0151		18.0151			18.0151	18.0151
Spec. Heat kJ/kg-C	4.4058		4.5373			4.7092	2.0286
Therm Cond W/m-K	0.6825		0.6875			0.6788	0.0308
Viscosity cP	0.266		0.1901			0.1535	0.0148
Z Factor	0.0088		0.0073			0.0049	0.952
Component Flow, kg-mol/h							
CH4	0.000		0.000			0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000		0.000			0.000	0.000
CO2	0.000		0.000			0.000	0.000
H2	0.000		0.000			0.000	0.000
H2O	352.498		352.498			0.378	352.121
N2	0.000		0.000			0.000	0.000
O2	0.000		0.000			0.000	0.000
TOTAL, kg-mol/h	352.498		352.498			0.378	352.121

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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 15°C AMBIENT AIR

Stream Description	Superheated Steam	Superheater Duty	Steam to Process Unit	Ambient Air	Air Compressor Discharge	Air Compressor GHP	Process Air to Combustor
Temperature K	536		536	288	543		543
Pressure kPa	896		896	86	621		621
Molar Flow kgmole/h	352.12		340.65	2202.73	2202.73		0.99
Mass Flow kg/h	6343.5		6136.9	63408.4	63408.4		28.4
Enthalpy kJ/h	6.246E+06	1.129E+06	6.043E+06	1.846E+07	3.509E+07	1.662E+07	1.574E+04
Density kg/m3	3.7298		3.7298	1.0363	3.9504		3.9504
Mole Wt.	18.0151		18.0151	28.7863	28.7863		28.7863
Spec. Heat kJ/kg-C	2.0458		2.0458	1.0147	1.0548		1.0548
Therm Cond W/m-K	0.0404		0.0404	0.0251	0.0414		0.0414
Viscosity cP	0.0186		0.0186	0.0182	0.0289		0.0289
Z Factor	0.9716		0.9716	0.9993	1.0017		1.0017
Component Flow, kg-mol/h							
CH4	0.000		0.000	0.000	0.000		0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000		0.000	0.000	0.000		0.000
CO2	0.000		0.000	6.608	6.608		0.003
H2	0.000		0.000	0.000	0.000		0.000
H2O	352.121		340.651	22.027	22.027		0.010
N2	0.000		0.000	1718.128	1718.128		0.771
O2	0.000		0.000	455.965	455.965		0.205
TOTAL, kg-mol/h	352.121		340.651	2202.728	2202.728		0.988

TABLE 21
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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 15°C AMBIENT AIR

Stream Description	Process Air to Recycle Blower	Turbine Combustor Feed	Turbine Inlet	Turbine Exhaust	Turbine GHP	Flue Gas Exit N.G. Pre- heater	Flue Gas Exit Steam Superheater
Temperature K	543	978	978	670		656	644
Pressure kPa	621	603	594	105		104	103
Molar Flow kgmole/h	722.39	2677.41	2677.41	2677.41		2677.41	2677.41
Mass Flow kg/h	20794.9	71802.7	71802.7	71802.7		71802.7	71802.7
Enthalpy kJ/h	1.151E+07	8.413E+07	8.413E+07	5.573E+07	2.840E+07	5.452E+07	5.339E+07
Density kg/m ³	3.9504	1.9859	1.9561	0.5044		0.5116	0.5183
Mole Wt.	28.7863	26.818	26.818	26.818		26.818	26.818
Spec. Heat kJ/kg-C	1.0548	1.3355	1.3355	1.2379		1.2336	1.2296
Therm Cond W/m-K	0.0414	0.0693	0.0693	0.0492		0.0483	0.0474
Viscosity cP	0.0289	0.0388	0.0388	0.0293		0.0289	0.0285
Z Factor	1.0017	1.0012	1.0012	1.0001		1.0001	1.0001
Component Flow, kg-mol/h							
CH ₄	0.000	0.000	0.000	0.000		0.000	0.000
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.000	0.000	0.000	0.000		0.000	0.000
CO ₂	2.167	137.963	137.963	137.963		137.963	137.963
H ₂	0.000	0.000	0.000	0.000		0.000	0.000
H ₂ O	7.224	619.980	619.980	619.980		619.980	619.980
N ₂	563.462	1720.832	1720.832	1720.832		1720.832	1720.832
O ₂	149.534	198.635	198.635	198.635		198.635	198.635
TOTAL, kg-mol/h	722.387	2677.409	2677.409	2677.409		2677.409	2677.409

TABLE 21
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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 15°C AMBIENT AIR

Stream Description	Flue Gas Exit Steam Boiler	Flue Gas Exit BFW Heater	Flue Gas Exit Demin Feedwater Heater	Anode Recycle Gas	Anode Recycle Gas	Ejector Exit Stream	Exchanger Feed Stream
Temperature K	482	469	447	978	978	838	802
Pressure kPa	103	102	101	598	598	607	607
Molar Flow kgmole/h	2677.41	2677.41	2677.41	17.23	17.23	26.69	30.41
Mass Flow kg/h	71802.7	71802.7	71802.7	511.4	511.4	681.9	744.6
Enthalpy kJ/h	3.942E+07	3.832E+07	3.647E+07	6.493E+05	6.493E+05	8.171E+05	9.061E+05
Density kg/m ³	0.6875	0.7018	0.7312	2.1832	2.1832	2.2301	2.2309
Mole Wt.	26.818	26.818	26.818	29.6879	29.6881	25.5493	24.4832
Spec. Heat kJ/kg-C	1.1823	1.179	1.1735	1.5613	1.5613	1.676	1.8317
Therm Cond W/m-K	0.0364	0.0355	0.034	0.0796	0.0796	0.0697	0.0726
Viscosity cP	0.0231	0.0227	0.0218	0.0367	0.0367	0.0292	0.0279
Z Factor	0.9996	0.9995	0.9993	1.0001	1.0001	0.9982	0.9985
Component Flow, kg-mol/h							
CH ₄	0.000	0.000	0.000	0.030	0.030	0.030	3.605
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.000	0.000	0.000	0.412	0.412	0.412	0.412
CO ₂	137.963	137.963	137.963	7.953	7.953	7.953	8.028
H ₂	0.000	0.000	0.000	0.633	0.633	0.633	0.633
H ₂ O	619.980	619.980	619.980	8.158	8.157	17.620	17.620
N ₂	1720.832	1720.832	1720.832	0.040	0.040	0.040	0.114
O ₂	198.635	198.635	198.635	0.000	0.000	0.000	0.000
TOTAL, kg-mol/h	2677.409	2677.409	2677.409	17.226	17.226	26.689	30.412

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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 15°C AMBIENT AIR

Stream Description	Pre-reformer Feed Stream	Reformer Feed Stream	Reformer Exit	Reformer Process Duty	Anode Feed	MCFC Stack Energy Recovery	Anode Exhaust
Temperature K	921	800	1061		950		978
Pressure kPa	603	603	603		599		598
Molar Flow kgmole/h	30.41	32.19	37.45		37.45		49.55
Mass Flow kg/h	744.6	744.6	744.6		744.6		1471.2
Enthalpy kJ/h	1.072E+06	9.297E+05	1.380E+06	1.023E+06	1.215E+06	1.797E+06	1.868E+06
Density kg/m3	1.9302	2.0984	1.3575		1.5079		2.1832
Mole Wt.	24.4832	23.1336	19.884		19.8842		29.6881
Spec. Heat kJ/kg-C	1.9115	1.8599	2.0255		1.9746		1.5613
Therm Cond W/m-K	0.0845	0.0847	0.1313		0.1201		0.0796
Viscosity cP	0.0319	0.0285	0.0338		0.0312		0.0367
Z Factor	0.9995	0.9991	1.0007		1.0005		1.0001
Component Flow, kg-mol/h							
CH4	3.605	2.718	0.088		0.088		0.088
C2H6							
C3H8							
n-C4H10							
CO	0.412	0.556	4.699		4.700		1.186
CO2	8.028	8.771	7.258		7.258		22.880
H2	0.633	4.038	10.414		10.414		1.821
H2O	17.620	15.990	14.873		14.873		23.466
N2	0.114	0.114	0.114		0.114		0.114
O2	0.000	0.000	0.000		0.000		0.000
TOTAL, kg-mol/h	30.412	32.186	37.446		37.446		49.554

TABLE 21
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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 15°C AMBIENT AIR

Stream Description	Anode Exhaust to Combustor	Cathode Recycle Gas	Cathode Recycle Gas	Cathode Recycle Bypass	Cathode Recycle to Combustor	Combined Feed to Combustor	Combustor Effluent
Temperature K	978	978	978	978	978	975	1072
Pressure kPa	598	603	603	603	603	598	597
Molar Flow kgmole/h	32.33	227.19	227.19	110.46	116.73	150.05	149.07
Mass Flow kg/h	959.8	6092.8	6092.8	2962.3	3130.5	4118.7	4118.7
Enthalpy kJ/h	1.218E+06	7.139E+06	7.139E+06	3.471E+06	3.668E+06	4.902E+06	5.449E+06
Density kg/m3	2.1832	1.9859	1.9859	1.9859	1.9859	2.0217	1.849
Mole Wt.	29.6881	26.818	26.818	26.818	26.818	27.4493	27.6299
Spec. Heat kJ/kg-C	1.5613	1.3355	1.3355	1.3355	1.3355	1.386	1.4141
Therm Cond W/m-K	0.0796	0.0693	0.0693	0.0693	0.0693	0.0713	0.076
Viscosity cP	0.0367	0.0388	0.0388	0.0388	0.0388	0.0377	0.0405
Z Factor	1.0001	1.0012	1.0012	1.0012	1.0012	1.001	1.001
Component Flow, kg-mol/h							
CH4	0.057	0.000	0.000	0.000	0.000	0.057	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.773	0.000	0.000	0.000	0.000	0.773	0.000
CO2	14.926	11.707	11.707	5.692	6.015	20.944	21.775
H2	1.188	0.000	0.000	0.000	0.000	1.188	0.000
H2O	15.309	52.607	52.608	25.578	27.030	42.348	43.650
N2	0.074	146.021	146.021	70.995	75.026	75.871	75.871
O2	0.000	16.856	16.855	8.195	8.660	8.865	7.770
TOTAL, kg-mol/h	32.328	227.191	227.191	110.460	116.731	150.047	149.066

TABLE 21
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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 15°C AMBIENT AIR

Stream Description	Reformer Flue Gas	Power Module Gas to Blower	Recycle Blower Suction	Recycle Blower Discharge	Recycle Blower GHP	Cathode Exhaust Gas	Cathode Exhaust to Expander
Temperature K	893	928	864	867		978	978
Pressure kPa	596	596	596	603		603	603
Molar Flow kgmole/h	149.07	259.53	3836.70	3836.70		301.56	74.37
Mass Flow kg/h	4118.7	7081.0	105766.8	105766.8		8087.3	1994.5
Enthalpy kJ/h	4.425E+06	7.897E+06	1.063E+08	1.066E+08	3.842E+05	9.476E+06	2.337E+06
Density kg/m3	2.2181	2.1075	2.2869	2.3031		1.9859	1.9859
Mole Wt.	27.6299	27.2843	27.5671	27.5671		26.818	26.818
Spec. Heat kJ/kg-C	1.3577	1.3491	1.2889	1.2898		1.3355	1.3355
Therm Cond W/m-K	0.065	0.0668	0.0622	0.0623		0.0693	0.0693
Viscosity cP	0.0353	0.0366	0.0356	0.0357		0.0388	0.0388
Z Factor	1.0009	1.001	1.0011	1.0012		1.0012	1.0012
Component Flow, kg-mol/h							
CH4	0.000	0.000	0.000	0.000		0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000	0.000	0.000		0.000	0.000
CO2	21.775	27.466	331.762	331.762		15.539	3.832
H2	0.000	0.000	0.000	0.000		0.000	0.000
H2O	43.650	69.228	837.958	837.958		69.830	17.222
N2	75.871	146.866	2325.859	2325.859		193.822	47.801
O2	7.770	15.965	341.119	341.119		22.373	5.518
TOTAL, kg-mol/h	149.066	259.526	3836.698	3836.698		301.563	74.372

TABLE 22
Stream Summary Information
(Page 1 of 9)

FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / -20°C AMBIENT AIR

Stream Description	Natural Gas Supply	Discharge from Natural Gas Compressor	Natural Gas Compressor GHP	Natural Gas Fuel to Combustor	Natural Gas to Process Unit	Pre-heated Natural Gas	Natural Gas Pre-heater Duty
Temperature K	288	426		426	426	616	
Pressure kPa	200	749		749	749	715	
Molar Flow kgmole/h	134.04	134.04		0.00	134.04	134.04	
Mass Flow kg/h	2257.4	2257.4		0.0	2257.4	2257.4	
Enthalpy kJ/h	1.287E+06	1.990E+06	7.032E+05	0.000E+00	1.990E+06	3.203E+06	1.212E+06
Density kg/m3	1.4125	3.5809		3.5809	3.5809	2.3479	
Mole Wt.	16.8416	16.8416		16.8416	16.8416	16.8416	
Spec. Heat kJ/kg-C	2.1213	2.499		2.499	2.499	3.1371	
Therm Cond W/m-K	0.0317	0.053		0.053	0.053	0.0862	
Viscosity cP	0.0111	0.0154		0.0154	0.0154	0.0201	
Z Factor	0.9951	0.9962		0.9962	0.9962	1.0006	
Component Flow, kg-mol/h							
CH4	128.676	128.676		0.000	128.676	128.676	
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000		0.000	0.000	0.000	
CO2	2.681	2.681		0.000	2.681	2.681	
H2	0.000	0.000		0.000	0.000	0.000	
H2O	0.000	0.000		0.000	0.000	0.000	
N2	2.681	2.681		0.000	2.681	2.681	
O2	0.000	0.000		0.000	0.000	0.000	
TOTAL, kg-mol/h	134.038	134.038		0.000	134.038	134.038	

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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / -20°C AMBIENT AIR

Stream Description	Demineralized Feedwater	Pre-heated Feedwater	Demin. Pre-heater Duty	Steam to Deaerator	Boiler Feedwater	Deaerator Vent	BFW Pump Suction
Temperature K	289	358		536	378	378	378
Pressure kPa	276	241		896	119	119	148
Molar Flow kgmole/h	297.61	297.61		10.02	307.50	0.13	307.50
Mass Flow kg/h	5361.5	5361.5		180.5	5539.7	2.3	5539.7
Enthalpy kJ/h	-1.049E+07	-8.876E+06	1.611E+06	1.777E+05	-8.700E+06	1.580E+03	-8.700E+06
Density kg/m ³	1014.4741	960.4306		3.7298	944.1871	0.6873	944.2007
Mole Wt.	18.0151	18.0151		18.0151	18.0151	18.0151	18.0151
Spec. Heat kJ/kg-C	4.3154	4.3654		2.0458	4.4086	1.9114	4.4085
Therm Cond W/m-K	0.5963	0.673		0.0404	0.6824	0.0248	0.6824
Viscosity cP	1.1197	0.3301		0.0186	0.2666	0.0122	0.2666
Z Factor	0.002	0.0015		0.9716	0.0007	0.9902	0.0009
Component Flow, kg-mol/h							
CH ₄	0.000	0.000		0.000	0.000	0.000	0.000
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.000	0.000		0.000	0.000	0.000	0.000
CO ₂	0.000	0.000		0.000	0.000	0.000	0.000
H ₂	0.000	0.000		0.000	0.000	0.000	0.000
H ₂ O	297.610	297.610		10.019	307.503	0.126	307.503
N ₂	0.000	0.000		0.000	0.000	0.000	0.000
O ₂	0.000	0.000		0.000	0.000	0.000	0.000
TOTAL, kg-mol/h	297.610	297.610		10.019	307.503	0.126	307.503

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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / -20°C AMBIENT AIR

Stream Description	BFW Pump Discharge	BFW Pump HP	BFW to Steam Drum	BFW Heater Duty	Boiler Duty	Boiler Blowdown	Boiler Steam
Temperature K	378		416			448	448
Pressure kPa	1448		1276			896	896
Molar Flow kgmole/h	307.50		307.50			0.38	307.13
Mass Flow kg/h	5539.7		5539.7			6.8	5532.9
Enthalpy kJ/h	-8.689E+06	1.089E+04	-7.732E+06	9.569E+05	1.219E+07	-8.494E+03	4.463E+06
Density kg/m3	944.6234		910.4422			879.3838	4.5488
Mole Wt.	18.0151		18.0151			18.0151	18.0151
Spec. Heat kJ/kg-C	4.4058		4.5373			4.7092	2.0286
Therm Cond W/m-K	0.6825		0.6875			0.6788	0.0308
Viscosity cP	0.266		0.1901			0.1535	0.0148
Z Factor	0.0088		0.0073			0.0049	0.952
Component Flow, kg-mol/h							
CH4	0.000		0.000			0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000		0.000			0.000	0.000
CO2	0.000		0.000			0.000	0.000
H2	0.000		0.000			0.000	0.000
H2O	307.503		307.503			0.378	307.125
N2	0.000		0.000			0.000	0.000
O2	0.000		0.000			0.000	0.000
TOTAL, kg-mol/h	307.503		307.503			0.378	307.125

TABLE 22
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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / -20°C AMBIENT AIR

Stream Description	Superheated Steam	Superheater Duty	Steam to Process Unit	Ambient Air	Air Compressor Discharge	Air Compressor GHP	Process Air to Combustor
Temperature K	536		536	244	463		463
Pressure kPa	896		896	86	621		621
Molar Flow kgmole/h	307.13		297.11	1926.24	1926.24		0.99
Mass Flow kg/h	5532.9		5352.4	55659.1	55659.1		28.5
Enthalpy kJ/h	5.448E+06	9.847E+05	5.270E+06	1.366E+07	2.600E+07	1.234E+07	1.334E+04
Density kg/m3	3.7298		3.7298	1.2278	4.6519		4.6519
Mole Wt.	18.0151		18.0151	28.8951	28.8951		28.8951
Spec. Heat kJ/kg-C	2.0458		2.0458	1.0079	1.0344		1.0344
Therm Cond W/m-K	0.0404		0.0404	0.022	0.0366		0.0366
Viscosity cP	0.0186		0.0186	0.016	0.0259		0.0259
Z Factor	0.9716		0.9716	0.9987	1.0013		1.0013
Component Flow, kg-mol/h							
CH4	0.000		0.000	0.000	0.000		0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000		0.000	0.000	0.000		0.000
CO2	0.000		0.000	5.837	5.837		0.003
H2	0.000		0.000	0.000	0.000		0.000
H2O	307.125		297.106	0.000	0.000		0.000
N2	0.000		0.000	1517.647	1517.647		0.778
O2	0.000		0.000	402.760	402.760		0.207
TOTAL, kg-mol/h	307.125		297.106	1926.244	1926.244		0.988

TABLE 22
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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / -20°C AMBIENT AIR

Stream Description	Process Air to Recycle Blower	Turbine Combustor Feed	Turbine Inlet	Turbine Exhaust	Turbine GHP	Flue Gas Exit N.G. Pre-heater	Flue Gas Exit Steam Superheater
Temperature K	463	978	978	671		656	643
Pressure kPa	621	603	594	105		104	103
Molar Flow kgmole/h	630.23	2357.39	2357.39	2357.39		2357.39	2357.39
Mass Flow kg/h	18210.4	63268.6	63268.6	63268.6		63268.6	63268.6
Enthalpy kJ/h	8.508E+06	7.428E+07	7.428E+07	4.926E+07	2.503E+07	4.804E+07	4.706E+07
Density kg/m3	4.6519	1.9874	1.9576	0.5041		0.5126	0.5192
Mole Wt.	28.8951	26.8384	26.8384	26.8384		26.8384	26.8384
Spec. Heat kJ/kg-C	1.0344	1.3398	1.3398	1.2416		1.2366	1.2326
Therm Cond W/m-K	0.0366	0.0693	0.0693	0.0492		0.0482	0.0473
Viscosity cP	0.0259	0.0386	0.0386	0.0292		0.0287	0.0284
Z Factor	1.0013	1.0012	1.0012	1.0001		1.0001	1.0001
Component Flow, kg-mol/h							
CH4	0.000	0.000	0.000	0.000		0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000	0.000	0.000		0.000	0.000
CO2	1.910	137.192	137.192	137.192		137.192	137.192
H2	0.000	0.000	0.000	0.000		0.000	0.000
H2O	0.000	554.450	554.450	554.450		554.450	554.450
N2	496.541	1520.333	1520.333	1520.333		1520.333	1520.333
O2	131.774	145.413	145.413	145.413		145.413	145.413
TOTAL, kg-mol/h	630.225	2357.388	2357.388	2357.388		2357.388	2357.388

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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / -20°C AMBIENT AIR

Stream Description	Flue Gas Exit Steam Boiler	Flue Gas Exit BFW Heater	Flue Gas Exit Demin Feedwater Heater	Anode Recycle Gas	Anode Recycle Gas	Ejector Exit Stream	Exchanger Feed Stream
Temperature K	484	471	449	978	978	851	812
Pressure kPa	103	102	101	598	598	607	607
Molar Flow kgmole/h	2357.39	2357.39	2357.39	17.26	17.26	25.51	29.23
Mass Flow kg/h	63268.6	63268.6	63268.6	520.4	520.4	669.0	731.8
Enthalpy kJ/h	3.487E+07	3.392E+07	3.231E+07	6.523E+05	6.523E+05	7.987E+05	8.877E+05
Density kg/m3	0.6861	0.7	0.7288	2.2173	2.2173	2.2527	2.2522
Mole Wt.	26.8384	26.8384	26.8384	30.1547	30.1547	26.2273	25.0319
Spec. Heat kJ/kg-C	1.1855	1.1821	1.1766	1.5441	1.5441	1.6501	1.8124
Therm Cond W/m-K	0.0364	0.0355	0.034	0.0792	0.0792	0.0706	0.0736
Viscosity cP	0.0231	0.0226	0.0218	0.0369	0.0369	0.0302	0.0287
Z Factor	0.9996	0.9995	0.9993	1.0002	1.0002	0.9985	0.9987
Component Flow, kg-mol/h							
CH4	0.000	0.000	0.000	0.035	0.035	0.035	3.610
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000	0.000	0.443	0.443	0.443	0.443
CO2	137.192	137.192	137.192	8.263	8.263	8.263	8.338
H2	0.000	0.000	0.000	0.630	0.630	0.630	0.630
H2O	554.450	554.450	554.450	7.844	7.844	16.097	16.097
N2	1520.333	1520.333	1520.333	0.041	0.041	0.041	0.116
O2	145.413	145.413	145.413	0.000	0.000	0.000	0.000
TOTAL, kg-mol/h	2357.388	2357.388	2357.388	17.257	17.257	25.510	29.233

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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / -20°C AMBIENT AIR

Stream Description	Pre-reformer Feed Stream	Reformer Feed Stream	Reformer Exit	Reformer Process Duty	Anode Feed	MCFC Stack Energy Recovery	Anode Exhaust
Temperature K	930	805	1061		950		978
Pressure kPa	603	603	603		599		598
Molar Flow kgmole/h	29.23	30.99	36.25		36.25		48.36
Mass Flow kg/h	731.8	731.8	731.8		731.8		1458.3
Enthalpy kJ/h	1.048E+06	9.053E+05	1.336E+06	1.012E+06	1.175E+06	1.797E+06	1.828E+06
Density kg/m3	1.9531	2.1265	1.3779		1.5306		2.2173
Mole Wt.	25.0319	23.6139	20.1843		20.1843		30.1547
Spec. Heat kJ/kg-C	1.8905	1.8367	1.9948		1.9448		1.5441
Therm Cond W/m-K	0.0853	0.0853	0.1308		0.1197		0.0792
Viscosity cP	0.0324	0.0289	0.0341		0.0315		0.0369
Z Factor	0.9997	0.9993	1.0007		1.0005		1.0002
Component Flow, kg-mol/h							
CH4	3.610	2.732	0.099		0.099		0.099
C2H6							
C3H8							
n-C4H10							
CO	0.443	0.639	5.001		5.001		1.242
CO2	8.338	9.020	7.290		7.290		23.157
H2	0.630	3.945	10.114		10.114		1.765
H2O	16.097	14.537	13.634		13.634		21.983
N2	0.116	0.116	0.116		0.116		0.116
O2	0.000	0.000	0.000		0.000		0.000
TOTAL, kg-mol/h	29.233	30.988	36.254		36.254		48.362

TABLE 22
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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / -20°C AMBIENT AIR

Stream Description	Anode Exhaust to Combustor	Cathode Recycle Gas	Cathode Recycle Gas	Cathode Recycle Bypass	Cathode Recycle to Combustor	Combined Feed to Combustor	Combustor Effluent
Temperature K	978	978	978	978	978	975	1072
Pressure kPa	598	603	603	603	603	598	597
Molar Flow kgmole/h	31.10	237.42	237.42	120.64	116.78	148.87	147.91
Mass Flow kg/h	938.0	6372.0	6372.0	3237.8	3134.2	4100.7	4100.7
Enthalpy kJ/h	1.176E+06	7.482E+06	7.482E+06	3.802E+06	3.680E+06	4.869E+06	5.416E+06
Density kg/m3	2.2173	1.9874	1.9874	1.9874	1.9874	2.0296	1.8553
Mole Wt.	30.1547	26.8384	26.8384	26.8384	26.8384	27.545	27.725
Spec. Heat kJ/kg-C	1.5441	1.3398	1.3398	1.3398	1.3398	1.3842	1.4127
Therm Cond W/m-K	0.0792	0.0693	0.0693	0.0693	0.0693	0.0712	0.0759
Viscosity cP	0.0369	0.0386	0.0386	0.0386	0.0386	0.0378	0.0405
Z Factor	1.0002	1.0012	1.0012	1.0012	1.0012	1.001	1.001
Component Flow, kg-mol/h							
CH4	0.064	0.000	0.000	0.000	0.000	0.064	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.799	0.000	0.000	0.000	0.000	0.799	0.000
CO2	14.894	13.817	13.817	7.021	6.796	21.693	22.556
H2	1.135	0.000	0.000	0.000	0.000	1.135	0.000
H2O	14.139	55.841	55.841	28.374	27.466	41.605	42.868
N2	0.074	153.119	153.119	77.804	75.315	76.168	76.168
O2	0.000	14.645	14.645	7.442	7.204	7.410	6.315
TOTAL, kg-mol/h	31.105	237.422	237.422	120.641	116.781	148.874	147.907

TABLE 22
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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / -20°C AMBIENT AIR

Stream Description	Reformer Flue Gas	Power Module Gas to Blower	Recycle Blower Suction	Recycle Blower Discharge	Recycle Blower GHP	Cathode Exhaust Gas	Cathode Exhaust to Expander
Temperature K	894	930	864	866		978	978
Pressure kPa	596	596	596	603		603	603
Molar Flow kgmole/h	147.91	268.55	3852.79	3852.79		302.90	65.48
Mass Flow kg/h	4100.7	7338.5	106272.8	106272.8		8129.5	1757.5
Enthalpy kJ/h	4.403E+06	8.205E+06	1.070E+08	1.074E+08	3.859E+05	9.545E+06	2.063E+06
Density kg/m3	2.2232	2.1049	2.2883	2.3046		1.9874	1.9874
Mole Wt.	27.725	27.3267	27.5833	27.5833		26.8384	26.8384
Spec. Heat kJ/kg-C	1.3564	1.3497	1.2926	1.2935		1.3398	1.3398
Therm Cond W/m-K	0.065	0.0669	0.0621	0.0623		0.0693	0.0693
Viscosity cP	0.0354	0.0367	0.0355	0.0356		0.0386	0.0386
Z Factor	1.0009	1.001	1.0011	1.0011		1.0012	1.0012
Component Flow, kg-mol/h							
CH4	0.000	0.000	0.000	0.000		0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000	0.000	0.000		0.000	0.000
CO2	22.556	29.577	356.830	356.830		17.628	3.811
H2	0.000	0.000	0.000	0.000		0.000	0.000
H2O	42.868	71.242	854.906	854.906		71.242	15.401
N2	76.168	153.972	2344.201	2344.201		195.350	42.231
O2	6.315	13.757	296.859	296.859		18.684	4.039
TOTAL, kg-mol/h	147.907	268.547	3852.795	3852.795		302.905	65.483

TABLE 23
Stream Summary Information
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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Natural Gas Supply	Discharge from Natural Gas Compressor	Natural Gas Compressor GHP	Natural Gas Fuel to Combustor	Natural Gas to Process Unit	Pre-heated Natural Gas	Natural Gas Pre-heater Duty
Temperature K	288	426		426	426	616	
Pressure kPa	200	749		749	749	715	
Molar Flow kgmole/h	134.04	134.04		0.00	134.04	134.04	
Mass Flow kg/h	2257.4	2257.4		0.0	2257.4	2257.4	
Enthalpy kJ/h	1.287E+06	1.990E+06	7.032E+05	0.000E+00	1.990E+06	3.203E+06	1.212E+06
Density kg/m3	1.4125	3.5809		3.5809	3.5809	2.3479	
Mole Wt.	16.8416	16.8416		16.8416	16.8416	16.8416	
Spec. Heat kJ/kg-C	2.1213	2.499		2.499	2.499	3.1371	
Therm Cond W/m-K	0.0317	0.053		0.053	0.053	0.0862	
Viscosity cP	0.0111	0.0154		0.0154	0.0154	0.0201	
Z Factor	0.9951	0.9962		0.9962	0.9962	1.0006	
Component Flow, kg-mol/h							
CH4	128.676	128.676		0.000	128.676	128.676	
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000		0.000	0.000	0.000	
CO2	2.681	2.681		0.000	2.681	2.681	
H2	0.000	0.000		0.000	0.000	0.000	
H2O	0.000	0.000		0.000	0.000	0.000	
N2	2.681	2.681		0.000	2.681	2.681	
O2	0.000	0.000		0.000	0.000	0.000	
TOTAL, kg-mol/h	134.038	134.038		0.000	134.038	134.038	

TABLE 23
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FULL-LOAD, START-OF-RUN OPERATION - 5.000' ELEV. / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Demineralized Feedwater	Pre-heated Feedwater	Demin. Pre-heater Duty	Steam to Deaerator	Boiler Feedwater	Deaerator Vent	BFW Pump Suction
Temperature K	289	358		536	378	378	378
Pressure kPa	276	241		896	119	119	148
Molar Flow kgmole/h	376.99	376.99		12.66	389.53	0.13	389.53
Mass Flow kg/h	6791.5	6791.5		228.1	7017.3	2.3	7017.3
Enthalpy kJ/h	-1.328E+07	-1.124E+07	2.041E+06	2.246E+05	-1.102E+07	1.569E+03	-1.102E+07
Density kg/m3	1014.4741	960.4306		3.7298	944.1871	0.6873	944.201
Mole Wt.	18.0151	18.0151		18.0151	18.0151	18.0151	18.0151
Spec. Heat kJ/kg-C	4.3154	4.3654		2.0458	4.4086	1.9114	4.4085
Therm Cond W/m-K	0.5963	0.673		0.0404	0.6824	0.0248	0.6824
Viscosity cP	1.1197	0.3301		0.0186	0.2666	0.0122	0.2666
Z Factor	0.002	0.0015		0.9716	0.0007	0.9902	0.0009
Component Flow, kg-mol/h							
CH4	0.000	0.000		0.000	0.000	0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000		0.000	0.000	0.000	0.000
CO2	0.000	0.000		0.000	0.000	0.000	0.000
H2	0.000	0.000		0.000	0.000	0.000	0.000
H2O	376.989	376.989		12.662	389.525	0.126	389.525
N2	0.000	0.000		0.000	0.000	0.000	0.000
O2	0.000	0.000		0.000	0.000	0.000	0.000
TOTAL, kg-mol/h	376.989	376.989		12.662	389.525	0.126	389.525

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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	BFW Pump Discharge	BFW Pump HP	BFW to Steam Drum	BFW Heater Duty	Boiler Duty	Boiler Blowdown	Boiler Steam
Temperature K	378		416			448	448
Pressure kPa	1448		1276			896	896
Molar Flow kgmole/h	389.53		389.53			0.38	389.15
Mass Flow kg/h	7017.3		7017.3			6.8	7010.5
Enthalpy kJ/h	-1.101E+07	1.380E+04	-9.795E+06	1.212E+06	1.544E+07	-8.494E+03	5.655E+06
Density kg/m3	944.6237		910.4422			879.3838	4.5488
Mole Wt.	18.0151		18.0151			18.0151	18.0151
Spec. Heat kJ/kg-C	4.4058		4.5373			4.7092	2.0286
Therm Cond W/m-K	0.6825		0.6875			0.6788	0.0308
Viscosity cP	0.266		0.1901			0.1535	0.0148
Z Factor	0.0088		0.0073			0.0049	0.952
Component Flow, kg-mol/h							
CH4	0.000		0.000			0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000		0.000			0.000	0.000
CO2	0.000		0.000			0.000	0.000
H2	0.000		0.000			0.000	0.000
H2O	389.525		389.525			0.378	389.148
N2	0.000		0.000			0.000	0.000
O2	0.000		0.000			0.000	0.000
TOTAL, kg-mol/h	389.525		389.525			0.378	389.148

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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Superheated Steam	Superheater Duty	Steam to Process Unit	Ambient Air	Air Compressor Discharge	Air Compressor GHP	Process Air to Combustor
Temperature K	536		536	322	605		605
Pressure kPa	896		896	86	621		621
Molar Flow kgmole/h	389.15		376.49	2504.67	2504.67		0.99
Mass Flow kg/h	7010.5		6782.4	72372.7	72372.7		28.5
Enthalpy kJ/h	6.903E+06	1.248E+06	6.678E+06	2.344E+07	4.455E+07	2.111E+07	1.757E+04
Density kg/m3	3.7298		3.7298	0.9303	3.5597		3.5597
Mole Wt.	18.0151		18.0151	28.8951	28.8951		28.8951
Spec. Heat kJ/kg-C	2.0458		2.0458	1.0116	1.0616		1.0616
Therm Cond W/m-K	0.0404		0.0404	0.0275	0.045		0.045
Viscosity cP	0.0186		0.0186	0.0199	0.0315		0.0315
Z Factor	0.9716		0.9716	0.9997	1.0019		1.0019
Component Flow, kg-mol/h							
CH4	0.000		0.000	0.000	0.000		0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000		0.000	0.000	0.000		0.000
CO2	0.000		0.000	7.590	7.590		0.003
H2	0.000		0.000	0.000	0.000		0.000
H2O	389.148		376.486	0.000	0.000		0.000
N2	0.000		0.000	1973.376	1973.376		0.778
O2	0.000		0.000	523.704	523.704		0.207
TOTAL, kg-mol/h	389.148		376.486	2504.670	2504.670		0.988

TABLE 23
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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Process Air to Recycle Blower	Turbine Combustor Feed	Turbine Inlet	Turbine Exhaust	Turbine GHP	Flue Gas Exit N.G. Pre-heater	Flue Gas Exit Steam Superheater
Temperature K	605	978	978	668		656	643
Pressure kPa	621	603	594	105		104	103
Molar Flow kgmole/h	823.03	3015.11	3015.11	3015.11		3015.11	3015.11
Mass Flow kg/h	23781.7	81409.7	81409.7	81409.7		81409.7	81409.7
Enthalpy kJ/h	1.464E+07	9.423E+07	9.423E+07	6.229E+07	3.194E+07	6.108E+07	5.983E+07
Density kg/m3	3.5597	1.9993	1.9693	0.5092		0.5153	0.5219
Mole Wt.	28.8951	27.0006	27.0006	27.0006		27.0006	27.0006
Spec. Heat kJ/kg-C	1.0616	1.3166	1.3166	1.2212		1.2174	1.2136
Therm Cond W/m-K	0.045	0.069	0.069	0.0491		0.0482	0.0474
Viscosity cP	0.0315	0.0393	0.0393	0.0295		0.0292	0.0288
Z Factor	1.0019	1.0012	1.0012	1.0001		1.0001	1.0001
Component Flow, kg-mol/h							
CH4	0.000	0.000	0.000	0.000		0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000	0.000	0.000		0.000	0.000
CO2	2.494	138.949	138.949	138.949		138.949	138.949
H2	0.000	0.000	0.000	0.000		0.000	0.000
H2O	0.000	633.851	633.851	633.851		633.851	633.851
N2	648.451	1975.985	1975.985	1975.985		1975.985	1975.985
O2	172.089	266.326	266.326	266.326		266.326	266.326
TOTAL, kg-mol/h	823.034	3015.111	3015.111	3015.111		3015.111	3015.111

TABLE 23
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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Flue Gas Exit Steam Boiler	Flue Gas Exit BFW Heater	Flue Gas Exit Demin Feedwater Heater	Anode Recycle Gas	Anode Recycle Gas	Ejector Exit Stream	Exchanger Feed Stream
Temperature K	484	471	450	978	978	838	806
Pressure kPa	103	102	101	598	598	607	607
Molar Flow kgmole/h	3015.11	3015.11	3015.11	19.28	19.28	29.74	33.46
Mass Flow kg/h	81409.7	81409.7	81409.7	565.2	565.2	753.6	816.3
Enthalpy kJ/h	4.439E+07	4.318E+07	4.114E+07	7.248E+05	7.248E+05	9.104E+05	9.993E+05
Density kg/m ³	0.6893	0.7033	0.7321	2.1559	2.1559	2.2099	2.2127
Mole Wt.	27.0006	27.0006	27.0006	29.3156	29.3156	25.3415	24.3957
Spec. Heat kJ/kg-C	1.1682	1.165	1.1597	1.575	1.575	1.6864	1.8282
Therm Cond W/m-K	0.0366	0.0358	0.0343	0.08	0.08	0.07	0.0726
Viscosity cP	0.0235	0.023	0.0222	0.0365	0.0365	0.029	0.028
Z Factor	0.9996	0.9996	0.9994	1	1	0.9982	0.9984
Component Flow, kg-mol/h							
CH ₄	0.000	0.000	0.000	0.025	0.025	0.025	3.600
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.000	0.000	0.000	0.443	0.443	0.443	0.443
CO ₂	138.949	138.949	138.949	8.641	8.641	8.641	8.715
H ₂	0.000	0.000	0.000	0.722	0.722	0.722	0.722
H ₂ O	633.851	633.851	633.851	9.406	9.406	19.864	19.864
N ₂	1975.985	1975.985	1975.985	0.043	0.043	0.043	0.118
O ₂	266.326	266.326	266.326	0.000	0.000	0.000	0.000
TOTAL, kg-mol/h	3015.111	3015.111	3015.111	19.280	19.280	29.738	33.461

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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Pre-reformer Feed Stream	Reformer Feed Stream	Reformer Exit	Reformer Process Duty	Anode Feed	MCFC Stack Energy Recovery	Anode Exhaust
Temperature K	924	802	1061		950		978
Pressure kPa	603	603	603		599		598
Molar Flow kgmole/h	33.46	35.41	40.52		40.52		52.63
Mass Flow kg/h	816.3	816.3	816.3		816.3		1542.9
Enthalpy kJ/h	1.180E+06	1.024E+06	1.503E+06	1.036E+06	1.322E+06	1.797E+06	1.979E+06
Density kg/m ³	1.9164	2.0844	1.3753		1.5278		2.1559
Mole Wt.	24.3957	23.0506	20.1447		20.1447		29.3156
Spec. Heat kJ/kg-C	1.9058	1.8555	2.0147		1.9638		1.575
Therm Cond W/m-K	0.0843	0.0846	0.1284		0.1174		0.08
Viscosity cP	0.032	0.0286	0.0339		0.0313		0.0365
Z Factor	0.9995	0.999	1.0006		1.0004		1
Component Flow, kg-mol/h							
CH ₄	3.600	2.623	0.069		0.069		0.069
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.443	0.602	4.649		4.649		1.209
CO ₂	8.715	9.532	8.039		8.039		23.588
H ₂	0.722	4.468	10.638		10.638		1.971
H ₂ O	19.864	18.071	17.009		17.009		25.677
N ₂	0.118	0.118	0.118		0.118		0.118
O ₂	0.000	0.000	0.000		0.000		0.000
TOTAL, kg-mol/h	33.461	35.414	40.522		40.522		52.630

TABLE 23
(Page 8 of 9)

FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Anode Exhaust to Combustor	Cathode Recycle Gas	Cathode Recycle Gas	Cathode Recycle Bypass	Cathode Recycle to Combustor	Combined Feed to Combustor	Combustor Effluent
Temperature K	978	978	978	978	978	976	1072
Pressure kPa	598	603	603	603	603	598	597
Molar Flow kmole/h	33.35	218.31	218.31	100.69	117.61	151.95	150.94
Mass Flow kg/h	977.7	5894.4	5894.4	2718.8	3175.6	4181.8	4181.8
Enthalpy kJ/h	1.254E+06	6.823E+06	6.823E+06	3.147E+06	3.676E+06	4.947E+06	5.496E+06
Density kg/m ³	2.1559	1.9993	1.9993	1.9993	1.9993	2.0262	1.8539
Mole Wt.	29.3156	27.0005	27.0006	27.0005	27.0005	27.521	27.7046
Spec. Heat kJ/kg-C	1.575	1.3166	1.3166	1.3166	1.3166	1.3751	1.4025
Therm Cond W/m-K	0.08	0.069	0.069	0.069	0.069	0.0712	0.0758
Viscosity cP	0.0365	0.0393	0.0393	0.0393	0.0393	0.038	0.0408
Z Factor	1	1.0012	1.0012	1.0012	1.0012	1.001	1.0011
Component Flow, kg-mol/h							
CH ₄	0.044	0.000	0.000	0.000	0.000	0.044	0.000
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.766	0.000	0.000	0.000	0.000	0.766	0.000
CO ₂	14.947	10.061	10.061	4.640	5.420	20.370	21.180
H ₂	1.249	0.000	0.000	0.000	0.000	1.249	0.000
H ₂ O	16.271	45.894	45.894	21.169	24.725	40.996	42.332
N ₂	0.074	143.068	143.070	65.990	77.078	77.931	77.931
O ₂	0.000	19.283	19.283	8.894	10.388	10.595	9.500
TOTAL, kg-mol/h	33.350	218.306	218.308	100.694	117.612	151.950	150.943

TABLE 23
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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Reformer Flue Gas	Power Module Gas to Blower	Recycle Blower Suction	Recycle Blower Discharge	Recycle Blower GHP	Cathode Exhaust Gas	Cathode Exhaust to Expander
Temperature K	892	925	864	866		978	978
Pressure kPa	596	596	596	603		603	603
Molar Flow kgmole/h	150.94	251.64	3842.67	3842.67		302.06	83.75
Mass Flow kg/h	4181.8	6900.6	106588.8	106588.8		8155.8	2261.4
Enthalpy kJ/h	4.459E+06	7.606E+06	1.059E+08	1.063E+08	3.850E+05	9.441E+06	2.618E+06
Density kg/m ³	2.2264	2.1245	2.301	2.3173		1.9993	1.9993
Mole Wt.	27.7046	27.4229	27.7382	27.7382		27.0006	27.0006
Spec. Heat kJ/kg-C	1.3467	1.3355	1.2725	1.2734		1.3166	1.3166
Therm Cond W/m-K	0.0648	0.0665	0.0619	0.0621		0.069	0.069
Viscosity cP	0.0355	0.0368	0.036	0.0361		0.0393	0.0393
Z Factor	1.0009	1.0011	1.0012	1.0012		1.0012	1.0012
Component Flow, kg-mol/h							
CH ₄	0.000	0.000	0.000	0.000		0.000	0.000
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.000	0.000	0.000	0.000		0.000	0.000
CO ₂	21.180	25.820	312.337	312.337		13.920	3.860
H ₂	0.000	0.000	0.000	0.000		0.000	0.000
H ₂ O	42.332	63.501	762.008	762.008		63.501	17.607
N ₂	77.931	143.921	2375.506	2375.506		197.959	54.888
O ₂	9.500	18.394	392.821	392.821		26.681	7.398
TOTAL, kg-mol/h	150.943	251.636	3842.672	3842.672		302.061	83.753

TABLE 24
Stream Summary Information
(Page 1 of 9)

FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (100% REL. HUMIDITY) AMBIENT AIR

Stream Description	Natural Gas Supply	Discharge from Natural Gas Compressor	Natural Gas Compressor GHP	Natural Gas Fuel to Combustor	Natural Gas to Process Unit	Pre-heated Natural Gas	Natural Gas Pre-heater Duty
Temperature K	288	426		426	426	616	
Pressure kPa	200	749		749	749	715	
Molar Flow kgmole/h	134.04	134.04		0.00	134.04	134.04	
Mass Flow kg/h	2257.4	2257.4		0.0	2257.4	2257.4	
Enthalpy kJ/h	1.287E+06	1.990E+06	7.032E+05	0.000E+00	1.990E+06	3.203E+06	1.212E+06
Density kg/m ³	1.4125	3.5809		3.5809	3.5809	2.3479	
Mole Wt.	16.8416	16.8416		16.8416	16.8416	16.8416	
Spec. Heat kJ/kg-C	2.1213	2.499		2.499	2.499	3.1371	
Therm Cond W/m-K	0.0317	0.053		0.053	0.053	0.0862	
Viscosity cP	0.0111	0.0154		0.0154	0.0154	0.0201	
Z Factor	0.9951	0.9962		0.9962	0.9962	1.0006	
Component Flow, kg-mol/h							
CH ₄	128.676	128.676		0.000	128.676	128.676	
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.000	0.000		0.000	0.000	0.000	
CO ₂	2.681	2.681		0.000	2.681	2.681	
H ₂	0.000	0.000		0.000	0.000	0.000	
H ₂ O	0.000	0.000		0.000	0.000	0.000	
N ₂	2.681	2.681		0.000	2.681	2.681	
O ₂	0.000	0.000		0.000	0.000	0.000	
TOTAL, kg-mol/h	134.038	134.038		0.000	134.038	134.038	

TABLE 24
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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (100% REL. HUMIDITY) AMBIENT AIR

Stream Description	Demineralized Feedwater	Pre-heated Feedwater	Demin. Pre-heater Duty	Steam to Deaerator	Boiler Feedwater	Deaerator Vent	BFW Pump Suction
Temperature K	289	358		536	378	378	378
Pressure kPa	276	241		896	119	119	148
Molar Flow kgmole/h	383.79	383.79		12.89	396.56	0.13	396.56
Mass Flow kg/h	6914.1	6914.1		232.2	7144.0	2.3	7144.0
Enthalpy kJ/h	-1.352E+07	-1.145E+07	2.078E+06	2.286E+05	-1.122E+07	1.571E+03	-1.122E+07
Density kg/m3	1014.4741	960.4306		3.7298	944.1871	0.6873	944.201
Mole Wt.	18.0151	18.0151		18.0151	18.0151	18.0151	18.0151
Spec. Heat kJ/kg-C	4.3154	4.3654		2.0458	4.4086	1.9114	4.4085
Therm Cond W/m-K	0.5963	0.673		0.0404	0.6824	0.0248	0.6824
Viscosity cP	1.1197	0.3301		0.0186	0.2666	0.0122	0.2666
Z Factor	0.002	0.0015		0.9716	0.0007	0.9902	0.0009
Component Flow, kg-mol/h							
CH4	0.000	0.000		0.000	0.000	0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000		0.000	0.000	0.000	0.000
CO2	0.000	0.000		0.000	0.000	0.000	0.000
H2	0.000	0.000		0.000	0.000	0.000	0.000
H2O	383.793	383.793		12.888	396.556	0.126	396.556
N2	0.000	0.000		0.000	0.000	0.000	0.000
O2	0.000	0.000		0.000	0.000	0.000	0.000
TOTAL, kg-mol/h	383.793	383.793		12.888	396.556	0.126	396.556

TABLE 24
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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (100% REL. HUMIDITY) AMBIENT AIR

Stream Description	BFW Pump Discharge	BFW Pump HP	BFW to Steam Drum	BFW Heater Duty	Boiler Duty	Boiler Blowdown	Boiler Steam
Temperature K	378		416			448	448
Pressure kPa	1448		1276			896	896
Molar Flow kgmole/h	396.56		396.56			0.38	396.18
Mass Flow kg/h	7144.0		7144.0			6.8	7137.2
Enthalpy kJ/h	-1.121E+07	1.405E+04	-9.972E+06	1.234E+06	1.572E+07	-8.494E+03	5.758E+06
Density kg/m3	944.6237		910.4422			879.3838	4.5488
Mole Wt.	18.0151		18.0151			18.0151	18.0151
Spec. Heat kJ/kg-C	4.4058		4.5373			4.7092	2.0286
Therm Cond W/m-K	0.6825		0.6875			0.6788	0.0308
Viscosity cP	0.266		0.1901			0.1535	0.0148
Z Factor	0.0088		0.0073			0.0049	0.952
Component Flow, kg-mol/h							
CH4	0.000		0.000			0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000		0.000			0.000	0.000
CO2	0.000		0.000			0.000	0.000
H2	0.000		0.000			0.000	0.000
H2O	396.556		396.556			0.378	396.178
N2	0.000		0.000			0.000	0.000
O2	0.000		0.000			0.000	0.000
TOTAL, kg-mol/h	396.556		396.556			0.378	396.178

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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (100% REL. HUMIDITY) AMBIENT AIR

Stream Description	Superheated Steam	Superheater Duty	Steam to Process Unit	Ambient Air	Air Compressor Discharge	Air Compressor GHP	Process Air to Combustor
Temperature K	536		536	322	598		598
Pressure kPa	896		896	86	621		621
Molar Flow kgmole/h	396.18		383.29	2388.71	2388.71		0.99
Mass Flow kg/h	7137.2		6905.0	66028.4	66028.4		27.3
Enthalpy kJ/h	7.028E+06	1.270E+06	6.799E+06	2.271E+07	4.271E+07	2.001E+07	1.767E+04
Density kg/m3	3.7298		3.7298	0.8908	3.4474		3.4474
Mole Wt.	18.0151		18.0151	27.6418	27.6418		27.6418
Spec. Heat kJ/kg-C	2.0458		2.0458	1.0773	1.1339		1.1339
Therm Cond W/m-K	0.0404		0.0404	0.0267	0.0448		0.0448
Viscosity cP	0.0186		0.0186	0.0182	0.0289		0.0289
Z Factor	0.9716		0.9716	0.9988	1.0011		1.0011
Component Flow, kg-mol/h							
CH4	0.000		0.000	0.000	0.000		0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000		0.000	0.000	0.000		0.000
CO2	0.000		0.000	6.405	6.405		0.003
H2	0.000		0.000	0.000	0.000		0.000
H2O	396.178		383.290	275.160	275.160		0.114
N2	0.000		0.000	1665.224	1665.224		0.689
O2	0.000		0.000	441.925	441.925		0.183
TOTAL, kg-mol/h	396.178		383.290	2388.714	2388.714		0.988

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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (100% REL. HUMIDITY) AMBIENT AIR

Stream Description	Process Air to Recycle Blower	Turbine Combustor Feed	Turbine Inlet	Turbine Exhaust	Turbine GHP	Flue Gas Exit N.G. Pre- heater	Flue Gas Exit Steam Superheater
Temperature K	598	978	978	674		662	649
Pressure kPa	621	603	594	105		104	103
Molar Flow kgmole/h	784.38	2906.04	2906.04	2906.04		2906.04	2906.04
Mass Flow kg/h	21681.8	75190.7	75190.7	75190.7		75190.7	75190.7
Enthalpy kJ/h	1.403E+07	9.250E+07	9.250E+07	6.158E+07	3.092E+07	6.037E+07	5.910E+07
Density kg/m3	3.4474	1.9164	1.8877	0.4839		0.4898	0.4963
Mole Wt.	27.6418	25.8739	25.8739	25.8739		25.8739	25.8739
Spec. Heat kJ/kg-C	1.1339	1.4077	1.4077	1.3029		1.2986	1.2942
Therm Cond W/m-K	0.0448	0.0708	0.0708	0.05		0.0491	0.0481
Viscosity cP	0.0289	0.0371	0.0371	0.0284		0.028	0.0276
Z Factor	1.0011	1.001	1.001	1		1	0.9999
Component Flow, kg-mol/h							
CH4	0.000	0.000	0.000	0.000		0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000	0.000	0.000		0.000	0.000
CO2	2.103	137.770	137.770	137.770		137.770	137.770
H2	0.000	0.000	0.000	0.000		0.000	0.000
H2O	90.354	915.799	915.799	915.799		915.799	915.799
N2	546.810	1667.905	1667.905	1667.905		1667.905	1667.905
O2	145.115	184.569	184.569	184.569		184.569	184.569
TOTAL, kg-mol/h	784.382	2906.043	2906.043	2906.043		2906.043	2906.043

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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (100% REL. HUMIDITY) AMBIENT AIR

Stream Description	Flue Gas Exit Steam Boiler	Flue Gas Exit BFW Heater	Flue Gas Exit Demin Feedwater Heater	Anode Recycle Gas	Anode Recycle Gas	Ejector Exit Stream	Exchanger Feed Stream
Temperature K	484	470	448	978	978	839	807
Pressure kPa	103	102	101	598	598	607	607
Molar Flow kgmole/h	2906.04	2906.04	2906.04	19.67	19.67	30.32	34.04
Mass Flow kg/h	75190.7	75190.7	75190.7	575.4	575.4	767.2	829.9
Enthalpy kJ/h	4.338E+07	4.214E+07	4.007E+07	7.393E+05	7.393E+05	9.282E+05	1.017E+06
Density kg/m3	0.6617	0.6758	0.7049	2.151	2.151	2.2062	2.2094
Mole Wt.	25.8739	25.8739	25.8739	29.2483	29.2483	25.3038	24.3783
Spec. Heat kJ/kg-C	1.2414	1.2377	1.2316	1.5775	1.5775	1.6883	1.8277
Therm Cond W/m-K	0.0362	0.0353	0.0337	0.08	0.08	0.07	0.0726
Viscosity cP	0.022	0.0214	0.0205	0.0365	0.0365	0.029	0.028
Z Factor	0.9993	0.9992	0.999	1	1	0.9982	0.9984
Component Flow, kg-mol/h							
CH4	0.000	0.000	0.000	0.024	0.024	0.024	3.599
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000	0.000	0.448	0.448	0.448	0.448
CO2	137.770	137.770	137.770	8.769	8.769	8.769	8.843
H2	0.000	0.000	0.000	0.739	0.739	0.739	0.739
H2O	915.799	915.799	915.799	9.650	9.650	20.297	20.297
N2	1667.905	1667.905	1667.905	0.044	0.044	0.044	0.118
O2	184.569	184.569	184.569	0.000	0.000	0.000	0.000
TOTAL, kg-mol/h	2906.043	2906.043	2906.043	19.674	19.674	30.320	34.044

TABLE 24
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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (100% REL. HUMIDITY) AMBIENT AIR

Stream Description	Pre-reformer Feed Stream	Reformer Feed Stream	Reformer Exit	Reformer Process Duty	Anode Feed	MCFC Stack Energy Recovery	Anode Exhaust
Temperature K	925	803	1061		950		978
Pressure kPa	603	603	603		599		598
Molar Flow kgmole/h	34.04	36.03	41.11		41.11		53.22
Mass Flow kg/h	829.9	829.9	829.9		829.9		1556.5
Enthalpy kJ/h	1.200E+06	1.041E+06	1.526E+06	1.039E+06	1.343E+06	1.797E+06	2.000E+06
Density kg/m3	1.9138	2.0818	1.3783		1.5311		2.151
Mole Wt.	24.3783	23.0342	20.1885		20.1885		29.2483
Spec. Heat kJ/kg-C	1.905	1.8549	2.013		1.962		1.5775
Therm Cond W/m-K	0.0843	0.0846	0.1278		0.1169		0.08
Viscosity cP	0.032	0.0286	0.0339		0.0313		0.0365
Z Factor	0.9994	0.999	1.0006		1.0004		1
Component Flow, kg-mol/h							
CH4	3.599	2.605	0.066		0.066		0.066
C2H6							
C3H8							
n-C4H10							
CO	0.448	0.611	4.639		4.639		1.212
CO2	8.843	9.674	8.185		8.185		23.720
H2	0.739	4.549	10.679		10.679		1.998
H2O	20.297	18.473	17.422		17.422		26.103
N2	0.118	0.118	0.118		0.118		0.118
O2	0.000	0.000	0.000		0.000		0.000
TOTAL, kg-mol/h	34.044	36.030	41.109		41.109		53.217

TABLE 24
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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (100% REL. HUMIDITY) AMBIENT AIR

Stream Description	Anode Exhaust to Combustor	Cathode Recycle Gas	Cathode Recycle Gas	Cathode Recycle Bypass	Cathode Recycle to Combustor	Combined Feed to Combustor	Combustor Effluent
Temperature K	978	978	978	978	978	975	1072
Pressure kPa	598	603	603	603	603	598	597
Molar Flow kgmole/h	33.54	213.99	213.99	99.48	114.51	149.05	148.03
Mass Flow kg/h	981.1	5536.9	5536.9	2573.9	2962.9	3971.3	3971.3
Enthalpy kJ/h	1.261E+06	6.811E+06	6.811E+06	3.166E+06	3.645E+06	4.923E+06	5.472E+06
Density kg/m3	2.151	1.9164	1.9164	1.9164	1.9164	1.9624	1.7955
Mole Wt.	29.2483	25.8739	25.8739	25.8739	25.8739	26.645	26.8271
Spec. Heat kJ/kg-C	1.5775	1.4077	1.4077	1.4077	1.4077	1.4477	1.4775
Therm Cond W/m-K	0.08	0.0708	0.0708	0.0708	0.0708	0.0727	0.0774
Viscosity cP	0.0365	0.0371	0.0371	0.0371	0.0371	0.0367	0.0393
Z Factor	1	1.001	1.001	1.001	1.001	1.0008	1.0009
Component Flow, kg-mol/h							
CH4	0.042	0.000	0.000	0.000	0.000	0.042	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.764	0.000	0.000	0.000	0.000	0.764	0.000
CO2	14.951	10.145	10.145	4.716	5.429	20.383	21.188
H2	1.259	0.000	0.000	0.000	0.000	1.259	0.000
H2O	16.453	67.437	67.437	31.350	36.088	52.655	53.997
N2	0.074	122.821	122.821	57.096	65.725	66.488	66.488
O2	0.000	13.591	13.591	6.318	7.273	7.456	6.361
TOTAL, kg-mol/h	33.544	213.994	213.994	99.480	114.514	149.046	148.034

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FULL-LOAD, START-OF-RUN OPERATION - 5,000' ELEV. / 49°C (100% REL. HUMIDITY) AMBIENT AIR

Stream Description	Reformer Flue Gas	Power Module Gas to Blower	Recycle Blower Suction	Recycle Blower Discharge	Recycle Blower GHP	Cathode Exhaust Gas	Cathode Exhaust to Expander
Temperature K	891	925	864	866		978	978
Pressure kPa	596	596	596	603		603	603
Molar Flow kgmole/h	148.03	247.51	3754.55	3754.55		294.72	80.72
Mass Flow kg/h	3971.3	6545.3	100225.0	100225.0		7625.5	2088.6
Enthalpy kJ/h	4.433E+06	7.599E+06	1.052E+08	1.056E+08	3.760E+05	9.381E+06	2.569E+06
Density kg/m3	2.1578	2.0493	2.2151	2.2309		1.9164	1.9164
Mole Wt.	26.8271	26.444	26.6943	26.6943		25.8739	25.8739
Spec. Heat kJ/kg-C	1.4168	1.4139	1.3516	1.3526		1.4077	1.4077
Therm Cond W/m-K	0.066	0.068	0.0633	0.0635		0.0708	0.0708
Viscosity cP	0.0344	0.0354	0.0344	0.0344		0.0371	0.0371
Z Factor	1.0006	1.0008	1.0008	1.0009		1.001	1.001
Component Flow, kg-mol/h							
CH4	0.000	0.000	0.000	0.000		0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000	0.000	0.000		0.000	0.000
CO2	21.188	25.905	312.958	312.958		13.972	3.827
H2	0.000	0.000	0.000	0.000		0.000	0.000
H2O	53.997	85.347	1114.514	1114.514		92.876	25.439
N2	66.488	123.584	2029.817	2029.817		169.151	46.331
O2	6.361	12.679	297.264	297.264		18.718	5.127
TOTAL, kg-mol/h	148.034	247.514	3754.553	3754.553		294.718	80.723

TABLE 25
Stream Summary Information
(Page 1 of 9)

FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR

Stream Description	Natural Gas Supply	Discharge from Natural Gas Compressor	Natural Gas Compressor GHP	Natural Gas Fuel to Combustor	Natural Gas to Process Unit	Pre-heated Natural Gas	Natural Gas Pre-heater Duty
Temperature K	288	426		426	426	616	
Pressure kPa	200	749		749	749	715	
Molar Flow kgmole/h	134.04	134.04		0.00	134.04	134.04	
Mass Flow kg/h	2257.4	2257.4		0.0	2257.4	2257.4	
Enthalpy kJ/h	1.287E+06	1.990E+06	7.032E+05	0.000E+00	1.990E+06	3.203E+06	1.212E+06
Density kg/m3	1.4125	3.5809		3.5809	3.5809	2.3479	
Mole Wt.	16.8416	16.8416		16.8416	16.8416	16.8416	
Spec. Heat kJ/kg-C	2.1213	2.499		2.499	2.499	3.1371	
Therm Cond W/m-K	0.0317	0.053		0.053	0.053	0.0862	
Viscosity cP	0.0111	0.0154		0.0154	0.0154	0.0201	
Z Factor	0.9951	0.9962		0.9962	0.9962	1.0006	
Component Flow, kg-mol/h							
CH4	128.676	128.676		0.000	128.676	128.676	
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000		0.000	0.000	0.000	
CO2	2.681	2.681		0.000	2.681	2.681	
H2	0.000	0.000		0.000	0.000	0.000	
H2O	0.000	0.000		0.000	0.000	0.000	
N2	2.681	2.681		0.000	2.681	2.681	
O2	0.000	0.000		0.000	0.000	0.000	
TOTAL, kg-mol/h	134.038	134.038		0.000	134.038	134.038	

TABLE 25
(Page 2 of 9)

FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR

Stream Description	Demineralized Feedwater	Pre-heated Feedwater	Demin. Pre-heater Duty	Steam to Deaerator	Boiler Feedwater	Deaerator Vent	BFW Pump Suction
Temperature K	289	358		536	378	378	378
Pressure kPa	276	241		896	119	119	148
Molar Flow kgmole/h	365.65	365.65		12.28	377.81	0.13	377.81
Mass Flow kg/h	6587.2	6587.2		221.3	6806.2	2.3	6806.2
Enthalpy kJ/h	-1.289E+07	-1.091E+07	1.979E+06	2.179E+05	-1.069E+07	1.573E+03	-1.069E+07
Density kg/m ³	1014.4741	960.4306		3.7298	944.1871	0.6873	944.2007
Mole Wt.	18.0151	18.0151		18.0151	18.0151	18.0151	18.0151
Spec. Heat kJ/kg-C	4.3154	4.3654		2.0458	4.4086	1.9114	4.4085
Therm Cond W/m-K	0.5963	0.673		0.0404	0.6824	0.0248	0.6824
Viscosity cP	1.1197	0.3301		0.0186	0.2666	0.0122	0.2666
Z Factor	0.002	0.0015		0.9716	0.0007	0.9902	0.0009
Component Flow, kg-mol/h							
CH ₄	0.000	0.000		0.000	0.000	0.000	0.000
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.000	0.000		0.000	0.000	0.000	0.000
CO ₂	0.000	0.000		0.000	0.000	0.000	0.000
H ₂	0.000	0.000		0.000	0.000	0.000	0.000
H ₂ O	365.649	365.649		12.285	377.808	0.126	377.808
N ₂	0.000	0.000		0.000	0.000	0.000	0.000
O ₂	0.000	0.000		0.000	0.000	0.000	0.000
TOTAL, kg-mol/h	365.649	365.649		12.285	377.808	0.126	377.808

TABLE 25
(Page 3 of 9)

FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR

Stream Description	BFW Pump Discharge	BFW Pump HP	BFW to Steam Drum	BFW Heater Duty	Boiler Duty	Boiler Blowdown	Boiler Steam
Temperature K	378		416			448	448
Pressure kPa	1448		1276			896	896
Molar Flow kgmole/h	377.81		377.81			0.38	377.43
Mass Flow kg/h	6806.2		6806.2			6.8	6799.4
Enthalpy kJ/h	-1.068E+07	1.338E+04	-9.500E+06	1.176E+06	1.498E+07	-8.494E+03	5.485E+06
Density kg/m ³	944.6234		910.4422			879.3838	4.5488
Mole Wt.	18.0151		18.0151			18.0151	18.0151
Spec. Heat kJ/kg-C	4.4058		4.5373			4.7092	2.0286
Therm Cond W/m-K	0.6825		0.6875			0.6788	0.0308
Viscosity cP	0.266		0.1901			0.1535	0.0148
Z Factor	0.0088		0.0073			0.0049	0.952
Component Flow, kg-mol/h							
CH ₄	0.000		0.000			0.000	0.000
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.000		0.000			0.000	0.000
CO ₂	0.000		0.000			0.000	0.000
H ₂	0.000		0.000			0.000	0.000
H ₂ O	377.808		377.808			0.378	377.430
N ₂	0.000		0.000			0.000	0.000
O ₂	0.000		0.000			0.000	0.000
TOTAL, kg-mol/h	377.808		377.808			0.378	377.430

TABLE 25
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FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR

Stream Description	Superheated Steam	Superheater Duty	Steam to Process Unit	Ambient Air	Air Compressor Discharge	Air Compressor GHP	Process Air to Combustor
Temperature K	536		536	288	518		518
Pressure kPa	896		896	101	621		621
Molar Flow kgmole/h	377.43		365.15	2399.53	2399.53		0.99
Mass Flow kg/h	6799.4		6578.1	69073.6	69073.6		28.4
Enthalpy kJ/h	6.695E+06	1.210E+06	6.477E+06	2.011E+07	3.640E+07	1.629E+07	1.499E+04
Density kg/m3	3.7298		3.7298	1.2105	4.1419		4.1419
Mole Wt.	18.0151		18.0151	28.7863	28.7863		28.7863
Spec. Heat kJ/kg-C	2.0458		2.0458	1.015	1.0499		1.0499
Therm Cond W/m-K	0.0404		0.0404	0.0251	0.0399		0.0399
Viscosity cP	0.0186		0.0186	0.0182	0.0279		0.0279
Z Factor	0.9716		0.9716	0.9992	1.0016		1.0016
Component Flow, kg-mol/h							
CH4	0.000		0.000	0.000	0.000		0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000		0.000	0.000	0.000		0.000
CO2	0.000		0.000	7.199	7.199		0.003
H2	0.000		0.000	0.000	0.000		0.000
H2O	377.430		365.146	23.995	23.995		0.010
N2	0.000		0.000	1871.633	1871.633		0.771
O2	0.000		0.000	496.703	496.703		0.205
TOTAL, kg-mol/h	377.430		365.146	2399.530	2399.530		0.988

TABLE 25
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FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR

Stream Description	Process Air to Recycle Blower	Turbine Combustor Feed	Turbine Inlet	Turbine Exhaust	Turbine GHP	Flue Gas Exit N.G. Pre- heater	Flue Gas Exit Steam Superheater
Temperature K	518	978	978	669		657	644
Pressure kPa	621	603	594	105		104	103
Molar Flow kgmole/h	787.99	2898.71	2898.71	2898.71		2898.71	2898.71
Mass Flow kg/h	22683.3	77908.5	77908.5	77908.5		77908.5	77908.5
Enthalpy kJ/h	1.195E+07	9.085E+07	9.085E+07	6.012E+07	3.073E+07	5.890E+07	5.769E+07
Density kg/m3	4.1419	1.9902	1.9604	0.5062		0.5126	0.5192
Mole Wt.	28.7863	26.877	26.877	26.877		26.877	26.877
Spec. Heat kJ/kg-C	1.0499	1.3276	1.3276	1.231		1.227	1.2231
Therm Cond W/m-K	0.0399	0.0692	0.0692	0.0492		0.0483	0.0475
Viscosity cP	0.0279	0.039	0.039	0.0294		0.029	0.0286
Z Factor	1.0016	1.0012	1.0012	1.0001		1.0001	1.0001
Component Flow, kg-mol/h							
CH4	0.000	0.000	0.000	0.000		0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000	0.000	0.000		0.000	0.000
CO2	2.364	138.567	138.567	138.567		138.567	138.567
H2	0.000	0.000	0.000	0.000		0.000	0.000
H2O	7.880	646.523	646.523	646.523		646.523	646.523
N2	614.630	1874.294	1874.294	1874.294		1874.294	1874.294
O2	163.113	239.327	239.327	239.327		239.327	239.327
TOTAL, kg-mol/h	787.987	2898.710	2898.710	2898.710		2898.710	2898.710

TABLE 25
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FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR

Stream Description	Flue Gas Exit Steam Boiler	Flue Gas Exit BFW Heater	Flue Gas Exit Demin Feedwater Heater	Anode Recycle Gas	Anode Recycle Gas	Ejector Exit Stream	Exchanger Feed Stream
Temperature K	484	471	449	978	978	838	805
Pressure kPa	103	102	101	598	598	607	607
Molar Flow kgmole/h	2898.71	2898.71	2898.71	18.60	18.60	28.74	32.46
Mass Flow kg/h	77908.5	77908.5	77908.5	548.2	548.2	730.9	793.6
Enthalpy kJ/h	4.272E+07	4.154E+07	3.956E+07	7.004E+05	7.004E+05	8.803E+05	9.693E+05
Density kg/m ³	0.687	0.7011	0.7301	2.1678	2.1678	2.2185	2.2205
Mole Wt.	26.877	26.877	26.877	29.4773	29.4773	25.432	24.4467
Spec. Heat kJ/kg-C	1.1767	1.1735	1.1681	1.5706	1.5706	1.6829	1.8288
Therm Cond W/m-K	0.0365	0.0357	0.0342	0.0796	0.0796	0.0698	0.0725
Viscosity cP	0.0233	0.0228	0.022	0.0366	0.0366	0.0291	0.0279
Z Factor	0.9996	0.9995	0.9994	1.0001	1.0001	0.9982	0.9984
Component Flow, kg-mol/h							
CH ₄	0.000	0.000	0.000	0.042	0.042	0.042	3.616
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.000	0.000	0.000	0.411	0.411	0.411	0.411
CO ₂	138.567	138.567	138.567	8.434	8.434	8.434	8.508
H ₂	0.000	0.000	0.000	0.657	0.657	0.657	0.657
H ₂ O	646.523	646.523	646.523	9.011	9.011	19.154	19.154
N ₂	1874.294	1874.294	1874.294	0.042	0.042	0.042	0.116
O ₂	239.327	239.327	239.327	0.000	0.000	0.000	0.000
TOTAL, kg-mol/h	2898.710	2898.710	2898.710	18.597	18.597	28.740	32.463

TABLE 25
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FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR

Stream Description	Pre-reformer Feed Stream	Reformer Feed Stream	Reformer Exit	Reformer Process Duty	Anode Feed	MCFC Stack Energy Recovery	Anode Exhaust
Temperature K	923	801	1061		950		978
Pressure kPa	603	603	603		599		598
Molar Flow kgmole/h	32.46	34.37	39.46		39.46		51.57
Mass Flow kg/h	793.6	793.6	793.6		793.6		1520.2
Enthalpy kJ/h	1.145E+06	9.920E+05	1.463E+06	1.026E+06	1.287E+06	1.663E+06	1.942E+06
Density kg/m3	1.9226	2.0926	1.373		1.5251		2.1678
Mole Wt.	24.4467	23.0915	20.11		20.11		29.4773
Spec. Heat kJ/kg-C	1.9072	1.8558	2.0174		1.9664		1.5706
Therm Cond W/m-K	0.0842	0.0844	0.1289		0.1179		0.0796
Viscosity cP	0.032	0.0285	0.0339		0.0313		0.0366
Z Factor	0.9995	0.999	1.0006		1.0004		1.0001
Component Flow, kg-mol/h							
CH4	3.616	2.663	0.116		0.116		0.116
C2H6							
C3H8							
n-C4H10							
CO	0.411	0.579	4.614		4.614		1.139
CO2	8.508	9.293	7.805		7.805		23.388
H2	0.657	4.300	10.456		10.456		1.823
H2O	19.154	17.417	16.356		16.356		24.989
N2	0.116	0.116	0.116		0.116		0.116
O2	0.000	0.000	0.000		0.000		0.000
TOTAL, kg-mol/h	32.463	34.368	39.464		39.464		51.571

TABLE 25
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FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR

Stream Description	Anode Exhaust to Combustor	Cathode Recycle Gas	Cathode Recycle Gas	Cathode Recycle Bypass	Cathode Recycle to Combustor	Combined Feed to Combustor	Combustor Effluent
Temperature K	978	978	978	978	978	975	1072
Pressure kPa	598	603	603	603	603	598	597
Molar Flow kgmole/h	32.97	255.38	255.38	140.68	114.70	148.66	147.72
Mass Flow kg/h	972.0	6863.9	6863.9	3781.1	3082.8	4083.3	4083.3
Enthalpy kJ/h	1.242E+06	8.004E+06	8.004E+06	4.409E+06	3.595E+06	4.852E+06	5.394E+06
Density kg/m3	2.1678	1.9902	1.9902	1.9902	1.9902	2.0233	1.8498
Mole Wt.	29.4773	26.8769	26.877	26.8769	26.8769	27.4664	27.6425
Spec. Heat kJ/kg-C	1.5706	1.3276	1.3276	1.3276	1.3276	1.3833	1.4113
Therm Cond W/m-K	0.0796	0.0692	0.0692	0.0692	0.0692	0.0713	0.076
Viscosity cP	0.0366	0.039	0.039	0.039	0.039	0.0378	0.0405
Z Factor	1.0001	1.0012	1.0012	1.0012	1.0012	1.001	1.001
Component Flow, kg-mol/h							
CH4	0.074	0.000	0.000	0.000	0.000	0.074	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.729	0.000	0.000	0.000	0.000	0.729	0.000
CO2	14.954	12.208	12.208	6.725	5.483	20.440	21.243
H2	1.165	0.000	0.000	0.000	0.000	1.165	0.000
H2O	15.978	56.961	56.960	31.378	25.583	41.571	42.885
N2	0.074	165.129	165.129	90.963	74.165	75.010	75.010
O2	0.000	21.085	21.085	11.615	9.470	9.674	8.579
TOTAL, kg-mol/h	32.975	255.383	255.383	140.681	114.702	148.664	147.717

TABLE 25
(Page 9 of 9)

FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 15°C AMBIENT AIR

Stream Description	Reformer Flue Gas	Power Module Gas to Blower	Recycle Blower Suction	Recycle Blower Discharge	Recycle Blower GHP	Cathode Exhaust Gas	Cathode Exhaust to Expander
Temperature K	890	932	864	866		978	978
Pressure kPa	596	596	596	603		603	603
Molar Flow kgmole/h	147.72	288.40	4248.77	4248.77		335.90	80.52
Mass Flow kg/h	4083.3	7864.3	117055.4	117055.4		9028.0	2164.1
Enthalpy kJ/h	4.367E+06	8.776E+06	1.173E+08	1.177E+08	4.255E+05	1.053E+07	2.524E+06
Density kg/m3	2.2252	2.0974	2.2855	2.3017		1.9902	1.9902
Mole Wt.	27.6425	27.269	27.5504	27.5504		26.877	26.877
Spec. Heat kJ/kg-C	1.3543	1.3422	1.2832	1.2841		1.3276	1.3276
Therm Cond W/m-K	0.0648	0.067	0.0621	0.0623		0.0692	0.0692
Viscosity cP	0.0353	0.0369	0.0358	0.0358		0.039	0.039
Z Factor	1.0009	1.0011	1.0012	1.0012		1.0012	1.0012
Component Flow, kg-mol/h							
CH4	0.000	0.000	0.000	0.000		0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000	0.000	0.000		0.000	0.000
CO2	21.243	27.968	337.979	337.979		16.057	3.849
H2	0.000	0.000	0.000	0.000		0.000	0.000
H2O	42.885	74.262	899.028	899.028		74.919	17.959
N2	75.010	165.974	2606.316	2606.316		217.193	52.064
O2	8.579	20.194	405.444	405.444		27.733	6.648
TOTAL, kg-mol/h	147.717	288.398	4248.768	4248.768		335.902	80.520

TABLE 26
Stream Summary Information
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FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Natural Gas Supply	Discharge from Natural Gas Compressor	Natural Gas Compressor GHP	Natural Gas Fuel to Combustor	Natural Gas to Process Unit	Pre-heated Natural Gas	Natural Gas Pre-heater Duty
Temperature K	288	426		426	426	616	
Pressure kPa	200	749		749	749	715	
Molar Flow kgmole/h	134.04	134.04		0.00	134.04	134.04	
Mass Flow kg/h	2257.4	2257.4		0.0	2257.4	2257.4	
Enthalpy kJ/h	1.287E+06	1.990E+06	7.032E+05	0.000E+00	1.990E+06	3.203E+06	1.212E+06
Density kg/m3	1.4125	3.5809		3.5809	3.5809	2.3479	
Mole Wt.	16.8416	16.8416		16.8416	16.8416	16.8416	
Spec. Heat kJ/kg-C	2.1213	2.499		2.499	2.499	3.1371	
Therm Cond W/m-K	0.0317	0.053		0.053	0.053	0.0862	
Viscosity cP	0.0111	0.0154		0.0154	0.0154	0.0201	
Z Factor	0.9951	0.9962		0.9962	0.9962	1.0006	
Component Flow, kg-mol/h							
CH4	128.676	128.676		0.000	128.676	128.676	
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000		0.000	0.000	0.000	
CO2	2.681	2.681		0.000	2.681	2.681	
H2	0.000	0.000		0.000	0.000	0.000	
H2O	0.000	0.000		0.000	0.000	0.000	
N2	2.681	2.681		0.000	2.681	2.681	
O2	0.000	0.000		0.000	0.000	0.000	
TOTAL, kg-mol/h	134.038	134.038		0.000	134.038	134.038	

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FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Demineralized Feedwater	Pre-heated Feedwater	Demin. Pre-heater Duty	Steam to Deaerator	Boiler Feedwater	Deaerator Vent	BFW Pump Suction
Temperature K	289	358		536	378	378	378
Pressure kPa	276	241		896	119	119	148
Molar Flow kgmole/h	390.50	390.50		13.11	403.49	0.13	403.49
Mass Flow kg/h	7035.0	7035.0		236.2	7268.9	2.3	7268.9
Enthalpy kJ/h	-1.376E+07	-1.165E+07	2.114E+06	2.326E+05	-1.142E+07	1.571E+03	-1.142E+07
Density kg/m3	1014.4741	960.4306		3.7298	944.1871	0.6873	944.2013
Mole Wt.	18.0151	18.0151		18.0151	18.0151	18.0151	18.0151
Spec. Heat kJ/kg-C	4.3154	4.3654		2.0458	4.4086	1.9114	4.4085
Therm Cond W/m-K	0.5963	0.673		0.0404	0.6824	0.0248	0.6824
Viscosity cP	1.1197	0.3301		0.0186	0.2666	0.0122	0.2666
Z Factor	0.002	0.0015		0.9716	0.0007	0.9902	0.0009
Component Flow, kg-mol/h							
CH4	0.000	0.000		0.000	0.000	0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000		0.000	0.000	0.000	0.000
CO2	0.000	0.000		0.000	0.000	0.000	0.000
H2	0.000	0.000		0.000	0.000	0.000	0.000
H2O	390.504	390.504		13.111	403.489	0.126	403.489
N2	0.000	0.000		0.000	0.000	0.000	0.000
O2	0.000	0.000		0.000	0.000	0.000	0.000
TOTAL, kg-mol/h	390.504	390.504		13.111	403.489	0.126	403.489

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FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	BFW Pump Discharge	BFW Pump HP	BFW to Steam Drum	BFW Heater Duty	Boiler Duty	Boiler Blowdown	Boiler Steam
Temperature K	378		416			448	448
Pressure kPa	1448		1276			896	896
Molar Flow kgmole/h	403.49		403.49			0.38	403.11
Mass Flow kg/h	7268.9		7268.9			6.8	7262.1
Enthalpy kJ/h	-1.140E+07	1.429E+04	-1.015E+07	1.256E+06	1.600E+07	-8.494E+03	5.858E+06
Density kg/m3	944.624		910.4422			879.3838	4.5488
Mole Wt.	18.0151		18.0151			18.0151	18.0151
Spec. Heat kJ/kg-C	4.4058		4.5373			4.7092	2.0286
Therm Cond W/m-K	0.6825		0.6875			0.6788	0.0308
Viscosity cP	0.266		0.1901			0.1535	0.0148
Z Factor	0.0088		0.0073			0.0049	0.952
Component Flow, kg-mol/h							
CH4	0.000		0.000			0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000		0.000			0.000	0.000
CO2	0.000		0.000			0.000	0.000
H2	0.000		0.000			0.000	0.000
H2O	403.489		403.489			0.378	403.112
N2	0.000		0.000			0.000	0.000
O2	0.000		0.000			0.000	0.000
TOTAL, kg-mol/h	403.489		403.489			0.378	403.112

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FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Superheated Steam	Superheater Duty	Steam to Process Unit	Ambient Air	Air Compressor Discharge	Air Compressor GHP	Process Air to Combustor
Temperature K	536		536	322	577		577
Pressure kPa	896		896	101	621		621
Molar Flow kgmole/h	403.11		390.00	2709.78	2709.78		0.99
Mass Flow kg/h	7262.1		7025.9	78299.5	78299.5		28.5
Enthalpy kJ/h	7.151E+06	1.292E+06	6.918E+06	2.536E+07	4.591E+07	2.055E+07	1.674E+04
Density kg/m ³	3.7298		3.7298	1.0867	3.7303		3.7303
Mole Wt.	18.0151		18.0151	28.8951	28.8951		28.8951
Spec. Heat kJ/kg-C	2.0458		2.0458	1.0118	1.0557		1.0557
Therm Cond W/m-K	0.0404		0.0404	0.0275	0.0434		0.0434
Viscosity cP	0.0186		0.0186	0.0199	0.0304		0.0304
Z Factor	0.9716		0.9716	0.9997	1.0019		1.0019
Component Flow, kg-mol/h							
CH ₄	0.000		0.000	0.000	0.000		0.000
C ₂ H ₆							
C ₃ H ₈							
n-C ₄ H ₁₀							
CO	0.000		0.000	0.000	0.000		0.000
CO ₂	0.000		0.000	8.212	8.212		0.003
H ₂	0.000		0.000	0.000	0.000		0.000
H ₂ O	403.112		390.000	0.000	0.000		0.000
N ₂	0.000		0.000	2134.981	2134.981		0.778
O ₂	0.000		0.000	566.591	566.591		0.207
TOTAL, kg-mol/h	403.112		390.000	2709.784	2709.784		0.988

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FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Process Air to Recycle Blower	Turbine Combustor Feed	Turbine Inlet	Turbine Exhaust	Turbine GHP	Flue Gas Exit N.G. Pre- heater	Flue Gas Exit Steam Superheater
Temperature K	577	978	978	667		656	644
Pressure kPa	621	603	594	105		104	103
Molar Flow kgmole/h	891.41	3233.82	3233.82	3233.82		3233.82	3233.82
Mass Flow kg/h	25757.3	87582.5	87582.5	87582.5		87582.5	87582.5
Enthalpy kJ/h	1.510E+07	1.008E+08	1.008E+08	6.656E+07	3.424E+07	6.535E+07	6.406E+07
Density kg/m3	3.7303	2.0054	1.9753	0.5114		0.5169	0.5232
Mole Wt.	28.8951	27.0833	27.0833	27.0833		27.0833	27.0833
Spec. Heat kJ/kg-C	1.0557	1.3077	1.3077	1.2134		1.2099	1.2062
Therm Cond. W/m-K	0.0434	0.0688	0.0688	0.049		0.0482	0.0474
Viscosity cP	0.0304	0.0396	0.0396	0.0297		0.0293	0.0289
Z Factor	1.0019	1.0013	1.0013	1.0001		1.0001	1.0001
Component Flow, kg-mol/h							
CH4	0.000	0.000	0.000	0.000		0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000	0.000	0.000		0.000	0.000
CO2	2.701	139.571	139.571	139.571		139.571	139.571
H2	0.000	0.000	0.000	0.000		0.000	0.000
H2O	0.000	647.351	647.351	647.351		647.351	647.351
N2	702.319	2137.663	2137.663	2137.663		2137.663	2137.663
O2	186.385	309.237	309.237	309.237		309.237	309.237
TOTAL, kg-mol/h	891.405	3233.820	3233.821	3233.821		3233.821	3233.821

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FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Flue Gas Exit Steam Boiler	Flue Gas Exit BFW Heater	Flue Gas Exit Demin Feedwater Heater	Anode Recycle Gas	Anode Recycle Gas	Ejector Exit Stream	Exchanger Feed Stream
Temperature K	490	477	456	978	978	839	807
Pressure kPa	103	102	101	598	598	607	607
Molar Flow kgmole/h	3233.82	3233.82	3233.82	20.04	20.04	30.87	34.59
Mass Flow kg/h	87582.5	87582.5	87582.5	585.5	585.5	780.7	843.4
Enthalpy kJ/h	4.806E+07	4.681E+07	4.469E+07	7.532E+05	7.532E+05	9.454E+05	1.034E+06
Density kg/m3	0.6838	0.6968	0.7238	2.149	2.149	2.2045	2.2079
Mole Wt.	27.0833	27.0833	27.0833	29.2217	29.2218	25.2889	24.3797
Spec. Heat kJ/kg-C	1.1627	1.1596	1.1545	1.5799	1.5799	1.69	1.8271
Therm Cond W/m-K	0.0371	0.0362	0.0348	0.0799	0.0799	0.07	0.0725
Viscosity cP	0.0238	0.0234	0.0226	0.0364	0.0364	0.029	0.0279
Z Factor	0.9997	0.9996	0.9995	1	1	0.9981	0.9984
Component Flow, kg-mol/h							
CH4	0.000	0.000	0.000	0.037	0.037	0.037	3.611
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000	0.000	0.433	0.433	0.433	0.433
CO2	139.571	139.571	139.571	8.901	8.901	8.901	8.975
H2	0.000	0.000	0.000	0.721	0.721	0.721	0.721
H2O	647.351	647.351	647.351	9.899	9.899	20.733	20.733
N2	2137.663	2137.663	2137.663	0.044	0.044	0.044	0.119
O2	309.237	309.237	309.237	0.000	0.000	0.000	0.000
TOTAL, kg-mol/h	3233.821	3233.821	3233.821	20.036	20.036	30.869	34.593

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FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Pre-reformer Feed Stream	Reformer Feed Stream	Reformer Exit	Reformer Process Duty	Anode Feed	MCFC Stack Energy Recovery	Anode Exhaust
Temperature K	925	802	1061		950		978
Pressure kPa	603	603	603		599		598
Molar Flow kgmole/h	34.59	36.62	41.62		41.62		53.72
Mass Flow kg/h	843.4	843.4	843.4		843.4		1569.9
Enthalpy kJ/h	1.220E+06	1.058E+06	1.548E+06	1.036E+06	1.362E+06	1.663E+06	2.020E+06
Density kg/m3	1.9128	2.0825	1.3836		1.537		2.149
Mole Wt.	24.3797	23.0292	20.2648		20.2648		29.2218
Spec. Heat kJ/kg-C	1.9041	1.8536	2.0112		1.9601		1.5799
Therm Cond W/m-K	0.0841	0.0844	0.1271		0.1162		0.0799
Viscosity cP	0.032	0.0286	0.034		0.0314		0.0364
Z Factor	0.9994	0.9989	1.0006		1.0003		1
Component Flow, kg-mol/h							
CH4	3.611	2.597	0.099		0.099		0.099
C2H6							
C3H8							
n-C4H10							
CO	0.433	0.611	4.584		4.584		1.162
CO2	8.975	9.812	8.337		8.337		23.867
H2	0.721	4.601	10.619		10.619		1.934
H2O	20.733	18.882	17.859		17.859		26.544
N2	0.119	0.119	0.119		0.119		0.119
O2	0.000	0.000	0.000		0.000		0.000
TOTAL, kg-mol/h	34.593	36.621	41.617		41.617		53.725

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FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Anode Exhaust to Combustor	Cathode Recycle Gas	Cathode Recycle Gas	Cathode Recycle Bypass	Cathode Recycle to Combustor	Combined Feed to Combustor	Combustor Effluent
Temperature K	978	978	978	978	978	975	1072
Pressure kPa	598	603	603	603	603	598	597
Molar Flow kgmole/h	33.69	247.37	247.37	131.43	115.95	150.62	149.65
Mass Flow kg/h	984.4	6699.7	6699.7	3559.5	3140.2	4153.2	4153.2
Enthalpy kJ/h	1.267E+06	7.711E+06	7.711E+06	4.097E+06	3.614E+06	4.897E+06	5.441E+06
Density kg/m3	2.149	2.0054	2.0054	2.0054	2.0054	2.0304	1.8571
Mole Wt.	29.2218	27.0833	27.0833	27.0833	27.0833	27.5735	27.7523
Spec. Heat kJ/kg-C	1.5799	1.3077	1.3077	1.3077	1.3077	1.3703	1.3976
Therm Cond W/m-K	0.0799	0.0688	0.0688	0.0688	0.0688	0.0711	0.0757
Viscosity cP	0.0364	0.0396	0.0396	0.0396	0.0396	0.0381	0.0409
Z Factor	1	1.0013	1.0013	1.0013	1.0013	1.0011	1.0011
Component Flow, kg-mol/h							
CH4	0.062	0.000	0.000	0.000	0.000	0.062	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.728	0.000	0.000	0.000	0.000	0.728	0.000
CO2	14.966	10.677	10.677	5.672	5.004	19.973	20.764
H2	1.213	0.000	0.000	0.000	0.000	1.213	0.000
H2O	16.645	49.520	49.520	26.309	23.210	39.855	41.192
N2	0.074	163.522	163.522	86.877	76.645	77.498	77.498
O2	0.000	23.655	23.655	12.568	11.088	11.294	10.199
TOTAL, kg-mol/h	33.689	247.374	247.374	131.427	115.947	150.624	149.654

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FULL-LOAD, END-OF-RUN OPERATION - SEA LEVEL / 49°C (0% REL. HUMIDITY) AMBIENT AIR

Stream Description	Reformer Flue Gas	Power Module Gas to Blower	Recycle Blower Suction	Recycle Blower Discharge	Recycle Blower GHP	Cathode Exhaust Gas	Cathode Exhaust to Expander
Temperature K	890	930	864	866		978	978
Pressure kPa	596	596	596	603		603	603
Molar Flow kgmole/h	149.65	281.08	4264.37	4264.37		337.20	89.83
Mass Flow kg/h	4153.2	7712.7	118309.6	118309.6		9132.5	2432.8
Enthalpy kJ/h	4.405E+06	8.502E+06	1.171E+08	1.176E+08	4.271E+05	1.051E+07	2.800E+06
Density kg/m3	2.2349	2.1149	2.3015	2.3177		2.0054	2.0054
Mole Wt.	27.7523	27.4395	27.7437	27.7437		27.0833	27.0833
Spec. Heat kJ/kg·C	1.3416	1.3267	1.2658	1.2667		1.3077	1.3077
Therm Cond W/m·K	0.0646	0.0666	0.0618	0.062		0.0688	0.0688
Viscosity cP	0.0355	0.0372	0.0362	0.0363		0.0396	0.0396
Z Factor	1.0009	1.0011	1.0012	1.0013		1.0013	1.0013
Component Flow, kg-mol/h							
CH4	0.000	0.000	0.000	0.000		0.000	0.000
C2H6							
C3H8							
n-C4H10							
CO	0.000	0.000	0.000	0.000		0.000	0.000
CO2	20.764	26.436	319.936	319.936		14.554	3.877
H2	0.000	0.000	0.000	0.000		0.000	0.000
H2O	41.192	67.502	810.020	810.020		67.502	17.982
N2	77.498	164.375	2674.823	2674.823		222.902	59.380
O2	10.199	22.767	459.589	459.589		32.245	8.590
TOTAL, kg-mol/h	149.654	281.080	4264.368	4264.368		337.202	89.828

3.B.4. Control Concept

3.B.4.a Fuel Cell

Fuel Cell Stack Temperature Control

The primary temperature control of the fuel cell stack is by adjustment of the balance of plant process controls based upon the outlet temperature of the stack's cathode effluent gas flow. These process control adjustments are the regulation of the amount of natural gas introduced to the plant's fuel cell system to maintain the reformer combustor outlet temperature; regulation of the flow of compressed air introduced upstream of the cathode inlet; and the total recycle gas flow rate through the cathode of the fuel cell stack.

Air from the turbine is mixed with the reformer combustor effluent and is supplied to the inlet of the cathode recycle blower. The cathode recycle blower outlet provides the required cooling gas flow to the cathode inlet of the stack. The total flow supplied to the cathode side of the stack is set to meet the excess air required based upon the natural gas flow rate and the cooling flow demanded for cooling as determined by the cathode effluent temperature.

The anode inlet temperature is controlled by varying the natural gas flow bypassed around the reformer feed/effluent heat exchanger. The anode inlet temperature is monitored to ascertain that the proper amount of anode recycle is used to maintain a safe margin above the critical Boudovard carbon formation temperature.

Stack Clamping

A stack clamping system is required to maintain the correct clamping force on the fuel cell stack. The clamping system consists of a set of springs initially set at the factory and periodically adjusted during plant outages.

Reformer/Fuel Cell Containment Vessel Purge Gas Control

Purge air is used to prevent accumulation of combustible gas mixtures in the reformer/fuel cell containment vessel. The purge air is designed to continuously sweep through the containment vessel to the reformer combustor. The vessel pressure is controlled to be lower than the process fuel streams for the reformer and fuel cell systems. The flow rate of the purge air is adjusted based upon vessel effluent temperature or O₂ and CO₂ concentrations.

Fuel Cell Stack Output Current Control

The inverter used in the plant design will control the current draw from the stack based upon operating stack voltage. The current draw control limits will be based upon the maximum and minimum design limits of the plant. This approach allows stack operation based upon the actual realized stack performance at the most efficient stack operating point. Compensation and adjustment of the plant controls for the slow decline in stack performance, or "decay rate", of the

fuel cell stack are based upon the fuel gas flow controls following the operating current of the stack over time and providing an automatic unmanned response. In addition, the heat provided from the stack due to internal losses to the balance of plant is held constant with this control concept for improvements in thermal integration and minimizing the required turndown ratios and inherent inefficiencies of the various pieces of balance of plant equipment.

3.B.4.b Turbine Generator

The information contained below is conceptual in nature and subject to change with further HEFPP design and development work. The Rolls- Royce Allison gas turbine generator will be controlled by a programmable logic control system. The gas turbine generator programmable logic control system (PLC) will be interconnected by Ethernet to the plant distributed control system (DCS) for monitoring purposes only. All safety related interlocks such as start/stop/trip and combustor management system will be hardwired between the PLC and the DCS to avoid the potential signal delays associated with digital networks.

Operating Modes

The gas turbine generator is capable of operating in either island mode or electric grid connected mode on the PLC system. The signal to change from one mode to another will come from the DCS via a hardwired signal based on the position of the plant-to-electric-grid circuit breaker.

The gas turbine generator controls will allow the turbine generator to operate in standalone mode where natural gas is in the combustor (normally in plant startup conditions) or in fuel cell support mode where the fuel cell exhaust is used to drive the gas turbine and no gas is used in the combustor (normal operation).

Compressor Control

Compressor discharge pressure and volumetric flow rate are controlled primarily by the speed of the compressor section when in the fuel cell support mode. Since the turbine, compressor and alternator are on the same shaft and the turbine section removes thermal energy from the fuel cell, the alternator load will be used as the primary control mechanism for speed control of the turbine generator. The alternator load will be controlled through the power conditioner. The signal to raise or lower the compressor flow or pressure will be sent to the PLC from the DCS. The PLC will convert this signal into a corresponding target load signal, which is sent onto the power conditioner to change the generator load. If due to process conditions, the compressor is close to the surge line and load cannot be changed to avoid the surge, a compressor emergency surge protection control valve will receive a signal to open from the PLC, which will relieve the air pressure to atmosphere. Compressor mass flow and pressure instrumentation signals are fed into the PLC to monitor the compressor surge potential.

A compressor bypass flow control valve is also provided, which connects the compressor and turbine sections. During all modes of operation, the compressor bypass flow control valve is under the supervision of the PLC control system which receives the same input signals from the compressor mass flow and pressure instrumentation mentioned above. In the standalone mode, the compressor bypass flow control valve will be fully open. In the fuel cell support mode, the compressor bypass flow control valve will be fully closed. In any mode, if the turbine generator PLC determines that the compressor is approaching surge with a pre-selected offset, then the compressor bypass flow control valve will modulate to avoid the surge so safe operation can be maintained.

Overspeed Protection

The primary mode of controlling the turbine generator speed is with the generator load. The generator load will be controlled by the PLC through the turbine generator power conditioning system. Signals to raise or lower the speed will come from the plant DCS and go to the PLC. To prevent turbine generator overspeed from fuel cell depressurization, a waste gate valve will be provided to bypasses the turbine section. The waste gate valve will fully open on detection of an overspeed condition. This valve is an open/close type of control valve.

Frequency and Voltage Control

When connected to the electric utility grid, the electric utility controls the plant AC power frequency of the plant. However, the turbine generator power conditioner will have an acceptable operating range of 59.5 to 60.5 Hz. When the frequency of the electric utility grid falls outside the operating range, the power conditioner will send a signal to the electric utility point of common coupling (typically a circuit breaker) to open and disconnect from the electric grid. Should utility disturbance cause the circuit breaker to open and then be restored to acceptable frequency limits quickly, the turbine generator power conditioner will wait for grid voltage and frequency to be within acceptable limits for 5 minutes prior to giving signal to close the breaker for reconnection to the grid. During the 5 minute wait period frequency and voltage will be adjusted to match the electric grid. When operating in island mode (i.e. disconnected from the utility), the turbine generator power conditioner will act as the master controller for frequency with the fuel cell inverters (as slaves) paralleling the turbine generator power conditioner frequency. The PLC will be preprogrammed to control the power conditioner for a 60 Hz frequency for islanded operation. Throughout the islanded operation period, the utility electric grid frequency will be monitored by the plant DCS and the PLC/power conditioner. When reconnection with the electric grid is desired, the operator will send a signal through the DCS to the PLC to synchronize back with the electric grid. The plant frequency will be adjusted by the turbine generator power conditioner to match the electric grid frequency for interconnection to the electric grid. When the frequencies are matched, the utility to plant circuit breaker will be closed.

The power conditioner will see the alternator input frequency vary up to 275 Hz with the variable speed gas turbine driver. The power conditioner will rectify the alternator input power to DC power and then invert the DC power back to usable AC power at the proper frequency and minimizing any harmonics.

When connected to the grid, the utility will determine the voltage the plant will see as the power conditioner would be considered a current source. The utility voltage will be matched with the power conditioner output voltage by the power conditioner controls and the available current would be delivered to the power plant auxiliaries with the excess going to the electric grid. If voltage falls outside the operating range of 3744 to 4576 Volts, the power conditioner will send a signal to the utility to plant circuit breaker to disconnect.

In the islanded operating mode, the power conditioner will maintain the plant voltage at a nominal 4160 Volts dependent on the auxiliary load of the plant. The acceptable operating output

voltage range of the power conditioner will be 3744 to 4576 Volts. If the voltage falls outside the operating range, the power conditioner will trip the plant.

Power Factor Control

The power conditioner will have the ability to change the power factor within the operating range of 0.85 lag to lead when output is 10 percent or greater. The power conditioner will control the phase angle of the output current with respect to the line voltage. The power conditioner may have an optional power factor control capability which allows the operator, via the DCS and PLC, the capability to set a constant power factor or constant reactive power delivery.

3.B.4.c Overall Plant

In terms of hierarchy, the control system's primary function is to support the operation of the stack: - delivery of fuel and oxidant streams and the removal of waste heat.

The HEFPP is a base-load facility. For base-load power demand from the grid, a Distributed Control System (DCS) based plant control system controls the load by maintaining the required average current density for each MCFC stack in the network and by controlling the speed and power output of the turbine generator. For a particular current draw, or MCFC fuel/oxidant conversion from each stack, a particular flow of natural gas is required to maintain the per-pass utilization, at the design anode recycle ratio. Therefore, the stack current draw automatically resets the flow of natural gas fuel to the power module. To avoid oxidation of the anode itself, there must always be sufficient supply of fuel at the three-phase interface when current is being drawn from the stack.

To maintain isothermal conditions in the MCFC stacks at different current draws, the quantity of cathode feed gas is varied in response to the temperature of the cathode exhaust gas. Cathode feed gas temperature is controlled (above its minimum value) by the injection of cooling air from the compressor into the recycle blower suction.

Steam flow to the fuel processor is set on flow control. The steam supply pressure is maintained by a trim steam condenser and a steam vent, in the event of unplanned shutdowns.

Anode recycle flow is set by the steam flow to the ejector.

The steam reformer outlet temperature is controlled by adjusting the flows of heating gas from the reformer combustor and cathode recycle; a bypass arrangement. The temperature of the heating gas is controlled by trim air injection to the combustor's mixed feed stream. Air is supplied from the compressor.

Air for stack cooling, for combustor temperature control, and for power module vessel purging are all set by demand signals from within the power module. If the total air flow drops below the compressor's minimum flow, then turbine control system can increase compressor throughput by several different mechanisms.

System operating pressure is normally maintained by flow regulation at the turbine inlet; through nozzle opening, or through a throttle valve.

3.B.5. List of Major Equipment

ITEM #	MAJOR EQUIPMENT ITEM	No. UNITS	UNIT CAPACITY	MATERIAL OF CONSTRUCTION	DESIGN CHARACTERISTICS (if available)
C-200	Steam Drum	1	Separation of 6,124 kg/h of sat'd steam from BFW. Operating conditions: 896 kPa, 449 K	Carbon Steel	Elevated - natural circulation Design Pressure: 1,069 kPa, Design Temperature: 456 K
C-201	Deaerator	1	5,933 kg/h of demin. water. Operating conditions: 119 kPa, 378 K	Carbon Steel	Steam-stripped deaeration to < 7ppb O ₂ Design Pressure: 310 kPa & Full Vacuum Design Temperature: 408 K
C-202	Zinc Oxide Desulfurizer	1	2,257 kg/h of natural gas. Operating conditions: 720 kPa, 616 K	Low-Alloy Steel	Sulfur reduction from 4 ppmv to < 0.1 ppmv 1.50 m ID x 2.4 m T/T Design Pressure: 900 kPa Design Temperature: 680 K
E-200A	Natural Gas Pre-heater	1	Duty = 1,212,480 kJ/h	Low-Alloy Steel	Heat exchange coil in HRSG duct
E-200B	Steam Superheater	1	Duty = 1,089,840 kJ/h	Low-Alloy Steel	Heat exchange coil in HRSG duct
E-200C	Steam Boiler	1	Duty = 13,489,022 kJ/h	Carbon Steel	Natural circulation boiler in HRSG duct
E-200D	Boiler Feedwater Heater	1	Duty = 1,058,980 kJ/h	Carbon Steel	Heat exchange coil in HRSG duct
E-200E	Demineralized Water Heater	1	Duty = 1,782,940 kJ/h	Stainless Steel	Heat exchange coil in HRSG duct
K-400	Natural Gas Compressor	2	3,004 Nm ³ /h gas flow 703,236 kJ/h GHP 214 kW Motor		100% units - 1 operating, 1 standby Compressor E _{ad} : 70% Suction conditions: 200 kPa, 288 K Pressure Ratio = 3.7
K-600	Gas Turbine/Generator Package	1	3,429 kW Generator Output		Compressor: Pressure Ratio 6.2, E _{ad} : 84%, Flow 60,507 kg/h, T _{suction} : 288 K, GHP 3,964 kW Turbine: Pressure Ratio 5.7, E _{ad} : 91.9%, Flow 68,688 kg/h, T _{inlet} : 978 K, GHP 7,561 kW Generator: Eff. 98.5%
P-600	Boiler Feedwater Pump	2	6.4 m ³ /h, 1,300 kPa differential head, 3.9 kW Motor	CS Casing, High-Alloy Impeller	100% units - 1 operating, 1 standby Multi-stage centrifugal

ITEM #	MAJOR EQUIPMENT ITEM	No. UNITS	UNIT CAPACITY	MATERIAL OF CONSTRUCTION	DESIGN CHARACTERISTICS (if available)
C-500	Power Module Vessel	9	3.1 m ID x 14.0 m T/T Operating conditions: 603 kPa, 550 K	Refractory-lined Carbon Steel	Horizontal cylindrical vessel, center flanged for transportation and access. Each vessel is pressurized and contains 4 (four) Power Modules (see Power-Module equipment list). Vessel is refractory lined and purged with air. Design Pressure: 720 kPa, Design Temperature: 615 K
K-500	Recycle Gas Blower	3	85,949 Nm ³ /h gas flow 384,120 kJ/h GHP 117 kW Motor	High-Alloy Steel	Axial Fan/Blower. Differential head: 6.2 kPa, E _{ad} : 75% Suction conditions: 596 kPa, 864 K

ITEM #	MAJOR EQUIPMENT ITEM	No. UNITS	UNIT CAPACITY	MATERIAL OF CONSTRUCTION	DESIGN CHARACTERISTICS (if available)
E-501	Reformer Feed/Effluent Exchanger	36	Duty = 160,564 kJ/h	High-Alloy Steel	Operating differential pressure: < 10 kPa (hot side to cold side, inside to outside) Design Pressure: 100 kPa, Design Temperature: 1070 K
K-501	Anode Recycle Gas Ejector	36	Steam: 164.6 kg/h (872 kPa, 536 K) Anode recycle: 493.7 kg/h (604 kPa, 978 K) Pressure ratio = 1.01 Weight ratio = 3.0	High-Alloy Steel	Operating differential pressure: 280 kPa (motive steam section) Design Pressure: 500 kPa, Design Temperature: 980 K
P-501	Power Conversion System - Inverter	9	2,180 amps, 800-900V DC, 1968 kW input		4160 VAC, 3 phase, 60 Hz output - 1918 kW Efficiency: 97.5%
R-501A	Pre-reformer	36	721 kg/h feed gas 24 % CH ₄ conversion	Stainless Steel	Adiabatic catalytic reactor - cylindrical packed bed
R-501B	Reformer	36	721 kg/h feed gas 73.6 % CH ₄ conversion Duty = 1,018,806 kJ/h	Stainless Steel	Horizontally-stacked, plate-type catalytic reactor Design Temperature:
R-501C	MCFC Stack	36	492 kW DC output		Molten-carbonate Fuel Cell 758 mV, 200 mA/cm ² 300 plates, 10,900 cm ² /plate
R-501D	Reformer Combustor	36	4,096 kg/h feed gas 1,072 K exit temperature	High-Alloy Steel	Catalytic Combustor - combustion of residual H ₂ , CO, and CH ₄ in anode exhaust gas Design Temperature:
	Stepup Transformer	1	25MVA OFA 4160 V input 138KV output	Carbon Steel	Depends on utility requirements

3.C Economic Projections

3.C.1. Description of Estimate

The three most common techniques used for plant cost estimates are: - curve estimates, factored estimates, and definitive estimates.

A curve estimate relies on a historical relationship between plant capacity and cost for a particular technology. There is no historical database available yet for commercial MCFC power plants. Therefore, a curve-estimated cost is not applicable.

A factored cost estimate relies on empirical cost data collected over time from actual projects. Cost relationships between different system costs (major equipment, bulk materials, direct labor, indirect costs, etc.) are characterized from analysis of the historical data. Factors are assigned to represent the relationship. Once a cost system, say the equipment costs, is well known, the remaining plant costs can be obtained by applying these factors to the known cost items.

Again a historical database, from which factors can be derived, is not available for commercial MCFC power plants.

A definitive estimate requires a significant amount of detailed engineering to be performed (15-25%), so that material takeoffs of bulk material quantities can be used to develop accurate costs. A definitive estimate includes known site location, firm design basis, firm equipment quotations, knowledge of local construction labor markets, and a detailed construction plan.

Unfortunately, this study does not provide sufficient time or budget to develop a definitive estimate. In addition, as mentioned above, there is no historical database available to perform a curve or factored estimate. Therefore, in this case, capital costs were developed from vendor quotations and in-house equipment estimates, and installation/construction costs were developed using costs developed for the 450 kW MCFC PDI³ unit and applying best engineering judgement.

MC-Power and Rolls-Royce Allison provided costs for the MCFC stacks and the gas turbine, respectively.

The plant construction plan is to maximize shop assembly of equipment, piping and instrumentation as packages or modules, so as to minimize construction time in the field. Module sizes are constrained by permissible shipping dimensions and weights.

Per the design basis, the capital cost estimate is based on mature technologies and integration techniques, not first-of-a-kind systems. Therefore, the assumption is that initial R&D and engineering expenditures have already been recouped by equipment vendors, and that sales price is based on competition in a mature-technology market.

³ Process Design Improvement program – MC-Power/DOE Contract DE-FC21-95MC30133

3.C.2. Sources of Cost Data

Capital cost data for major equipment items were provided from vendor quotes. The fuel processor system costs were provided by IHI. The MCFC stack costs were provided by M-C Power. The gas-turbine generator costs were provided by Rolls-Royce Allison. The inverter costs were provided by Trace Technologies. The recycle blower costs were provided by Robinson Industries, Inc. The HRSG unit costs were provided by the G.C. Broach Company. Other costs were estimated from previous recent projects (with similar equipment) and the 1998 Richardson Process Plant Construction Estimating Standards.

3.C.3. Estimate of Capital and Operating Costs

Capital Costs

The capital cost estimate is based on the costs for new equipment, such as the fuel cell stacks and the plate reformers, being projected to the cost for the final stages of its evolutionary development, from both a technical standpoint and from a production/assembly standpoint. In addition, it is assumed that the HEFPP has been demonstrated to be commercially competitive (economically and environmentally) to the extent that a significant number of 20 MW units are installed each year.

Equipment vendors provided capital costs for equipment in response to preliminary equipment specification inquiries. The major equipment cost list appears in Table 12. Equipment quotes for the gas turbine generator, HRSG unit and solid-state inverters, etc. were included as received, but modified to include a 10% cost reduction for volume production, on the assumption that twenty, 20 MW HEFPP plants would be installed domestically, annually. Equipment costs for technologies continuing to undergo evolutionary development, such as the M-C Power MCFC and IHI Fuel Processor, were assumed to follow the standard cost decline curve from “high cost” for initial production, decaying to a “low-cost” plateau for future mass production. MCFC stack and fuel processor mass-production capital costs are projected to be 300 \$/kW and 145 \$/kW, respectively.

Sales tax on equipment purchases and transportation costs for equipment from the module assembly yard to site are assumed to be 6% and 5% of the equipment capital costs, respectively.

Bulk materials (piping, electrical, instrumentation, insulation, etc.) and labor (for equipment and bulk materials installation), subcontracts, and field indirect costs are accounted for under “Installation Costs”. Installation costs for the HEFPP were estimated from detailed module fabrication/assembly costs developed for a single 450 kW module, as part of the PDI program and on typical larger conventional power plant installation costs.

Since it is intended to compare the costs and COE for the HEFPP to those for conventional, commercially-proven technologies, such as combined-cycle gas turbine power generation plants, a process equipment contingency has been applied to those equipment in the HEFPP which are in the development cycle, and whose costs were projected to mature-equipment status – it being quite possible that unforeseen factors will increase the actual mature-equipment costs.

In addition, 15% of the cost for the process plant and general plant facilities has been applied to the total capital requirement for the HEFPP (and the conventional power generation plants) as a project contingency, to address the likelihood of other project-related cost overruns.

The sum of the major equipment installed costs is the Process Plant Cost (PPC).

General plant facilities accounts for miscellaneous materials, equipment and services needed to support commercial operation of the plant, e.g. fire protection systems, in-plant roads, boundary

fences, lighting, etc. The cost of these items is estimated to be 5% of PPC for the HEFPP and the gas turbine combined cycle option.

Home office engineering fees and service costs are estimated to be 7.5% of PPC with the high degree of plant modularization in both the HEFPP and the combined cycle plants.

Startup costs are assumed to consist of one month of operating labor, one month of maintenance materials, one month of consumables, one week of natural gas fuel, and one half percent of PPC for both types of plants.

Spare parts were assumed to be ½ percent of the process plant cost for both HEFPP and combined cycle plants.

Working capital is assumed to consist of three months of operating labor and two months of consumables (excluding natural gas) with a twenty-five percent contingency. Three months of maintenance materials are also included for both plants.

The total capital requirements for the HEFPP are shown in Table 13.

Operating Cost

Operating costs for the HEFPP are based on a 92% annual capacity factor and fifty weeks operation per year.

The capacity factor is an estimate, based on similar plant histories, of the percentage of name-plate capacity that a plant will actually achieve during operation. This figure is often less than 100%, due to the plant experiencing unanticipated reductions in power output, and/or unplanned shutdowns.

Two weeks per year is typical of the time period required to do annual plant maintenance.

For the present analysis, the cost of natural gas is fixed at \$3.00 MMBtu (HHV).

The HEFPP plant organization chart is shown in Table 11. Operators will work in 12 hour shifts. O&M supervisor and water treatment specialist will fill in on vacation/sick leave of operators. The operations and maintenance would be similar to those expected for a combined cycle power plant of similar size on a single power plant site. The combined cycle organization was based on September, 1994 Power Magazine article "Operating and Maintaining IPP Cogen Facilities (pages 13-32).

Maintenance material costs are assumed to be 1½% of PPC and are primarily consumables. This is a historically based value, and is typical for plants requiring light service demands.

Insurance and local taxes are assumed to be 2% of PPC. In less populated areas, local property taxes are between 1-2% of the direct-capital investment. Annual insurance rates typically amount to 1% of the direct-capital investment.

Catalysts and chemicals cost include the cost of boiler water treatment chemicals and reforming catalyst replacement. Since the reforming catalyst operates under mild conditions for steam reforming catalysts, its operating life is anticipated to be approximately ten years.

The plant annual operating costs for the HEFPP are shown in Table 14.

COE Analysis

The cost of electricity (COE) is determined by NPV analysis. The COE is adjusted until the sum of the present-value cash flows, at a 15% discount rate, over the life of the plant is equal to zero.

The following basis (in addition to items mentioned above) was used for the COE analysis:

- Plant design lifetime 25 yrs
- Capital charge rate 15%
- Inflation allowance none – constant dollars
- Additional capital expenditure MCFC stack replacement every 5 yrs. Stack salvage value is assumed to be 33% of the stack equipment cost
- Depreciation Double Declining Balance for 10 yrs, straight line remaining life
- State & Federal Taxes 43%

Plant revenues are based on power exports for SOR and EOR assuming straight-line decline in stack performance.

The total project capital requirement is assumed to be expended over a two-year period, with one-third of the expenditure occurring in the first year. The first year has permitting and engineering, while the second year has startup and initial operation.

Project cash flows as part of the economic evaluation are shown in Table 15.

TABLE 27
HEFPP
Plant Organization Chart

Salaries (1998 Basis)

Position	Quantity	Unit Annual Salary (\$)	Total Annual Salary (\$)
Plant Manager	1	80,000	80,000
O & M Supervisor	1	65,000	65,000
Administrative Assistant	1	25,000	25,000
Water Treatment Specialist	1	30,000	30,000
Mechanics	2	30,000	60,000
I/E Technicians	2	40,000	80,000
Control Room Operator	4	40,000	160,000
Roving Operator	4	35,000	140,000
TOTAL	16		640,000

Table 28
HEFPP
Major Equipment Cost List

ITEM #	MAJOR EQUIPMENT ITEM	EQUIPMENT COST/UNIT	SALES TAX/ UNIT	FREIGHT COST/UNIT	INSTALLATION COST/UNIT	TOTAL COST/UNIT	QUANTITY	TOTAL COST
	HRSG unit/deaerator	\$202,140	\$12,128	\$10,107	\$60,642	\$285,017	1	\$285,017
C-200	Steam Drum							
E-200A	E-200A Natural Gas Pre-heater							
E-200B	Steam Superheater							
E-200C	Boiler							
E-200D	Boiler Feedwater Heater							
E-200E	Demineralized Water Heater							
C-201	Deaerator							
C-202	Zinc Oxide Desulfurizer	\$11,179	\$671	\$559	\$27,948	\$40,356	1	\$40,356
K-400	Natural Gas Compressor	\$50,000	\$3,000	\$2,500	\$100,000	\$155,500	2	\$311,000
C-500	Power Module Vessel	\$84,667	\$5,080	\$4,233	\$127,001	\$220,981	9	\$1,988,828
K-500	Recycle Blower	\$108,000	\$6,480	\$5,400	\$162,000	\$281,880	3	\$845,640
K-501	Anode Recycle Gas Ejector	\$3,600	\$216	\$180	\$360	\$4,356	36	\$156,816
P-501	Power Conversion System - Inverter	\$180,000	\$10,800	\$9,000	\$27,000	\$226,800	9	\$2,041,200
	Fuel Processor (w/catalyst)	\$72,500	\$4,350	\$3,625	\$29,000	\$109,475	36	\$3,941,100
E-501	Reformer Feed/Effluent Exchanger							
R-501A	Pre-reformer							
R-501B	Reformer							
R-501D	Reformer Combustor							
R-501C	MCFC Stack	\$150,000	\$9,000	\$7,500	\$45,000	\$211,500	36	\$7,614,000
K-600	Gas Turbine-Generator Package	\$724,000	\$43,440	\$36,200	\$99,188	\$905,000	1	\$905,000
P-600	Boiler Feedwater Pump	\$2,500	\$150	\$125	\$7,500	\$10,275	2	\$20,550
	Plant Control System	\$650,000	\$39,000	\$32,500	\$65,000	\$786,500	1	\$786,500
	Electrical Switchgear	\$187,000	\$11,220	\$9,350	\$37,400	\$244,970	1	\$244,970
	Stepup Transformer	\$187,500	\$11,250	\$9,375	\$37,500	\$245,625	1	\$245,625
								\$19,426,602

Table 29
Electric Power Generation Cost

AREA No.	PLANT SECTION DESCRIPTION	PROCESS CONT., %	COST, K\$ w/o CONT.
	HRSG Unit/Deaerator		285
	Zinc Oxide Desulfurizer		40
	Natural Gas Compressor		311
	Power Module Vessel	5	1,989
	Recycle Blower	10	846
	Anode Recycle Gas Ejector		157
	Inverter	10	2,041
	Fuel Processor	10	3,941
	MCFC Stack	10	7,614
	Gas Turbine/Generator Package		905
	BFW Pump		21
	Plant Control System		787
	Electrical Switchgear		245
	Stepup Transformer		246
	Subtotal, Process Plant Cost		19,427
	General Plant Facilities		971
	Engineering Fees		1,457
	Process Contingency (Using conntingencies listed above)		1,544
	Project Contingency, 15 % Proc. Plt. & Gen. Plt. Fac.		3,060
	Total Plant Cost		26,458
	Prepaid Royalties		
	Initial Catalyst and Chemical Inventory		15
	Startup Costs		334
	Spare Parts		97
	Working Capital		301
	Land, <u> 1 </u> Acres		25
	Total Capital Requirement (TCR)		27,230

Table 30
Annual Operating Costs

COST ITEM		QUANTITY		UNIT \$ PRICE		ANNUAL COST, K\$
Natural Gas	Fuel Type	<u>2609.3</u>	MMBtu/D	<u>\$ 3.00</u>	/MMBtu	\$ 2,527.46
Consumable Materials						
	Demineralized Water	<u>37.67</u>	Mgal/D	<u>\$ 3.00</u>	/Unit	\$ 36.49
	Catalysts & Chemicals	_____	Unit/D	_____	/Unit	\$ 100.00
	Fuel Cell Stack Replacement	_____	Unit/D	_____	/Unit	
		_____	Unit/D	_____	/Unit	
	Disposal Costs (If any)	_____	Unit/D	_____	/Unit	
Plant Labor						
	Operating Labor (incl. benef.)	<u>See Plant</u>	Organization Chart	_____		\$ 640
	Supervision & Clerical					
	Maintenance Costs					\$ 291
	Insurance & Local Taxes					\$ 389
	Royalties					
	Other Operating Costs					
Total Operating Costs						\$ 3,983.88
By-Product Credits						
_____		_____	Unit/D	_____	/Unit	
_____		_____	Unit/D	_____	/Unit	
_____		_____	Unit/D	_____	/Unit	
_____		_____	Unit/D	_____	/Unit	
_____		_____	Unit/D	_____	/Unit	
Total By-Product Credits						
Net Operating Costs						\$ 3,983.88

TABLE 31

COE - \$/kW-h, NPV Calculation																
CASE: 20 MW MCFC-Based HEFPP Facility																
TOTAL PROJECT COST 27.23 \$MM																
End of Proj. Year	End of Opn. Year	Investment-related Cash Flows \$1,000			Before Tax Operating Cash Flows \$1,000			Tax Calculations \$1,000					Cash Flows	NPV		
		Total Plant Cost (a)	Working Capital (b)	Total	Sales Revenue (c)	Production Costs (d)	Net Opn. Cash Flows	Depreciation (DDB) (e)	Tax Allowance Earned in Previous Year	Taxable Profit in Previous Year	Tax Payable (f)	After Tax Operating Cash Flow	Project Net Cash Flow	Discounted at IRR of 15.00%		
1		(8,819)	(152)	(8,971)									(8,971)	(8,971)		
2		(17,639)	(620)	(18,259)									(18,259)	(15,877)		
3	1	0	0	0	9,758	(3,984)	5,774					5,774	5,774	4,366		
4	2	0	0	0	9,666	(3,984)	5,682	8,172	8,172	(2,397)	1,031	6,712	6,712	4,414		
5	3	0	0	0	9,573	(3,984)	5,589	6,537	6,537	(856)	368	5,957	5,957	3,406		
6	4	0	0	0	9,485	(3,984)	5,501	5,230	5,230	359	(154)	5,347	5,347	2,658		
7	5	0	0	0	9,387	(3,984)	5,404	4,184	4,184	1,317	(566)	4,837	4,837	2,091		
8	6	(3,600)	0	(3,600)	9,758	(3,984)	5,774	3,347	3,347	2,056	(884)	4,890	1,290	485		
9	7	0	0	0	9,666	(3,984)	5,682	2,678	2,678	3,097	(1,332)	4,350	4,350	1,422		
10	8	0	0	0	9,573	(3,984)	5,589	2,142	2,142	3,540	(1,522)	4,067	4,067	1,156		
11	9	0	0	0	9,485	(3,984)	5,501	1,714	1,714	3,875	(1,666)	3,835	3,835	948		
12	10	0	0	0	9,387	(3,984)	5,404	1,371	1,371	4,130	(1,776)	3,628	3,628	780		
13	11	(3,600)	0	(3,600)	9,758	(3,984)	5,774	1,097	1,097	4,307	(1,852)	3,922	322	60		
14	12	0	0	0	9,666	(3,984)	5,682	313	313	5,461	(2,348)	3,333	3,333	542		
15	13	0	0	0	9,573	(3,984)	5,589	313	313	5,368	(2,308)	3,281	3,281	464		
16	14	0	0	0	9,485	(3,984)	5,501	313	313	5,276	(2,269)	3,233	3,233	397		
17	15	0	0	0	9,387	(3,984)	5,404	313	313	5,188	(2,231)	3,173	3,173	339		
18	16	(3,600)	0	(3,600)	9,758	(3,984)	5,774	313	313	5,090	(2,189)	3,586	(14)	(1)		
19	17	0	0	0	9,666	(3,984)	5,682	313	313	5,461	(2,348)	3,333	3,333	269		
20	18	0	0	0	9,573	(3,984)	5,589	313	313	5,368	(2,308)	3,281	3,281	231		
21	19	0	0	0	9,485	(3,984)	5,501	313	313	5,276	(2,269)	3,233	3,233	198		
22	20	0	0	0	9,387	(3,984)	5,404	313	313	5,188	(2,231)	3,173	3,173	169		
23	21	(3,600)	0	(3,600)	9,758	(3,984)	5,774	313	313	5,090	(2,189)	3,586	(14)	(1)		
24	22	0	0	0	9,666	(3,984)	5,682	313	313	5,461	(2,348)	3,333	3,333	134		
25	23	0	0	0	9,573	(3,984)	5,589	313	313	5,368	(2,308)	3,281	3,281	115		
26	24	0	0	0	9,485	(3,984)	5,501	313	313	5,276	(2,269)	3,233	3,233	98		
27	25	0	1,000	1,000	9,387	(3,984)	5,404	313	313	5,188	(2,231)	3,173	4,173	110		
28	26															
29	27															
		(40,858)			40,858							NPV =		\$0.00		
(*)		Fuel Cell Stack replacement every 5 years, with credit taken for salvage @ 33% of stack cost								Operating Days/Year		351				
(a)		1998 Capital Investment of..... 27.23 \$MM								SOR Power Output, MW		20.1 @ cap fac.		0.92		
(b)		Recovered at end of project								Elect Value, \$/kW h		0.063				
(c)		Electricity Value x Power Output (@cap fac.) x Opn. Days/year								EOR Power Output, MW		19.2				
(d)		From Operating Cost Table														
(e)		Double-declining-balance depreciation for 10 years, followed by straight-line depreciation for remaining project life. Zero salvage value														
(f)		Combined State and Federal Tax rate of 43%														

3.C.4. Economic Analysis Results

The COE for a mature technology 20 MW MCFC-based power plant (for the above design basis and assumptions) is 6.3 cents/kW-hr. The total project capital requirement for the turnkey facility is \$1362/kW.

In performing the economic analysis for the MCFC-based HEFPP, no cash-flow credits were assigned for the reduced emissions of air pollutants and greenhouse gases characteristics of this type of facility.

3.C.5. Comparison of Fuel Cell Power Plant with Competing Technology

This section compares the MCFC-based HEFPP power plant to similar sized, conventional, fossil-fuel power generation systems. The power generation system selected includes a gas turbine based combined-cycle gas turbine producing 22 MW (16 MW GT, 6 MW steam turbine) gross and 20.2 MW net. Each system is assumed to operate as a base-load unit.

The power plant performance, costs, and COE for a 15% capital charge rate, for each of these units are compared below.

1. Power Plant Efficiency

This performance criteria refers to the efficiency at which each of these power generation systems converts chemical energy in the fuel to net electric power, i.e. the quantity of power available for export to the grid once auxiliary loads have been subtracted.

Combined Cycle Gas Turbine performance figures were obtained from 1998 Gas Turbine World Handbook.

The Table below summarizes the efficiency for each plant.

Plant Efficiency		
	CC-GT	HEFPP
Efficiency (LHV),%	43.3	70.1
kJ/kW-h	8,314	5,134
kcal/kW-h	1,986	1,227
Btu/kW-h	7,882	4,868
Nat. Gas, Std.* M ³ /kW-h	0.242	0.149

*0°C, 101.3 kPa

Clearly, the MCFC-based HEFPP is significantly more efficient at converting chemical energy in fossil-fuel reserves to electric power.

Note: No attempt has been made to increase plant efficiency by including additional heat recovery to low-grade, closed-loop heating cycles, or refrigeration cycles.

2. Plant Emissions

Known pollutants associated with fossil-fuel power plants include sulfur oxides (SO_x), nitrogen oxides (NO_x), carbon monoxide, unburned hydrocarbons, and particulate matter.

NO_x

To meet the prevailing emission requirements, gas turbine combined cycle plants must include either low NO_x burners and/or post-combustion NO_x reduction technology. Dry, low-NO_x burners typically achieve NO_x emissions in the range of ppmv. ABB was contacted and the GT-

35 used in the study presently has a NO_x value of 45 ppm. The 25 ppm value will not be available until 2001. The study however has assumed that the 25 ppm NO_x machine is available for purposes of the study. This is roughly equivalent to 2,500 metric tonnes of NO_x emissions over the 25-year life of the 22 MW CC-GT plant. Further reductions in NO_x emissions can be achieved by including best available control technology (BACT) such as a selective catalytic reduction unit (SCR), or catalytic control technology like SCONO_xTM, which has recently pushed the lowest achievable emissions rate for NO_x to 2-3 ppmv. Emission of 3 ppmv NO_x over the plant life is equivalent to emitting 320 metric tonnes of NO_x. Typical performance for SCRs is in the region of 9 ppmv NO_x. By using the SCR technology, unreacted ammonia release into the atmosphere requires monitoring in the exhaust stack. Other pollutants such as NO_x, CO, and unburned hydrocarbons must also be monitored by a continuous emissions monitoring system. This system requires a significant amount of maintenance and utilizes a significant amount of calibration gases for proper operation.

NO_x production in the HEFPP is expected to be extremely low. South Coast Air Quality Management District (SCAQMD)(located in the Los Angeles area) tests for Phosphoric Acid Fuel Cell (PAFC) atmospheric emissions, measured NO_x emissions as <1 ppmv. MCFCs are expected to achieve even lower NO_x emission levels, since less combustion is needed to provide the heat for the fuel processing step. In addition, since all of the gas from the fuel processor's catalytic combustor passes through the MCFC cathode compartment, which is an excellent "NO_x scrubber"⁴, the emission of NO_x from the stack should be negligible. NO_x measurements taken by IHI on their MCFC stack were undetectable on equipment with a 1 ppmv lower detectable limit.

At 1 ppmv NO_x emission, the quantity of NO_x emitted over the 25-year life of the HEFPP would be 23 metric tonnes.

It is important to emphasize that combustion turbine based power plants, where the turbine combustor continuously fires fuel to sustain the cycle, cannot yet achieve the NO_x emission performance of a MCFC-based power plant.

The Table below illustrates the estimated NO_x emissions from each plant.

NO_x Emissions

	CC-GT	HEFPP
NO _x Emission (low NO _x bnr. + SCR)	3 ppmv	<1 ppm
NO _x emitted over life of plant, MT	316	<23

SO_x

The amount of SO_x generated depends on the quantity of fuel combusted and its sulfur content. Since the MCFC-based HEFPP 1) desulfurizes its fuel feedstock, 2) has a higher natural-gas-conversion efficiency, 3) will capture desulfurization trace sulfur (<0.1 ppmv) in the pre-reforming/reforming section, and 4) does not fire the gas turbine combustor during normal

⁴ A.J. Appelby, Characteristics of Fuel Cell Systems.

operation, the amount of sulfur emitted from the plant can be considered negligible, or insignificant.

The Table below illustrates the estimated SO_x emissions from each plant.

SO_x Emissions		
	CC-GT	HEFPP
SO _x emitted over life of plant, MT	12.2	<0.2

CO₂

There is increasing support being given to the viewpoint that the continued emission of so-called greenhouse gases is causing changes to the global climate. Human-based activities leading to the emission of carbon dioxide, methane, chlor-fluorocarbons, and other gases are believed to be the main contributors to the climate change.⁵ At the Kyoto conference, Annex 1 countries agreed to an average reduction of 5.2% in greenhouse gas emissions by the period 2008-2012.

By virtue of its superior fuel conversion efficiency, the MCFC-based HEFPP has one of the lowest CO₂ emissions rating of any commercial, or near-commercial, fossil-fuelled power plants.

The Table below illustrates the estimated CO₂ emissions from each plant.

CO₂ Emissions		
	CC-GT	HEFPP
CO ₂ emitted over life of plant, MT	2.1x10 ⁶	1.1x10 ⁶

As summarized in the three tables above, the MCFC-based HEFPP emits significantly less air pollutants and greenhouse gases than the competing fossil-fuel power generation systems.

Power Plant Economic Analysis

1. Cost of Electricity

The cost of producing electric power for each system is summarized below. The basis for comparison is:-natural gas price \$3.00/MMBtu (HHV), capital charge rate 15%, capacity factor 92%, and plant life 25-years.

The COE comparison is summarized below, while detailed capital and operating costs and the NPV analyses for the gas turbine combined cycle facilities are presented in Tables 32, 33, and 34.

The costs for the gas turbine combined cycle were obtained from recent power plant publications. Costs for SCR NO_x control systems, plant control systems, and continuous

⁵ K. Thambimuthu and P. Freund IEA Greenhouse Gas R&D Programme.

emissions monitoring systems were added to the gas turbine simple cycle and combined cycle projects. Project line-item costs below the PPC (engineering fees, project contingency, working capital, spare parts, maintenance, local taxes and insurance, etc.) were applied at the same rates as used for the HEFPP capital and operating cost estimates. No process equipment contingencies were applied to the gas turbine or steam turbine cycles – although a 10% contingency was applied to the estimated cost for the SCR. Based on discussions with ABB, the entire plant would have a major overhaul covering the gas turbine generator, steam turbine generator, HRSG, cooling tower, condenser and other balance of plant equipment every five years. ABB recommended that the five year overhaul costs should be based on \$2.00/MWh of operation. This major overhaul would last four weeks and would impact the sales revenues for the overhaul years. It should be noted that the gas turbine will have to come off line for ten hours for water washing the compressor section every two months. Power plant output also degrades approximately one half percent per year between plant major overhauls due to gas turbine and steam turbine performance degradation. Catalyst and chemical costs will be significantly higher for the combined cycle plant due to plant lube oils, cooling tower water treatment chemicals, CEM calibration gases, gas turbine washing fluids and SCR ammonia.

COE Comparison

	CC-GT	HEFPP**
Total Plant Investment, \$MM	23.2	27.2
	\$/kW	
Annual Operating Costs, \$MM	1161	1362
(Based on \$3.00/MMBtu gas)	6.13	3.98
COE* for \$1.50/MMBtu gas, ¢/kW-h	5.6	5.4
COE*, for \$3.00/MMBtu gas, ¢/kW-h	7.1	6.3
COE*, for \$5.00/MMBtu gas, ¢/kW-h	9.1	7.4

* -after depreciation and State and Federal taxes.

** -no financial credits have been taken for the environmental benefits of the MCFC-based HEFPP.

TABLE 32
Combined Cycle Gas Turbine
Power Plant Capital Cost

AREA No.	PLANT SECTION DESCRIPTION	PROCESS CONT., %	COST, K\$ w/o CONT.
	Gas turbine generator		6,882
	HRSG		
	Steam turbine generator set		4,440
	Condenser		
	Cooling Tower		
	Transformer & Switchgear		
	Nat Gas Compressor		600
	Balance of Plant		3,885
	BACT SCR System (3ppmv NOx)	10	600
	Continuous Emissions Monitoring System		237
	Plant Control System (DCS)		787
	Subtotal, Process Plant Cost		17,431
	General Plant Facilities		872
	Engineering Fees		1,307
	Process Contingency (Using conntingencies listed above)		60
	Project Contingency, 15 % Proc. Plt. & Gen. Plt. Fac.		2,745
	Total Plant Cost		22,415
	Prepaid Royalties		
	Initial Catalyst and Chemical Inventory		20
	Startup Costs		362
	Spare Parts		87
	Working Capital		315
	Land, 1 Acres		25
	Total Capital Requirement (TCR)		23,225
		\$/kW	1,161

TABLE 33
Combined Cycle Gas Turbine
Annual Operating Costs

COST ITEM		QUANTITY		UNIT \$ PRICE		ANNUAL COST, K\$
Natural Gas	Fuel Type	<u>4,789</u>	MMBtu/D	<u>\$ 3.00</u>	/MMBtu	\$ 4,639
Consumable Materials						
	Demineralized Water	<u>3</u>	Mgal/D	<u>\$ 3.00</u>	/Unit	\$ 3
	Catalysts & Chemicals	<u> </u>	Unit/D	<u> </u>	/Unit	\$ 238
	Turbine Blade Replacement	<u> </u>	Unit/D	<u> </u>	/Unit	
		<u> </u>	Unit/D	<u> </u>	/Unit	
	Disposal Costs (If any)	<u> </u>	Unit/D	<u> </u>	/Unit	
Plant Labor						
	Operating Labor (incl. benef.)	<u>Operating Staff Number and Costs Similar to HEFPP</u>				\$ 640
	Supervision & Clerical					
	Maintenance Costs					\$ 261
	Insurance & Local Taxes					\$ 349
	Royalties					
	Other Operating Costs					
Total Operating Costs						\$ 6,129
By-Product Credits						
<u> </u>		<u> </u>	Unit/D	<u> </u>	/Unit	
<u> </u>		<u> </u>	Unit/D	<u> </u>	/Unit	
<u> </u>		<u> </u>	Unit/D	<u> </u>	/Unit	
<u> </u>		<u> </u>	Unit/D	<u> </u>	/Unit	
<u> </u>		<u> </u>	Unit/D	<u> </u>	/Unit	
Total By-Product Credits						
Net Operating Costs						\$ 6,129

TABLE 34

COE - \$/kWh, NPV Calculation														
CASE: 22 MW CC-GT Plant														
TOTAL PROJECT COST 23.23 \$MM														
End of Proj. Year	End of Opn. Year	Investment-related Cash Flows \$1,000			Before Tax Operating Cash Flows \$1,000			Tax Calculations \$1,000					Cash Flows	NPV
		Total Plant Cost	Working Capital	Total	Sales Revenue	Production Costs	Net Opn. Cash Flows	Depreciation (DDB)	Tax Allowance Earned in Previous Year	Taxable Profit in Previous Year	Tax Payable	After Tax Operating Cash Flow	Project Net Cash Flow	Discounted at IRR of
		(a)*	(b)		(c)	(d)		(e)			(f)			15.00%
1		(7,472)	(161)	(7,632)									(7,632)	(7,632)
2		(14,944)	(649)	(15,593)									(15,593)	(13,559)
3	1	0	0	0	11,098	(6,129)	4,969					4,969	4,969	3,757
4	2	0	0	0	11,043	(6,129)	4,914	5,121	5,121	(152)	65	4,979	4,979	3,274
5	3	0	0	0	10,988	(6,129)	4,858	4,097	4,097	817	(351)	4,507	4,507	2,577
6	4	0	0	0	10,932	(6,129)	4,803	3,278	3,278	1,581	(680)	4,123	4,123	2,050
7	5	0	0	0	10,877	(6,129)	4,747	2,622	2,622	2,181	(938)	3,810	3,810	1,647
8	6	(798)	0	(798)	10,655	(6,129)	4,525	2,098	2,098	2,650	(1,139)	3,386	2,588	973
9	7	0	0	0	11,043	(6,129)	4,914	1,678	1,678	2,847	(1,224)	3,689	3,689	1,206
10	8	0	0	0	10,988	(6,129)	4,858	1,342	1,342	3,571	(1,536)	3,323	3,323	944
11	9	0	0	0	10,932	(6,129)	4,803	1,074	1,074	3,784	(1,627)	3,176	3,176	785
12	10	0	0	0	10,877	(6,129)	4,747	859	859	3,944	(1,696)	3,052	3,052	656
13	11	(798)	0	(798)	10,655	(6,129)	4,525	687	687	4,060	(1,746)	2,780	1,982	370
14	12	0	0	0	11,043	(6,129)	4,914	196	196	4,329	(1,861)	3,052	3,052	496
15	13	0	0	0	10,988	(6,129)	4,858	196	196	4,717	(2,028)	2,830	2,830	400
16	14	0	0	0	10,932	(6,129)	4,803	196	196	4,662	(2,005)	2,798	2,798	344
17	15	0	0	0	10,877	(6,129)	4,747	196	196	4,606	(1,981)	2,767	2,767	296
18	16	(798)	0	(798)	10,655	(6,129)	4,525	196	196	4,551	(1,957)	2,568	1,771	165
19	17	0	0	0	11,043	(6,129)	4,914	196	196	4,329	(1,861)	3,052	3,052	247
20	18	0	0	0	10,988	(6,129)	4,858	196	196	4,717	(2,028)	2,830	2,830	199
21	19	0	0	0	10,932	(6,129)	4,803	196	196	4,662	(2,005)	2,798	2,798	171
22	20	0	0	0	10,877	(6,129)	4,747	196	196	4,606	(1,981)	2,767	2,767	147
23	21	(798)	0	(798)	10,655	(6,129)	4,525	196	196	4,551	(1,957)	2,568	1,771	82
24	22	0	0	0	11,043	(6,129)	4,914	196	196	4,329	(1,861)	3,052	3,052	123
25	23	0	0	0	10,988	(6,129)	4,858	196	196	4,717	(2,028)	2,830	2,830	99
26	24	0	0	0	10,932	(6,129)	4,803	196	196	4,662	(2,005)	2,798	2,798	85
27	25	0	1,000	1,000	10,877	(6,129)	4,747	196	196	4,606	(1,981)	2,767	3,767	99
28	26													
29	27													
		(25,606)			25,606								NPV = \$0.00	
(*)		Gas Turbine, HRSG & Steam Turbine, and BOP Major Overhaul every 5 years with outside contractors.							Operating Days/Year		351			
(a)		1998 Capital Investment of..... 23.23 \$MM							SOR Power Output, MW		20.2 @ cap fac.		0.92	
(b)		Recovered at end of project							Elect Value, \$/kWh		0.071			
(c)		Elect Value x Power Output (@cap fac.) x Opn. Days/year (Power Degrades @1%/yr between rebuilds)							EOR Power Output, MW		19.7			
(d)		From Operating Cost Table												
(e)		Double-declining-balance depreciation for 10 years, followed by straight line depreciation for remaining project life. Zero salvage value												
(f)		Combined State and Federal Tax rate of 43%.												

4. CONCLUSION

This study has confirmed that Pressurized Molten Carbonate Fuel Cell technology can be combined with gas turbine generator technology to produce a power plant with 70% low heating value efficiency when utilizing natural gas as a fuel. The study also confirmed that the HEFPP concept produced significantly lower 6.3 cents/kWh cost of electricity when compared to a similar sized gas turbine combined cycle plant with a 7.1 cents/kWh cost of electricity based on a natural gas fuel cost of \$3.00/MMBtu. The HEFPP cost of electricity sensitivity to fuel cost was investigated with fuels ranging from \$1.50 to \$5.00 per MMBtu. In all cases the HEFPP concept was considerably cheaper in cost of electricity than a gas turbine combined cycle plant of similar size.

This study has assumed that investments in the HEFPP components will be made to drive the component costs down and to eliminate the technical issues. Specific components requiring further investment and development time include fuel cell stacks, gas turbine generator, recycle blower, power module vessels, inverter, reformer, and plant control philosophy.

Most of the development items on the molten carbonate fuel cell are already being addressed under M-C Power's Process Design Improvement program funded by DOE.

With the HEFPP concept, the fuel cell will have design and development issues related to operation at 610 kPa. The gas compositions and flow rate will affect pressure drop and cell performance. The higher pressure may result in higher output per stack because the internal manifolds will shrink and the active area increase. Bench, small area, and full area stack performance testing will be required to confirm predicted current densities and voltages with actual gas compositions at the 610 kPa pressure. Endurance testing of fuel cell stacks will also be required to determine if pressure has an effect on stack components. Series connections for four cells will also have to be designed and tested.

Many of the turbine generator components are of new design or must be scaled up or down to meet the specific application. These components will require engineering, fabrication development, and prototype testing prior to use in the power plant. Specific examples include the active magnetic bearings, direct drive alternator, turbine generator controls, and the power conditioner.

The recycle blower temperature problems are being worked through at the Miramar plant, but the shaft sealing solutions at a pressure of 606 kPa must be developed and tested. The limited number of manufacturers with experience at this operating pressure also presents a problem.

The power module vessel will require some design work to conquer the design constraints imposed both internally and externally. In this case, Bechtel National, a vessel fabricator, and the module fabricator will have to work closely to develop the design. Input will be required from M-C Power and IHI for the fuel cell and reformer installation/operation/maintenance issues, respectively.

The inverter design and development issues appear much more straightforward than the other components mentioned above. The most important inverter design challenges include the efficiency improvement, technology selection, product repackaging, and performance testing issues. Most of the issues can be resolved by a contractual relationship with an inverter manufacturer. Once the product is developed and tested successfully, the inverter manufacturer may have a significant competitive advantage in technology and product introduction over rivals who have not developed or tested products.

The reformer design will require some development work to accommodate the higher operating pressure of 610 kPa. At higher pressures, the specific volume of the gases is smaller which will reduce the pipe diameter and plate spacing. The higher pressure will however require thicker pipe materials to be used. Some development work and testing will also be required to meet the catalyst life goal of 80,000 hours or ten years.

The overall plant control concept also will require refinement. Integration of the turbine generator and multiple fuel cell inverters will require design work. Strategies for response to electrical system transients such as loss of electric grid, ground faults in a single phase or neutral must be developed. Modeling work will be required to determine HEFPP capabilities for load following, load shedding, and plant turndown.

The plant control system will also have to integrate the individual component control systems for the gas turbine generator and inverter with the other equipment in the plant.

Finally, the technical challenges mentioned above are not insurmountable and can be overcome with time and targeted funding. Clearly, much of the technology exists today and has been proven on a smaller scale by M-C Power at Miramar and in the future will be proven by M-C Power's 450 kW class molten carbonate fuel cell hybrid power plant using a micro gas turbine generator. The HEFPP concept is feasible and predicted performance levels can be met. This study has presented the plans and approaches required to solve the performance, life and cost issues by 2010 or sooner. The proposed HEFPP concept will produce a low cost of energy and environmentally friendly plant with significantly lower emissions than the gas turbine generator combined cycle plants of today.

Fuel Cell/Micro-Turbine Combined Cycle

Final Report

August 1998 – December 1999

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ABSTRACT

A wide variety of conceptual design studies have been conducted that describe ultra-high efficiency fossil power plant cycles. The most promising of these ultra-high efficiency cycles incorporate high temperature fuel cells with a gas turbine. Combining fuel cells with a gas turbine increases overall cycle efficiency while reducing per kilowatt emissions. This study has demonstrated that the unique approach taken to combining a fuel cell and gas turbine has both technical and economic merit. The approach used in this study eliminates most of the gas turbine integration problems associated with hybrid fuel cell turbine systems. By using a micro-turbine, and a non-pressurized fuel cell the total system size (kW) and complexity has been reduced substantially from those presented in other studies, while maintaining over 70% efficiency. The reduced system size can be particularly attractive in the deregulated electrical generation/distribution environment where the market may not demand multi-megawatt central stations systems. The small size also opens up the niche markets to this high efficiency, low emission electrical generation option.

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List of Acronyms and Abbreviations

AC	Alternating Current
AES	Advanced Energy System
ASR	Area Specific Resistance ($O \cdot cm^2$)
BOP	Balance of Plant
cfm	Cubic feet per minute
COE	Cost of Electricity
Cpn TM	Co-planar, n-stack
DC	Direct Current
DOE	United States Department of Energy
GRI	Gas Research Institute
HEFPP	High Efficiency Fossil Power Plant
HHV	Higher Heating Value
HT	High Temperature
kW	Kilowatt (1000 W)
kWe	Kilowatt Electric (1000 W)
LHV	Lower Heating Value
MM Btu	Million British Thermal Units
MTI	McDermott Technology, Inc.
MW	Megawatt (1,000,000 W)
NREC	Northern Research and Engineering
ODS	Oxide Dispersion Strengthened
OEM	Original Equipment Manufacturer
PLC	Programmable Logic Controller
PM2000	Advanced metallic material from Plansee GmbH, Germany
PowerWorks TM	NREC's micro turbine
PSOFC	Planar Solid Oxide Fuel Cell
SI	SI is an abbreviation for "Le Systeme Internationale d'Unites."
SOFCo	Solid Oxide Fuel Cell Company Research and Development Limited Partnership with MTI and Ceramatec
TIT	Turbine Inlet Temperature
TCE	Coefficient of thermal expansion
VHT	High Temperature

EXECUTIVE SUMMARY

A wide variety of conceptual design studies have been conducted that describe ultra-high efficiency fossil power plant cycles. The most promising of these ultra-high efficiency cycles incorporate high temperature fuel cells with a gas turbine. Combining fuel cells with a gas turbine increases overall cycle efficiency while reducing per kilowatt emissions. Fuel cells are widely recognized as one of the most promising family of technologies to meet future power generation requirements. Since fuel cells directly convert fuel and an oxidant into electricity through an electrochemical process, they can achieve operating efficiencies approaching 70% - nearly twice the efficiency of conventional internal combustion engines. Fuel cells produce very low levels of pollutant emissions (NO_x , SO_x , and CO_2). They are also amenable to high-volume production as standardized power modules.

This conceptual study has demonstrated that the unique approach taken to combining a fuel cell and gas turbine has both technical and economic merit. By using a micro-turbine, and a non-pressurized fuel cell the total system size (kW) has been reduced substantially from those presented in other studies, while maintaining over 70% efficiency. The approach used in this study eliminates most of the gas turbine integration problems associated with hybrid fuel cell turbine systems. The reduced system size can be particularly attractive in the deregulated electrical generation/distribution environment where the market may not demand multi-megawatt central stations systems. The small size also opens up the niche markets to this high efficiency, low emission electrical generation option.

While the study has discovered no technical obstacles to success, a sub-scale technology demonstration would reduce the risk of performance and enable a full-scale commercial offering. Demonstrating a full size micro-turbine, with a single fuel cell module would prove the concept as well as the major components and balance of plant that would be needed in a full-scale system.

1 INTRODUCTION

Global demands for additional power generation over the next twenty years are about 2 million megawatts, of which 490,000 megawatts are projected to be powered by natural gas (McDermott internal study). As a result of utility deregulation in the U.S., concerns with real and perceived health issues, and capital costs associated with the distribution and transmission of electricity, approximately 30% of this additional natural gas capacity will consist of modular power plants located close to the user. Fuel cells combined with a micro-turbine are a logical candidate to meet this need. They offer modularity, increased fuel efficiency, and low emissions. Major gas and electric utilities have shown an interest in investing in both fuel cells and micro-turbines (McDermott confidential communications).

A wide variety of conceptual design studies have been conducted that describe ultra-high efficiency fossil power plant cycles. The most promising of these ultra-high efficiency cycles incorporate high temperature fuel cells with a gas turbine. Combining fuel cells with a gas turbine increases overall cycle efficiency while reducing per kilowatt emissions. Fuel cells are widely recognized as one of the most promising family of technologies to meet future power generation requirements. Since fuel cells directly convert fuel and an oxidant into electricity through an electrochemical process, they can achieve operating efficiencies approaching 70% - nearly twice the efficiency of conventional internal combustion engines. Fuel cells produce very low levels of pollutant emissions (NO_x , SO_x , and CO_2). They are also amenable to high-volume production as standardized power modules.

The operating characteristics of a fuel cell/micro-turbine power plant have several important ramifications to the energy service industry. Successful development and commercialization of dispersed fuel cell/micro-turbine power generators will allow:

- Siting flexibility with environmentally friendly energy systems,
- Improved quality of energy services at a reduced cost,
- Ability to rapidly respond to customer needs with modular energy systems,
- Improved utilization of clean natural gas, of which the nation has an abundant domestic supply, and
- Facilitate implementation of clean air policies

This report documents the results of a conceptual technical and economic evaluation of an innovative and unique integration of high temperature fuel cells with a gas turbine. The technical approach described in this program focuses on a planar solid oxide fuel cell (PSOFC) combined with a micro-turbine. PSOFCs have the potential for low cost manufacturability. McDermott Technology Inc. (MTI) has a development program in progress to address various methods of low cost, high volume manufacturing of PSOFCs and stacks. A low cost PSOFC combined with a sub-megawatt gas turbine creates a highly attractive product for the deregulated power market. Other studies have focused on a pressurized fuel cell gas turbine system. This study presents a unique, non-

pressurized approach to combining PSOFCs with gas turbines. One of the key issues addressed in this study is that of system economics versus efficiency. The objective is to optimize the economic viability associated with the development of PSOFC/micro-turbine systems while balancing the need for operating efficiency and low emissions. Part of this economic analysis will include an economic analysis of the PSOFC stack operating point.

Based upon previous analyses by MTI and other solid oxide fuel cell related companies, PSOFC/turbine systems have been shown to be capable of operating at efficiencies greater than 70%. Overall, the HEFPP program goals of developing a fuel cell / turbine power plant concept of 20 MW with a net efficiency of greater than 70% have been met. The goals have been exceeded in that the efficiency target of 70% has been met at a sub-megawatt plant size. The smaller plant size gives more flexibility in responding to market demands.

2 RESULTS AND DISCUSSION

2.1 Fuel Cell / Micro-turbine system analysis

The analysis of the fuel cell micro turbine combined cycle is described below. The overall process is described first followed by engine fuel cell integration concepts, design assumptions, a description of the major equipment, input data, a heat and material balance and then the modeling approach and methodology.

2.1.1 Process Description

This design utilizes a unique combination of fuel cell, turbine and recuperator to achieve a highly efficient cycle in a small, compact market-driven size. The flow and heat requirements of components in the micro-turbine and Solid Oxide Fuel Cell Company (SOFCo) CpnTM (Co-planar, n-stack) fuel cell module have been matched, resulting in a highly integrated package. The micro-turbine is a 70 kW gas turbine engine under development by Northern Research and Engineering Corporation (NREC). The SOFCo CpnTM concept evolved from recognizing the impact of the balance of plant (BOP) on the economy and efficiency of the total fuel cell system. The design optimizes the total fuel cell system and maximizes the efficiency of the system while simultaneously reducing the number of high temperature components peripheral to the stack. The CpnTM module, shown in Figure 1, consists of a multi-stack arrangement that enhances efficiency through effective thermal coupling of the stacks and the fuel processors. The CpnTM power system is comprised of planar PSOFC stacks, fuel processor components and the BOP equipment. The most significant feature of the CpnTM is the Thermally Integrated PSOFC Module that houses the fuel cell stacks, reformer catalyst tubes, and a spent fuel burner.

Figure 1: Cpn 4 stack module

A process schematic for the fuel cell/micro-turbine combined cycle is shown in Figure 2. The state parameters for the system are listed in Table 1, and the design parameters used in the system analysis are listed in Table 2. The air is first compressed in the compressor at a 3:1 pressure ratio. The air is then heated to 1600°F in a high temperature recuperator by utilizing exhaust gas from the CpnTM module. The hot, high-pressure air is then

Table 1 - State Parameters For 700 kW Fuel Cell/Micro-turbine Combined Cycle				
State Point	Flow kg/s (lbm/hr)	Temperature C (F)	Pressure kPa (psi)	Enthalpy J/kg (Btu/lbm)
1	0.662 (5256)	15 (59)	101.3 (14.7)	-1.35e5 (-58.04)
2	0.622 (5256)	178 (352)	304 (44.1)	3.11e4 (13.35)
3	0.662 (5256)	871 (1600)	300.9 (43.6)	8.01e5 (344.37)
4	0.662 (5256)	872 (1600)	293.5 (42.6)	8.01e5 (344.37)
5	0.622 (5256)	639 (1182)	106.9 (15.5)	5.31e5 (228.29)
6	0.274 (2175.7)	862 (1583)	104.4 (15.1)	7.72e5 (331.90)
7	0.022 (177)	15 (59)	204.7 (29.7)	-4.74e6 (-2037.83)
8	0.040 (315.4)	25 (77)	120 (17.4)	-1.60e7 (-6878.76)
9	0.040 (315.4)	108 (226)	120 (17.4)	-1.33e7 (-5717.97)
10	0.062 (492.4)	95 (202)	120 (17.4)	-1.02e7 (-4385.21)
11	0.062 (492.4)	253 (488)	120 (17.4)	-9.81e6 (-4217.54)
12	0.062 (492.4)	816 (1500)	120 (17.4)	-3.81e6 (-1638.01)
13	0.138 (1091.7)	862 (1583)	116.8 (16.9)	-9.96e6 (-4282.03)
14	0.724 (5748.4)	913 (1675)	103.7 (15.0)	-1.31e6 (-563.20)
15	0.724 (5748.4)	910 (1670)	103.7 (15.0)	-1.32e6 (-567.50)
16	0.724 (5748.4)	358 (676)	103.6 (15.0)	-2.02e6 (-868.44)
17	0.724 (5748.4)	330 (626)	102.8 (14.9)	-2.06e6 (-885.64)
18	0.724 (5748.4)	200 (391)	102.6 (14.9)	-2.21e6 (-950.13)

expanded through the turbine providing power for the compressor and electrical generation. The turbine produces 68.8 kWe of net electrical power or 9.5% of the total. The air is then sent to the fuel cell.

Natural gas is mixed with steam that was generated in the steam generator coil, and the mixture is then heated further in the fuel heater. The heated fuel/steam mixture is then sent to the steam reformer. In the steam reformer, the fuel-steam mixture passes over steam reforming catalyst and is processed into hydrogen rich reformat and sent to the fuel cell. The hydrogen and carbon monoxide in the fuel are electrochemically oxidized in the fuel cell producing electrical power. The fuel cell produces 657.6 kW of electrical power or 90.5% of the total. The unreacted fuel exiting the fuel cell is burned with the fuel cell cooling air in the fuel cell module enclosure, further boosting the exhaust temperature and providing heat to drive the steam reforming reactions in the steam reformer. The hot gas leaving the CpnTM is then sent to the high temperature recuperator

where it heats the compressed air, and then is sent to the fuel heater where it heats the fuel and steam mixture. The fuel heater exhaust is used to provide heat to generate the steam that is mixed with the natural gas. The exhaust exits the process at 200°C. The exhaust could be used to generate low-pressure process steam or space heating in a cogeneration heat exchanger.

Figure 2: Process Schematic for the Fuel Cell MicroTurbine Combined Cycle

Table 2 - Design Parameters for 700 kW Fuel Cell/Micro-Turbine Combined Cycle	
Fuel	Natural gas 0.96 CH ₄ , 0.02 N ₂ , 0.02 CO ₂ LHV = 4.81E7 J/kg (20,659 Btu/lbm)
Turbine pressure ratio	3:1
Recuperator effectiveness	0.947
Fuel cell	15,616 cells
Operating voltage	0.76 V/cell
System heat loss	0.5% of heat input
Inverter efficiency	95%
Generator efficiency	98%
Gear box loss	5%

The performance of the fuel cell /micro-turbine combined cycle is summarized in Table 3, and the duties of heat transfer equipment are listed in Table 4. The process produced 726.4 kWe of power at 71.2% LHV efficiency. For the HEFPP program requirement of a multi-megawatt system, the process can be considered a module. Twenty-eight modules would produce 18.4 MW. The modular concept is an attractive alternative for the power plant, providing flexibility in turndown, dispatching, and annual maintenance downtime.

Table 3 - Performance Summary for 700 kW Fuel Cell/Micro-Turbine Combined Cycle	
Mass flow rate of natural gas	80.3 kg/hr (177 lbm/hr)
Gas flow * LHV	1,072,897 W (3,656,643 Btu/hr)
Gross Power	761 kW
Net Power	726.4 kW
Efficiency, LHV	71.2
Contributions to Power	
Fuel cell	657.6 kW
Turbine	68.8 kW
Inverter loss	-34.6 kW
Generator loss	Included in turbine power calculation
Gear box loss	Included in turbine power calculation

Table 4 - Component Duty Summary for 700 kW Fuel Cell/Micro-Turbine Combined Cycle	
Component	Duty (kW)
Fuel heater	24.2 (82,703 Btu/hr)
Reformer	372.4 (1,270,532 Btu/hr)
Recuperator	510.2 (1,741,011 Btu/hr)
Spent fuel burner	42.5 (144,979 Btu/hr)
Steam generator	110.2 (375,896 Btu/hr)

Fundamental requirements for the engine operating in a PSOFC system are as follows:

- During the power plant startup cycle, the engine provides hot air for PSOFC preheat and eventual power generation. Through the preheat period (~20 hours) the engine will operate at a pre-selectable constant turbine inlet temperature, delivering between 45 and 80 kWe AC power to the grid depending on ambient temperature and the turbine-inlet temperature set point. The cell remains inactive during this phase.
- At the conclusion of the preheat cycle the cell will have reached thermal equilibrium at the engine exhaust temperature of roughly 1200°F, sufficient for reformer operation. The PSOFC controller then modulates fuel supply to the reformer in order to drive recuperator-inlet temperature toward the 1740°F design-point level. As recuperator preheating occurs fuel supply to the engine combustor is reduced gradually to zero, maintaining turbine-inlet temperature roughly at the 1600°F design target. Except for monitoring of safety conditions by the engine controller, engine operation is governed at this point entirely by the PSOFC controller.
- During normal “design-point” operation, running with the combustor off, engine power augments PSOFC electrical output roughly by 10% while supplying hot air to the cell. The engine controller continues to monitor safety conditions, alerting the PSOFC controller in the event of a fault.
- Under part-load demand with the combustor off, the engine provides reduced electrical output and flow, but generally a higher fraction of PSOFC power than at design conditions.

Additional flexibility in the management of the power plant starting sequence is made possible with the use of the hydraulic drive system fitted with this engine. This proprietary NREC technology relies on a miniature hydraulic turbine mounted on the gasifier shaft, fed by a high-velocity jet of lubricating oil drawn from the engine sump. An attractive feature of this system for the current application is its ability to run for extended periods, delivering 200 to 400 cfm to the PSOFC. This may be applied as a pre-starting or cool-down operating mode.

During normal power plant operation the combustor is not fired. The engine power and flow under these circumstances depends entirely on turbine-inlet temperature and ambient conditions, the former dependent chiefly on recuperator-inlet temperature. The generator remains synchronized to the utility grid in all conditions. An attribute of this system is that turbine-inlet temperature will not drop substantially during power plant turndown, hence the engine will continue to run at high efficiency. At low PSOFC current density, with flow roughly at the design value, oxidant utilization will be low, boosting PSOFC efficiency.

2.1.2 Engine/Fuel cell Integration Concepts

The critical engine/fuel cell integration challenge is the development of a recuperator capable of accepting gas-inlet temperatures in excess of 1740°F, well beyond the capability of superalloys in this service. The design concepts developed in this study rely on the use of the advanced material PM2000 (Plansee GmbH, Germany), a so-called oxide-dispersion-strengthened (ODS) powdered-metal alloy. Although some questions remain regarding formability of this material in our manufacturing process, provided the problems can be overcome (and we expect that they can) a recuperator very similar in design to that of our current unit will be suitable. This greatly simplifies the job of building the recuperator and of packaging it in our engine.

Despite our optimism that PM2000 can be made to work, we've allowed in our cost projections for a more proven alternative solution in the form of a "hybrid" recuperator. This is the concept put forth in our original proposal, which makes use of a high-temperature tube-shell unit inserted in series with a recuperator similar to our current design. Compared to the single-recuperator approach this concept carries a substantial cost penalty, mostly from the high cost of the tube-shell unit, but also from costs associated with modifying the existing recuperator case and supports. The hybrid approach also carries a performance penalty in the form of additional pressure loss for the same thermodynamic effectiveness.

The hybrid concept has been evaluated to the extent that rough cost projections can be made, but explicit design layouts have been developed only for the single high-temperature recuperator approach. In part this reflects our view as to the superiority of the latter concept, and our optimism that it can be made to work.

The remaining integration challenges are largely associated with ducting hot gases with acceptable pressure and heat losses.

The engine modeled in this study is based on NREC's PowerWorks™ engine. The PowerWorks™ engine was originally developed in the early 1980s under GRI sponsorship. It is now in its fourth generation of development. It incorporates a single-spool gasifier and a low-speed power turbine. A single-stage gear box reduces the 44,000 RPM power turbine to 3600 RPM so that a conventional generator can be used. As a stand-alone machine the PowerWorks engine is tightly packaged to achieve these

objectives, and significant re-orientation of components is needed to allow for ducting transitions to the fuel cell. The following existing engine systems will require substantial rework:

- ✓ chassis
- ✓ compressor-recuperator duct
- ✓ recuperator inlet plenum/header
- ✓ exhaust plenum
- ✓ lubrication-system piping

Two system layouts (Concepts 'A' and 'B') are shown in Figures 3 through 9. Both concepts make use of a single recuperator core identical in size to the current PowerWorks recuperator, consistent with the use of the advanced high-temperature material mentioned above.

Both concepts are topologically identical, and there is no clear choice with regard to ease of fabrication or cost. Pressure losses will be roughly comparable, the specifications discussed earlier having been used as an approximate basis for pipe sizing in both cases. Concept A has a small advantage in terms of exposed surface area of hot ducting, but at the expense of slightly more challenging fabrication requirements. Concept A may also pose a bit more difficulty in achieving a flow balance among the modules, and service accessibility looks to be not as good. For these reasons, Concept B has a slight edge, but the choice may ultimately come down to site requirements such as proximity of inlet/exhaust ducting and availability of floor space.

Identical construction is assumed for all fuel cell modules in both concepts. Because of the requirement for vertical stacking of the cells, the gas-inlet aperture on all but the uppermost module in the stack will face its neighbor above. This requires that a spacer be included between each module to provide area for a supply duct. It is assumed that these features, the spacers and supply duct, will be incorporated into the module design.

CONCEPT

Figure 3 Concept A Isometric View

Figure 4 Concept A, Plan View

CONCEPT A

Figure 5 Concept A, Elevation View

CONCEPT B

Figure 6 Concept B, Isometric View

CONCEPT B

Figure 7 Concept B, Isometric View

F.C. INLET DUCT
(16 TOTAL)

CONCEPT B

Figure 8 Concept B, Plan View

POWER-TURBINE
DISCHARGE
(F.C. INLET)

CONCEPT B

Figure 9 Concept B Elevation View

Manifold Systems

The manifold systems comprise large horizontal-running ducts from the engine and vertical ducts at each stack of fuel-cell modules. Ducts have been sized to limit flow velocity to 60 fps (feet per second), consistent with the pressure-loss specifications cited earlier and in an effort to promote uniform flow distribution. Three inches of insulation will limit outer-wall temperatures below 240°F.

Manifolds would be constructed from light-gauge superalloy sleeves wrapped with insulation and surrounded by heavier-gauge low-grade Stainless Steel. The .035-inch¹ thick inner sleeves are segmented, allowing relative sliding to accommodate thermal growth. This choice of wall thickness is based on a 10,000-hr life target based on a .010-inch margin. This construction minimizes the weight of expensive materials and avoids use of metal bellows. The strong outer casing also makes for a straightforward approach to supporting heavy piping, although for clarity the structural framework has been omitted from the figures.

Single-Recuperator Approach

This is the preferred option as discussed earlier, provided that the advanced material PM2000 can be successfully utilized. Strength, oxidation resistance, and thermal conductivity are ample at 1740°F based on published data. The material is currently available from Plansee in the appropriate gauge thicknesses.

Successful manufacture of a PM2000 recuperator requires that it be brazable, and that it be amenable to intricate forming in our fin-folding process. Current experience with the manufacture of honeycomb turbine seals proves brazeability, and tentatively indicates that formability will be acceptable. This latter question cannot yet be answered definitively, however. Samples have been sent to a die vendor for further examination of this question.

Cost of PM2000 remains a pressing issue for small quantities, but would be expected to become manageable for production quantities. For preliminary budgeting purposes a speculative cost of \$200 per lb. was used. Even at this premium price, for prototype development the single-recuperator approach maintains a compelling cost advantage over the tandem concept.

To explore producibility issues and to obtain operational experience, a partial heat-exchanger core should be built and rig-tested prior to moving forward with construction of a full unit. This approach was taken during the development phase of our current recuperator, enabling the transition to volume production with low risk. A recuperator core comprising ten cells is envisioned. Experience gained while preparing the test article would likely suggest modifications to the manufacturing process. Testing would

¹ The Aerospace Structural Metals Handbook indicates .002-inch loss of material in IN625 after 800 hours in 1,800°F air.

include coupon burst-tests to establish strength of the brazed structure, and partial-core burst tests to measure the strength of the heat exchanger. A rig similar to that shown in Figure 10 would be constructed using high-temperature materials, and thermal-cycling and endurance testing carried out. Partial-core testing is envisioned as a separate task and is not budgeted under the current program.

Tandem-Recuperator Approach

The tandem-recuperator approach represents a compromise that can hopefully be avoided, but investigation remains worthwhile in the event that the single-recuperator strategy is unsuccessful.

The strategy is to make use of a recuperator whose construction would resemble our current design, but likely made from a higher-temperature alloy; this is termed the *high-temperature* (HT) recuperator in what follows. In series with the HT recuperator is a commercial tube-shell unit capable of withstanding the full 1740°F requirement; this is the *very-high-temperature* (VHT) recuperator. Construction of the HT recuperator would be a very straightforward application of our current manufacturing technology, and carries very low risk. The VHT recuperator is the more serious design challenge.

Ceramics would appear to be an obvious choice for this application in view of their high-temperature capability and low thermal expansion. Tube-shell ceramic heat exchangers are under current development by United Technologies Corporation and CHX Engineering, with units having been successfully tested at temperatures up to 2000°F. For the proposed application the most compelling disadvantage of these units is their large size compared to a compact plate-fin design, which carries a penalty in terms of the cost of the unit and in terms of integration with the PSOFc power plant.

A cost tradeoff exists between the thermodynamic effectiveness of the two recuperators. For a constant overall effectiveness of the pair, increased effectiveness of the VHT unit reduces the inlet temperature to the HT unit, increasing the size and cost of the former while enabling the latter to be fabricated from lower-cost materials.

It is preferable from a cost perspective to find the temperature limit of the HT heat exchanger, as the VHT unit is expected to dominate the cost. Based on a rough preliminary study of this issue (see table attached), we settled on an effectiveness near 66% as a rough optimum for the VHT unit. For the HT unit this corresponds to an inlet temperature of 1400°F at PSOFc design conditions, although it is unclear at this point whether an IN625 HT heat exchanger will have a satisfactory life at this temperature differential and pressure loading ($\Delta T=1150^\circ\text{F}$, $\Delta p=38$ psi). A test program would be needed to confirm acceptability.

Figure 10 Recuperator Arrangement

Table 5: Hybrid Recuperator Options								
	ϵ_1	ϵ_2	T1, F	T2, F	T6, F	T3, F	T4, F	T5, F
IN625 HT Recup (cost basis)	66%	93%	1,740	1,400	1,600	550	350	1,327
347SS HT Recup (push)	75%	93%	1,740	1,250	1,600	521	350	1,187
347SS HT Recup (safe)	77%	93%	1,740	1,200	1,600	512	350	1,141
IN625 HT Recup (push)	56%	93%	1,740	1,500	1,600	569	350	1,420

2.1.3 Design Assumptions

The major system parameters used in this study are shown in Table 6.

Table 6 – Key System Parameters for 700 kW Fuel Cell/Micro-Turbine Combined Cycle	
<i>Equipment Assumptions</i>	
Number of fuel cell modules	16
Number of stacks per module	4
Number of cells per stack	244
Cell area	327 cm ² (50.7 in ²)
Cell voltage	0.76 V/cell
Cell operating temperature	862 °C (1583°F)
Pressure loss, fuel cell + reformer + burner	3447 Pa (0.5 psi)
Inverter efficiency	95%
Recuperator effectiveness	0.947
Turbine pressure ratio	3:1
<i>Process Engineering Assumptions</i>	
LHV of fuel	4.81E7 J/kg (20,659 Btu/lbm)
Heat loss from system	0.5% of heat input
Water flow	1.7 steam/carbon mole ratio
NOx emissions	Less than 1 ppm

The natural gas used in this study was specified by the DOE and is typical of a mid-range heating value gas delivered in the United States.

We assumed that thermal losses from the process are equal to 0.5% of the heat input. This is a reasonable assumption given the temperatures and sizes of equipment involved in the process. This assumption is also consistent with experience on similarly sized processes.

An inverter to convert DC to 60Hz AC voltage is a key component for any fuel cell power plant. Development of an inverter is not envisioned to be part of this program. Currently, inverters with 95% inverter efficiency are commercially available, and this was the assumed efficiency used in this study.

2.1.4 Major Equipment

Fuel Cell Module

SOFCo's CPnTM module design provided the basis for the fuel cell module used in this power plant. The modified CPnTM concept used in this module design, thermally integrates the PSOFC stacks and the methane steam reformer, as well as the air and fuel manifolds. The module was scaled to 43 kW and a preliminary layout was developed. During this design effort the specifications of the burner that utilized fuel cell exhaust were revised and the spent fuel burner was eliminated. The spent fuel is now burned in the enclosure. The burner specification task in the program plan was revised. The burner specified in this program task was shifted to the micro-turbine startup combustor. The catalytic steam reformer was sized using commercially available catalyst, assumed to be Haldor-Topsoe R67R or equivalent. The methane steam reformer in an integral component of the fuel cell module, and thermally, is highly coupled to the fuel cell stacks. The catalyst loading of the steam reformer was sized conservatively at 600/hr gas space velocity.

Note that on the process schematic the desulfurizer was not shown. The desulfurizer was not modeled in the simulation as this is a mature, stable technology. However the desulfurizer was sized. A desulfurizer sized for a five year life of the sorbent would be a vessel 15.25 cm (6.0 in.) in diameter and 122 cm (48.0 in.) long. The sorbent was assumed to be Haldor-Topsoe HTZ-3 or equivalent. After 5 years the sorbent is easily changed and the spent sorbent is non-hazardous and needs no special disposal.

Engine

The PowerWorksTM engine incorporates the most widely accepted industrial gas turbine mechanical configuration, known commonly as a *free-power-turbine* design. The gasifier turbo-compressor section delivers hot pressurized combustion gas to the power turbine, which provides a versatile and mechanically simple power-take-off. The mechanical design is such that the power turbine is overhung from its bearing core and thermally isolated from the load. Thermal isolation of the hot sections from the load is fundamental to maintaining a stable rotating assembly, and minimizes performance losses.

In addition to simplifying the load connection, the twin turbines split the cycle work, thus operating at roughly half the stress of a single turbine assigned to the same duty. The gasifier section is formed from a low cost turbocharger. NREC customizes the aerodynamics and ruggedizes the turbine housings.

The turbomachine utilizes proven pressurized-oil floating-ring journal bearings. These bearings are the most reliable used in the turbomachinery field, often compiling hundreds of thousands of trouble free hours of operation in a gas turbine engine. Large dimensional clearances on a thick oil film make them exceptionally durable and tolerant to erosion from contaminants. In geared applications, an angular contact ball bearing is

used on the load end of the shaft. The B1 life of all bearings in the PowerWorks™ turbomachinery, as defined from a large industry data base, exceeds 100,000 hours.

Recuperator

NREC's recuperator has been designed for the challenging "micro-turbine" product specifications. Low cost and exceptional durability are its primary features. The design has been thoroughly tested over thousands of hours of extreme cycling. No other commercial recuperator could stand up to the high pressure and rapid thermal cycling that has been prescribed by our US Navy qualification program. NREC began production of the recuperator in our newly capitalized facility in Portsmouth, NH in April, 1997.

Two alternative recuperator strategies are proposed in connection with the current program. In both cases the design is substantially identical to that of our production unit, but higher-temperature materials are substituted. These strategies are discussed in a separate section of this report.

Combustor

The combustor proposed for the integrated PSOFC package would be a modification of the standard patented PowerWorks™ design, originally developed in 1990 in collaboration with SoCal Gas. It has consistently demonstrated NOx levels below 9ppmv, with exceptionally good turndown stability and proven durability.

Departure from the standard PowerWorks™ design is needed to limit combustor pressure loss during unfired operation. Combustor inlet temperature under these conditions will be in the vicinity of 1600F, whereas the current running condition is around 1200F. The design change needed to accommodate this difference is straightforward, and is roughly a matter of increasing the effective flow area of the combustor.

Generator/gearbox

The standard PowerWorks™ package incorporates a single-stage helical gear set to transfer power from the turbine to the 3600 RPM generator. The low-torque, high-sliding-velocity results in exceptional design-life margins. At the conditions specified for the PSOFC, the gear and bearing life exceed one million hours.

A commercial 2-pole 3600 RPM induction generator is standard with the PowerWorks™ package, and for a production version of the proposed system would be the probable choice. The manufacturer predicts a B10 life of 160,000 hours for normal service. The generator has been conservatively selected and operates in a cool, clean, low-vibration environment. For cold weather and extended peaking-power operation, a higher power rated generator can be provided. An optional synchronous generator can also be substituted for grid-isolated operation, as proposed in connection with the current experimental program.

Controls and starting

The PowerWorks™ engine is currently controlled by an industrial programmable logic controller (PLC) while undergoing laboratory testing. The production version of the product will incorporate Ingersoll-Rand's standard Intellesys™ micro-processor based controller. The PLC is well suited for the initial PSOFC/GT demonstration unit because of its versatility, allowing basic engine control and safety functions to be integrated readily with those of the PSOFC. During start-up, the controller monitors the power-turbine speed as it accelerates toward synchronous operation, at which point the induction generator is latched to the grid and remains at a fixed 3600 rpm. During the PSOFC preheat period, the controller governs engine fuel throttle to maintain the prescribed turbine-inlet temperature set-point.

The engine is started by activating the hydraulic starter, a miniature turbine located between the bearings of the gasifier turbocompressor. This can drive the gasifier to modest speeds for indefinite periods without harming the engine or components. Depending upon the capacity and set-point of the PowerWorks oil pump, the engine centrifugal compressor delivers 15 to 25% of the rated flow through the system. After starting (igniting) the engine the oil pump drops back to a low speed-setting as it continues to feed lubricant to the bearings.

Natural gas boosting system

NREC provides a special-duty natural-gas booster package built from a mature Ingersoll-Rand oil-free compressor product. It is capable of delivering between 10 and 60 icfm to about 60 icfm (inlet cubic feet per minute). For the proposed application the booster would operate at roughly 25 icfm with a parasitic electrical power consumption of about 2 kW.

Alternative Turbomachinery concepts evaluated

Over the course of the PowerWorks™ development, trades were evaluated in a number of areas relevant to this project. Significant results and conclusions are discussed in the following paragraphs.

Single-shaft vs. twin-shaft turbomachines

Several attempts have been made to integrate a shaft-speed alternator into the single spool turbo-compressor. Locating the alternator between the bearings with an over-hung turbine and compressor is a common mechanical arrangement, implemented in the AES 50 kWe cogeneration project by Allied Signal (1984-1990) and the Chrysler Patriot by SatCon and NREC (1994-1996). One of the attractions of this arrangement is that it affords a clear aerodynamic path for the inlet and exit flows from a radial turbine and centrifugal compressor. The primary challenge in this design is the cooling system

associated with the alternator and bearings. The high power-density of the high-speed alternator, with combined electrical and windage losses of nominally 10%, coupled with the close proximity of the turbine section, demands large quantities of liquid cooling. Neither of the two programs cited above resolved the interrelated cooling, stress, and dynamics issues associated with this configuration.

Relocation of the high-speed alternator to the inlet of the compressor avoids many of the problems encountered with the alternator cooling. The disadvantages are increased bearing-system cost, and performance losses. To support the dynamic system, usually three rather than two high-speed bearings are required. This results in tight-tolerance manufacturing methods typical of the aerospace industry. Avoidance of this manufacturing operation is a primary distinction between high cost aerospace turbocompressors and the common industrial turbocharger.

The performance compromises associated with the compressor-end shaft speed alternators stem from heating of compressor-inlet air, inlet pressure drop, and mechanical losses. The Brayton cycle's sensitivity to temperature ratio makes the first effect predominant. The inlet-cooled alternator and bearings would liberate approximately 10% to 12% of the shaft power as electrical and windage losses, raising inlet temperature by an amount sufficient to decrease engine efficiency by 1 to 2 percentage points and power by 4 to 8%, depending on operating conditions. Combined with an inlet pressure drop estimated at roughly 1%, the net effect would be to reduce power by 7% and efficiency by 6% at nominal PSOFC design conditions.

Alternator selection: high-speed permanent magnetic vs. commercial low-speed generator

The versatile PowerWorks™ power take-off has been designed to adapt to either high-speed or 3600 rpm loads. The power-take-off shaft is in a cool region and supported by rugged conventional bearings. Either a shaft-speed permanent magnet alternator or a low-speed generator are adaptable to the PowerWorks™ engine. For high quality AC applications, the standard 2-pole commercial generator is the preferred choice. Lower cost and proven reliability are the dominating factors in grid-compatible AC power generation applications.

Compared to the rare-earth magnet alternators, the PowerWorks™ system with low-speed generator is more efficient on a total system basis. Table 7 compares electrical conversion efficiencies for the candidates.

Table 7 Comparison of Transmission Efficiencies

Component, efficiency	Shaft-speed alternator	PowerWorks™
bearings, turbine	98%	98%
alternator/generator, electrical	95%	95 % (including bearings)
bearings, alt/gen	98%	---
alt/gen windage	96%	>99%
inverter/rectifier	95%	
gearbox, mesh		>98.5%
Total conversion efficiency	83%	>90%

Note: In the PowerWorks™ drive train, there are only two bearings on the turbine shaft, and two supplied with the generator. There are no other bearings specifically associated with the “gearbox”. PowerWorks™ generator manufacturer’s data shows combined efficiency (electrical, bearings, windage, etc.) of 94.9% at 25 kWe, 95.4% at 50 kW-e, 94.7% at 75 kWe.

Either an induction generator or a synchronous generator may be used in the PowerWorks™ package. The induction generator has the advantage of low cost and the broadest utility acceptance. Ingersoll-Rand, one of the largest induction motor/generator purchasers, receives the competitive OEM price of about 20 to 25\$/kWe for this size induction motor. Efficiencies greater than 94% are guaranteed by the suppliers. Equally importantly, the reliability of this type of generator is well known and excellent. Data supporting a statistical mean time between forced outage of 318,300 hours has been compiled by GRI and Ingersoll-Rand from the various manufacturers.

The synchronous generator is mechanically connected to the PowerWorks™ package in exactly the same manner as the induction generator. Synchronous generators have the added benefit of stand-alone capability and on-site power-factor correction. This can be a compelling economic advantage to industry users who pay premiums to their local utility for substandard power factors. Coupled with an inverter system, required for the PSOFC, the synchronous generator could provide vital power-factor correction. The synchronous generator is also the preferred choice when “block-loading” occurs, a common stand-alone specification. In the integrated PSOFC application, this feature of the synchronous generator could improve the system response to abrupt load changes.

As a future product enhancement, a direct drive such as an air conditioning chiller or some other industrial load might be considered with the PowerWorks™ packaged system. This would further improve overall conversion efficiencies and net system pay-back.

Bearing selection

Over NREC’s 40-year history in the turbomachinery field, many types of gas-turbine engine bearings have been evaluated. For the PowerWorks™ product, a variety of

bearing configurations were analyzed including rolling contact, and journals employing air, refrigerant vapor, water, and oil.

Anti-friction rolling contact bearings are the most efficient, provided the DN (diameter x speed) rating is maintained at appropriate levels. Losses are 1/10th to 1/5th that of journal bearings. They have been reliably used for many years in gas turbines. With the maturity of a large well-developed statistical data base, the bearing life is accurately predicted. At the design conditions of the PowerWorks™ power turbine, angular contact ball bearings are the best choice, providing a life in excess of 80,000 hours at the extreme power condition of 105 kW (cold day).

Air journal bearings, not yet used in the gas turbine field, are best suited for ultra-clean environments within tightly-controlled temperatures. Other than some experimental gas turbines, their experience has been in cool environments on aircraft air-cycle machines. Air journal bearings also have the added limitations of higher windage losses and greater parasitic cooling losses as compared to oil journals. Their tighter tolerance components make these bearings more expensive than most other bearings.

Several types of oil journal bearings are used in the turbomachinery field. The principal attraction is the “zero wear” experienced as metal contact is isolated by a film of lubricant. Pre-lubrication from either the pump or a bladder-type accumulator minimizes starting wear. The floating-sleeve type, selected for the PowerWorks™, uses a free-floating ring between the static and rotating bearing surfaces. This modern bearing has lower losses than conventional sleeve-type bearings and provides improved stability. These bearings have become the standard on low cost turbochargers, costing only a few dollars to manufacture.

2.1.5 Input Data and Heat and Material Balance

We modeled our fuel cell/micro-turbine combined cycle process using the commercially available ASPEN Plus process simulation software package. The process flowsheet shown in Figure 1 along with the design criteria shown in Tables 2 and 5 were used to build the ASPEN simulation. ASPEN does not contain a standard unit operation for solid oxide fuel cells. MTI in collaboration with SOFCo had previously developed a proprietary model based on SOFCo FORTRAN subroutines. The proprietary model was fully integrated into the ASPEN simulation. The physical and thermodynamic property data used in our study came from ASPEN’s extensive and widely-recognized property database.

Detailed heat and material balances were performed on the completed process model. For ease of reference, we have summarized the ASPEN heat and mass balance results in Appendix A. The results are organized around the major components of the system.

2.1.6 Modeling Approach and Methodology

The overall approach used in this study was to use accepted chemical process engineering methodology. The overall process was modeled by using the commercially available ASPEN Plus process simulation software package. Major components were modeled rigorously. The fuel cell stacks were modeled within ASPEN using a rigorous FORTRAN model. The gas turbine (compressor and turbine) was modeled rigorously using the software package GATE/CYCLE. The results from the rigorous model were compiled into a curve fit, and the curve fit was incorporated into the ASPEN process simulation. The recuperator and fuel heater were also modeled separately, and the results incorporated into the simulation.

A simplified process flow diagram was generated for this report, as shown in Figure 2. Also included is a simplified stream summary in Table 1, and a component duty summary in Table 4.

A performance map was constructed using the method developed in Reference 4 to aid in analyzing the PSOFCs. Detailed stack models have been used to predict PSOFC response to changes in operating conditions. Integrated into a process model, the results yield accurate predictions of system performance. Results from a number of cases may be assembled to construct performance maps. While these models provide detailed resolution of processes and conditions, they are complex and cumbersome for operating point analysis and optimization. Since the performance of the system is highly dependent on the performance of the fuel cell stacks, a more useful approach is to define relationships that govern stack performance. These relationships are then combined to create a closed form parametric model suitable for application in the construction of performance maps and operating point optimization and analysis. The governing performance parameters for PSOFC stacks are fuel flow, area specific resistance, and operating voltage. The functional form of the model and the boundaries of the operating envelope provide useful insight into PSOFC operating characteristics and an improved means of guiding the selection of economically viable operating conditions. The performance map is discussed in Section 2.2.1.

Modeling of Engine Performance

Prediction of detailed engine performance has been carried out using the GateCycle code developed by Enter Software (Menlo Park, CA) in conjunction with NREC performance models. This approach relies on turbomachinery performance maps whose broad characteristics are simulated using NREC software, but which have been calibrated in detail based on component test data. One exception is the power turbine whose performance is entirely based on a model simulation, given that this component is to be redesigned and no test data are available.

Pressure losses at nominal design conditions are shown in Table 8, with approximate corrections applied at off-design flow conditions.

Table 8 Component Pressure Loss	
Component	$\Delta p/p$
Inlet filter/ducting	0.5%
Recuperator cold-side	1.0%
Recuperator hot-side	3.2%
Starting/preheat combustor	2.5%
Fuel-cell module	2.4%
Module inlet manifold	0.8%
Module exit manifold	0.5%
Fuel preheater	0.7%
Exhaust ducting	0.3%

Pressure loss for the PSOFC module was supplied by MTI based on laboratory measurements, plus a conservative allowance for ducting. Other component pressure losses were based on NREC measurements for existing hardware, applying estimated corrections where appropriate. These specifications formed the basis for duct sizing in the conceptual designs presented earlier.

For prescribed turbomachinery performance together with the specifications above, engine behavior is governed entirely by turbine-inlet and ambient temperature (TIT and T_{amb}). Engine output electrical power and flow conditions delivered to the PSOFC are summarized in Figures 11 through 13 over the anticipated range in these parameters.

Recuperator effectiveness does not enter directly into engine performance projections, but this parameter does govern the TIT that will be achieved at a given operating condition. For the proposed recuperator, a conservative design-point projection for air-side effectiveness exceeds 90%, defined by $\epsilon = (T_3 - T_2)/(T_{13} - T_2)$. This projection results in overall recuperator dimensions roughly equal to those of the current PowerWorks engine. Figure 14 shows the effect of recuperator hot-side inlet temperatures corresponding to the projections of Figures 11 through 13, based on the conservative effectiveness value of 90% (assumed constant throughout the operating range). From the plot it's evident that recuperator-inlet temperature is largely a function of TIT, because compressor-discharge temperature doesn't vary over wide limits.

Figures 11 through 14 demonstrate that TIT and T_{amb} , or more roughly the engine *temperature ratio* TIT/T_{amb} , exert substantial impact on engine output and flow. This means that sensitive control of PSOFC flowrate can be achieved by modulating the recuperator-inlet temperature (T_{13}), which depends in turn on the rate of fuel supply to the burner/reformer.

The detailed projections of Figures 15 through 18 were useful in characterizing engine performance for purposes of the MTI integrated system model. Here the engine is represented in simplified form as a single compressor and turbine each having prescribed efficiency, pressure ratio, and flowrate. To support application of this model for three

values of T_{amb} spanning the projected operating range (0F, 59F, 100F), the required inputs were generated as functions of TIT by performing detailed off-design simulations, yielding the results of Figures 15 through 18. The curve fits shown were then incorporated into the integrated system model.

Some minor simplifications were made in carrying out this procedure. A constant mechanical loss equal to 2% of overall turbine power, and a generator shaft-to-electrical efficiency of 94% were assumed. Although more accurate prescriptions are made in the full GATE/CYCLE off-design model, discrepancies are very small. A further modeling difference is that a single overall expansion efficiency was used for the two-turbine system, but this introduces no error with respect to the thermodynamic variables presented to the PSOFC system.

Figure 11 Exhaust Temperature

Figure 12 Compressor Flow

Figure 13 Engine Electrical Power Output

Figure 14 Hot Side Recuperator Inlet Temp.

Figure 15 Compressor Pressure Ratio

Compressor Flow

Figure 16 Compressor Flow

Figure 17 Compressor Efficiency

Figure18 Overall Expansion Efficiency

2.2 Process/Equipment Uncertainties and Development Requirements

Having chosen the design flow conditions for the PSOFC power plant that correspond with the capacity of the PowerWorks™ 70kW engine, most production engine components can be used directly. In particular, no changes are anticipated for the compressor or high-pressure turbine and associated ducting. Starting, lubrication, and fuel-delivery systems will also remain unchanged. Redesign of the combustor and waste-heat recovery system (now fuel preheater) will be relatively minor, with the overall dimensions and packaging for these components unaffected (see below).

The power turbine is the single exception to the application of production hardware. Although the turbine rotor can be used without modification, the stationary housing will require rework in order to accommodate the increased pressure loss introduced by the PSOFC system. The impact of this pressure loss would be reduced pressure-ratio across the gasifier turbine, limiting engine power and flow especially on hot days. The change to the power turbine housing will increase the power-turbine flow capacity.

Redesign of the combustor is dictated by the need to minimize pressure loss at design conditions, where the combustor is switched off and represents a parasitic loss. Combustor inlet temperature is approximately 1200F at nominal PowerWorks™ design conditions, but for the proposed application this value increases to 1600F. Relatively straightforward design changes should make it possible to hold pressure loss within the 2.5% design specification, which is required for stable combustor performance during starting and preheat. Cold-flow testing of the redesigned combustor is planned in order to meet these objectives.

Adaptation of the current PowerWorks™ heat-recovery system to meet fuel preheating demands is largely a matter of substituting a redesigned heat-exchanger core for the current commercial finned-tube unit. The present device supplies hot water at a constant temperature, as achieved by pivoting the core out of the flow path under thermostatic control. For the proposed application this pivoting feature is probably unnecessary, because the fuel demand does not vary widely and precise control of the delivery temperature is less critical.

2.2.1 Fuel Cell Issues

The power and efficiency realized by operation of Solid Oxide Fuel Cell (PSOFC) systems are determined not only by the stack characteristics, but also by the operating conditions. The choice of operating conditions enable a wide range of delivered power and efficiency from a particular PSOFC device. Operating parameters may be selected to maximize power while constraining efficiency, maximize efficiency while constraining power, or optimization of a function of both variables such as cost of electricity. While detailed stack models may be used to predict PSOFC response at a specific operating point, these complex models are unnecessarily cumbersome for operating point analysis and optimization. Relationships between fuel flow, area specific resistance, and operating voltage were defined to develop a closed form parametric model. This model

was applied to construct a performance map for operating point analysis and economic optimization. The functional form of the model and the boundaries of the operating envelope provide useful insight into PSOFC operating characteristics and a simple means of selecting operating conditions.

The choice of stack temperature, fuel flow, and operating voltage enable a wide range of delivered power and efficiency from a particular PSOFC device. Thus, the challenge of formulating a relationship between PSOFC power and efficiency in terms of controllable operating parameters is undertaken. It has been shown that the stack efficiency can be expressed simply in terms of operating voltage, fuel utilization, and thermoneutral voltage (Reference 1).

$$h_{stack} = \frac{V_{op}U}{E_{tn}} \quad (1)$$

h_{stack} = fuel cell stack efficiency

V_{op} = operating voltage

U = fuel utilization

E_{tn} = thermoneutral voltage

Of these three factors, operating voltage is the only independent variable.

Thermoneutral voltage is a property of the fuel defined as the heating value of the fuel divided by the number of Faradays of charge resulting from complete electrochemical oxidation of the fuel.

$$E_{th} = \frac{\Delta H}{nF} \quad (2)$$

where:

ΔH = molar heating value of the fuel (LHV) for methane 890.347 kJ/g - mole

n = number of electrons transferred per molecule of fuel to completely oxidize the fuel, 8 for methane

F = Faraday's number, 96,487 Coulombs per g - mole

Utilization, defined as the ratio of delivered current to stoichiometric current, is an outcome which is dependent on stack resistance, operating voltage and fuel flow rate.

$$U = \frac{I}{\dot{n}nF} \quad (3)$$

where:

I = delivered current

$\dot{n}$ = fuel flow rate

n = number of electrons transferred per molecule of fuel to completely oxidize the fuel,

8 for methane

F = Faraday's number, 96,487 Coulombs per g - mole

For a given stack resistance and fuel composition, the only independent variables available to effect changes in power and efficiency are operating voltage and fuel flow rate. Stack resistance is of course a function of operating temperature. However, operating temperature is not considered as an independent variable here as it is assumed that to minimize resistance, the stack is operated at the highest temperature consistent with stack life and system balance of plant constraints.

It would appear from equation [1] that a high operating voltage is required for high efficiency. At a fixed fuel flow rate however, utilization declines with increasing operating voltage. Power also decreases as operating voltage is raised from the maximum power voltage which in common experience is slightly less than half the open circuit voltage (OCV). The greatest efficiency at any given fuel flow is obtained at the operating voltage which results in the highest power output.

Detailed system models were used in this program to predict overall system performance for a range of parameters. While these models provide detailed resolution of processes and conditions in the stack or balance of plant, they are unnecessarily complex and cumbersome for operating point analysis and optimization. A more useful approach is to define relationships between fuel flow, area specific resistance, and operating voltage. These relationships can then be combined to create a closed form parametric model suitable for application in the construction of performance maps and operating point optimization and analysis. The functional form of the model and the boundaries of the operating envelope, provide useful insight into PSOFC operating characteristics and an improved means of selecting operating conditions. Since the PSOFC is the highest efficiency and highest cost components the results of the PSOFC performance map translate to performance for the entire system.

A useful metric is derived by casting fuel flow rate as an electrochemical term which represents the average current density required for 100% fuel utilization.

$$j_f = \frac{\dot{n}nF}{A} \quad (4)$$

j_f = average current required for 100% fuel utilization divided by cell area

A = cell area

The driving voltage, or difference between the reversible potential and operating voltage ($E_m - V_{op}$), required to sustain the full utilization current density can then be calculated using the stack area specific resistance (ASR, or R'').

$$\Delta V_f = j_f R'' \quad (5)$$

ΔV_f = driving voltage

j_f = average current required for 100% fuel utilization divided by cell area

R'' = Area Specific Resistance or ASR, (Ω - cm^2)

Combining equation 4 with equation 5 gives

$$\Delta V_f = \frac{\dot{n}nFR''}{A} \quad (6)$$

ΔV_f = driving voltage

$\dot{n}$ = fuel flow rate

n = number of electrons transferred per molecule of fuel to completely oxidize the fuel, 8 for methane

F = Faraday's number, 96,487 Coulombs per g - mole

R'' = Area Specific Resistance or ASR, (Ω - cm^2)

A = cell area (cm^2)

Assuming that ASR is constant for a particular stack or set of stacks, and that n is a constant for a particular fuel, then the driving voltage is dependant on fuel flow rate and the number of cells (or cell area). Further details of the operating point model may be found in Reference 4.

The performance map will show that optimizing for maximum efficiency will lead to a large stack size and thus a large capital investment. In practice, objectives may dictate selection of an operating point designed to maximize power while constraining efficiency, maximize efficiency while constraining power, or optimize a function of both power and efficiency such as cost of electricity (COE). The relative priorities of efficiency, power, and COE establish the appropriate objectives and constraints.

A map of PSOFC performance in (V_{op} , ΔV_f) space was created using the approach developed in Reference 4. These results are for a plug flow PSOFC configuration (e.g., cross flow or counter flow) under isothermal conditions at 800 °C. Inlet fuel composition was methane with 2.0 steam to carbon ratio. The RAS was specified at 0.5 $\Sigma\text{-cm}^2$. The utilization parameter was varied from 0.3 to 0.98, while the ΔV_f parameter (representative of fuel flow and cell area) varied from 0.095 to 0.7 volts. The map of efficiency, power density, current density and utilization in the V_{op} , ΔV_f plane is shown in Figure. The domain is bounded to the right (high operating voltage) by utilization falling below 30% and to the left (low operating voltage) by utilization exceeding 98%.

Several interesting observations can be made by a thorough examination of this mapping. There are two operating regimes demarked by the kink in the high utilization boundary. At values of ΔV_f less than about 0.2 volts, power is fuel flow limited, while above 0.2

volts ΔV_f , power is resistance limited. The location of the transition is a function of ASR. Power drops by about half from the point of maximum power to the point of maximum efficiency. Efficiency drops by a similar ratio between the high efficiency and high power points. At a constant fuel flow (lines of constant ΔV_f), the operating voltage for maximum efficiency and maximum power coincide. Along a line of constant efficiency, maximum power is achieved at the greatest fuel flow (ΔV_f) possible.

For the PSOFC/Micro-Turbine combined cycle with a net system efficiency of greater than 70%, the operating point for the fuel cell is at an efficiency of 68%. As is shown in Figure, this shifts the operating point to the extreme lower right corner of the performance map. The lower right corner of the performance map corresponds to a high V_{op} and a low ΔV_f . Operating at high V_{op} results in low power density for a given stack. For a given fuel flow and stack ASR, operating at low ΔV_f requires a large stack area, as is shown by inspection of Equation 6. A large stack area requires many fuel cells and fuel cell stacks. Until PSOFCs become the low cost component in the PSOFC/Micro-Turbine cycle, optimizing the system economically will result in a system that has less fuel cell area than the system that is optimized on efficiency. The 18 MW system described above, would require over 18,000 square meters of cell area. Clearly for an economically viable market introduction, the cell area must be reduced.

A judiciously selected operating point is essential to extracting the most value from a PSOFC installation. The closed form parametric model presented in Reference 4 was used to create a performance map to aid the process of understanding economical operating point selection. The form of the model and the boundaries of the performance maps also provide insight valuable in the selection process. A commonly stated goal of high operating voltage (e.g., 0.7-0.8V) is not expected to be the highest efficiency point except at relatively low fuel flows and high cell area. It can be shown, by inspection of Figure 15, that curves of efficiency are highest in the lower right corner, and curves of power density are highest in the upper left corner. Therefore, commonly quoted goals for high efficiencies and high power densities are not likely to be achieved simultaneously.

Figure19: PSOFC Performance Map

2.3 System Capital Costs – Nth of a Kind Projections

Based upon a 727kW sized system described in Section 2, two primary subsystems were analyzed with regard to both existing and nth of a kind cost estimates. Using 50 units as the nth of a kind volume for both the microturbine subsystem and for the fuel cell related subsystem, estimates include capital costs, variable operating costs and replacement costs.

Using MTI's proprietary "Upgrades and Enhancements Financial Evaluation Program" software and incorporating a industry recognized standards for Cost of Electricity (COE) calculation, a wide variety of plant equipment, operations, and variable cost simulations were conducted to evaluate different scenarios for system configuration. Using the system described within Section 2, component costs were compiled and nth of a kind costs were estimated based upon; volume manufacturing cost reductions, technology innovations leading to lower material, process and manufacturing costs, and performance enhancement to component and system operations leading to lower cost components.

Initial installation nth of a kind capital costs (including a 10% installation charge) were \$454,850 for the 727kW system leading to a \$625/kW capital cost. Nth of a kind microturbine subsystem costs were \$85,000 and fuel cell subsystem costs accounted for the remaining \$328,500 (excluding the 10% installation charges in both cases).

In addition to the initial capital costs, a conservative approach to one-time fixed costs (assuming non-depreciable capital costs) was utilized to replace both subsystems (in their entirety) at the end of each subsystems assumed lifetime (multiple times over the 25 year plant life). Using an 80,000 hour lifetime for the microturbine subsystem and a 40,000 hour assumed lifetime for the fuel cell subsystem, the following fixed cost capital replacement schedule represents overall capital costs over the 25 year plant life.

Table 9 PSOFC/Microturbine Capital Costs		
ITEM	YEAR	COST
Initial Capital Cost (Fuel Cell/ Microturbine) Subsystems with 10% Installation Factor	1998	\$454,850
Fuel Cell System Replacement	2003	\$361,350
Fuel Cell System Replacement & Microturbine System Replacement	2008	\$454,850
Fuel Cell System Replacement	2013	\$361,350
Fuel Cell System Replacement & Microturbine System Replacement	2018	\$682,275*

*Includes pro-rated capital costs for subsystems whose lifetimes do not exactly fit into the 25 year plant life period assumed in the case.

2.4 ANNUAL OPERATING COSTS

A number of different assumptions were employed to estimate nth of a kind operating costs for the system. These include:

- 70% efficiency for the system,
- Capacity Factor: 50/52 Week Operating @ 92% Capacity = 88.4% overall capacity (with 56.33MWh/Year electric generation),
- Cost of Capital at 15%,
- Cost Basis of mid-1998 U.S. Dollars,
- COE Evaluation Method using Constant Dollars,
- Cost of Natural Gas at \$3.00 MM Btu (HHV),
- 1/3 man year of operator labor (at a total cost of \$93,951/year),
- \$36,000/year maintenance,
- Water usage costs of \$1.54/day.

Using the MTI's "Upgrades and Enhancements Financial Evaluation Program" software to determine COE, a total COE (including Levelized Capital, Variable Cost and Fixed Cost is 5824 mills/kWh or 5.824 ¢/kWh.

2.5 OPPORTUNITIES FOR IMPROVEMENT AND SUGGESTED WORK

The key to successful development of this technology is reduced component costs and higher fuel cell power output. Reducing the component cost and increasing cell power output is the objective of the SOFCo fuel cell development program. Other key issues to be resolved involve the integration of the major components and their controls into a functional package. We believe the best way to demonstrate the technology while minimizing costs and reducing uncertainty is to develop a proof of concept demonstration.

2.5.1 Proof of Concept Demonstration – 180 kW System

A proof of concept that includes all components of the full-scale system can be demonstrated at 180 kW. The demonstration would include a full-scale engine. The recuperator would be slightly smaller than full scale as the recuperator hot side inlet temperature is higher. The fuel heater would also be slightly smaller than full scale as the fuel side flows are lower. Both the recuperator and fuel heater would require a moderate increase in surface area to achieve full scale. The major departure from full scale is that only one fuel cell module would be included. The module can be demonstrated at full scale size, or if full scale stacks are not available for demonstration, the stacks could also be simulated. At the successful conclusion of the proof of concept demonstration, all that would be needed to field a full-scale system would be the inclusion of additional fuel cell modules and ducting.

The process description of the 180 kW proof of concept demonstration, shown in Figure 23, is similar to the full-scale system described in section 2.1.1. The air is

compressed in the compressor at a 3:1 pressure ratio. The air is then heated to 1600°F in a high temperature recuperator by utilizing exhaust gas from the CpnTM module. The hot, high-pressure air is then expanded through the turbine providing power for the compressor and electrical generation. The turbine produces 69.1 kWe of net electrical power or 39% of the total. The air is then sent to the fuel cell.

Natural gas is mixed with steam that was generated in the steam generator coil, the mixture is then heated further in the fuel heater. The heated fuel/steam mixture is then sent to the steam reformer. In the steam reformer, the fuel-steam mixture passes over steam reforming catalyst and is processed into hydrogen rich reformat and sent to the fuel cell.

The hydrogen and carbon monoxide in the reformat is electrochemically oxidized in the fuel cell producing electrical power. The fuel cell produces 109 kW of electrical power or 61.2% of the total. The unreacted fuel is burned with the fuel cell cooling air in the fuel cell module enclosure, further boosting the exhaust temperature and providing heat to drive the steam reforming reactions in the steam reformer. The hot gas leaving the CpnTM is then sent to the high temperature recuperator where it heats the compressed air, and then is sent to the fuel heater where it heats the fuel and steam mixture. The fuel heater exhaust is used to generate steam to that is mixed with the natural gas. The exhaust exits the process at 259°C. The exhaust could be used to generate low-pressure process steam or space heating in a cogeneration heat exchanger.

The state parameters for the streams shown in 180 kW process schematic are listed in Table10. The 180 kW system design parameters are listed in Table11, and the component duties are summarized in Table 13. The performance for the 180 kW system is listed in Table 12.

Figure 20: 180 kW PSOFC/MicroTurbine System

Table 10 - State Parameters For 180 kW Fuel Cell/Micro-turbine Combined Cycle				
State Point	Flow kg/s (lbm/hr)	Temperature C (F)	Pressure kPa (psi)	Enthalpy J/kg (Btu/lbm)
1	0.662 (5256)	15 (59)	101.3 (14.7)	-1.35e5 (-58.04)
2	0.622 (5256)	178 (352)	304 (44.1)	3.11e4 (13.35)
3	0.662 (5256)	875 (1606)	304 (44.1)	8.05e5 (346.26)
4	0.662 (5256)	875 (1606)	304 (44.1)	8.05e5 (346.26)
5	0.622 (5256)	642 (1187)	106.9 (15.5)	5.35e5 (228.83)
6	0.123 (975.7)	862 (1583)	106.8 (15.5)	7.83e5 (336.50)
7	0.009 (69.5)	15 (59)	204.8 (29.7)	-4.74e6 (-2037.08)
8	0.014 (109.3)	25 (77)	120 (17.4)	-1.60e7 (-6897.80)
9	0.014 (109.3)	108 (226)	120 (17.4)	-1.33e7 (-5706.12)
10	0.023 (178.8)	93 (200)	120 (17.4)	-9.95e6 (-4279.84)
11	0.023 (178.8)	252 (485)	120 (17.4)	-9.56e6 (-4109.37)
12	0.023 (178.8)	816 (1500)	120 (17.4)	-3.16e6 (-1359.78)
13	0.036 (286.8)	862 (1583)	120 (17.4)	-7.57e6 (-3254.09)
14	0.685 (5434.8)	951 (1745)	106.8 (15.5)	1.33e4 (14.33)
15	0.685 (5434.8)	950 (1743)	106.8 (15.5)	3.19e4 (13.71)
16	0.685 (5434.8)	321 (610)	106.8 (15.5)	-7.17e5 (-308.25)
17	0.685 (5434.8)	309 (589)	106.8 (15.5)	-7.30e5 (-313.86)
18	0.685 (5434.8)	259 (498)	106.8 (15.5)	-7.86e5 (-337.82)

Table 11 - Design Parameters for 180 kW Fuel Cell/Micro-Turbine Combined Cycle	
Fuel	Natural gas 0.96 CH ₄ , 0.02 N ₂ , 0.02 CO ₂ LHV = 4.81E7 J/kg (20,659 Btu/lbm)
Turbine pressure ratio	3:1
Recuperator effectiveness	0.902
Fuel cell	976 cells
Operating voltage	0.70 V/cell
System heat loss	0.5% of heat input
Inverter efficiency	95%
Generator efficiency	98%
Gear box loss	5%

Table 12 - Performance Summary for 180 kW Fuel Cell/Micro-Turbine Combined Cycle	
Mass flow rate of natural gas	31.5 kg/hr (69.5 lbm/hr)
Gas flow * LHV	420,972 W (1,435,801 Btu/hr)
Gross Power	184.3 kW
Net Power	178.5 kW
Efficiency, LHV	44.6
Contributions to Power	
Fuel cell	109.2 kW
Turbine	69.3 kW
Inverter loss	-5.8 kW
Generator loss	
Gear box loss	

Table 13 - Component Duty Summary for 180 kW Fuel Cell/Micro-Turbine Combined Cycle	
Component	Duty (kW)
Fuel heater	8.9 (30,477 Btu/hr)
Reformer	79.9 (272,723 Btu/hr)
Recuperator	512.8 (1,749,778 Btu/hr)
Spent fuel burner	-79.9 (-272,723 Btu/hr)
Steam generator	38.2 (130,233 Btu/hr)

3 CONCLUSION

This study has demonstrated that the unique approach taken to combining a fuel cell and gas turbine has both technical and economic merit. By using a micro-turbine, and a non-pressurized fuel cell the total system size has been reduced substantially from those presented in other studies, while maintaining over 70% efficiency. The reduced system size can be particularly attractive in the deregulated electrical generation/distribution environment where the market may not demand multi-megawatt central stations systems. The small size also opens up the niche markets to this high efficiency, low emission electrical generation option.

While the study has discovered no technical obstacles to success, a sub-scale technology demonstration would reduce the risk of performance and enable a full-scale commercial offering. Demonstrating a full size micro-turbine, with a single fuel cell module would prove the concept as well as the major components and BOP that would be needed in a full-scale system.

The major hurdle to the commercialization of the technology is economics. The costs of major components must be reduced from the present levels. For some components, such as the micro-turbine, volume production may be enough to reduce costs. For the fuel cells, low cost manufacturing, as well as volume production are need to achieve cost projections.

4 REFERENCES

1. J. Hartvigsen, S. Elangovan and A. Khandkar, in *Solid Oxide Fuel Cells*, S. C. Singhal and H. Iwahara, Editors, PV 93-4, p. 878, The Electrochemical Society Proceedings Series, Pennington, NJ (1993).
2. J. Hartvigsen, S. Elangovan and A. Khandkar, in *Science and Technology of Zirconia V*, S. P. S. Badwal, M. J. Bannister, R. H. J. Hannink, Editors, p. 682, Technomic Publishing Company Inc., Lancaster, PA (1993).
3. R. Herbin, J. M. Fiard and J. R. Ferguson, in First European Solid Oxide Fuel Cell Forum Proceedings, U. Bossel Editor, p317, Lucerne, Switzerland (1994).
4. J. Hartvigsen, S. Elangovan and A. Khandkar, in Third European Solid Oxide Fuel Cell Forum Proceedings, P. Stevens Editor, p517, Nantes, France (1998).

5 APPENDICES

A. ECONOMIC ANALYSIS

Babcock & Wilcox a McDermott company	Upgrades and Enhancements Financial Evaluation Program
1. PROJECT INFORMATION SUMMARY	
Project Title DOE PSOFC/MIDROTURBINE COMBINED CYCLE (727 kW)	
Case No. <div></div>	
Customer Data:	
Company: <div></div>	
Plant Name: <div></div>	
Unit No(s).: <div></div>	
Contact Name: <div></div>	
Contact Phone: <div></div>	
Contact Fax: <div></div>	
Contact Electronic Mail Address: <div></div>	
Contact Mailing Address and Other Information: <div></div> <div></div> <div></div> <div></div>	
Description The analysis is for a 727 kW PSOFC/Microturbine combined Cycle Power Plant that utilizes approximately 650kW of PSOFCs and 70 kW of Microturbine generation to achieve a 70% efficiency rating. Plant operations assume a 50/52 week operation (with a planned 2 week service outage and a 92% capacity factor. Capacity factor input into the B&W analysis model utilizes a compound capacity factor rate (50/52*92) to achieve an overall capacity rate of 88.46% A 25 year total plant live includes stack replacement every 5 years and microturbine replacemnt every 8.33 years. Ancillary equipment replacemnet is included in O&M costs.	
Prepared by <div></div>	Date <div></div>

2. UNIT CAPACITY AND OPERATING INFORMATION SUMMARY

Project Title: DOE PSOFC/MIDROTURBINE COMBINED CYCLE (727 kW) Case No.:
--

HEAT RATE DATA

Annual Average Heat Rate

Base Case (Before this installation)	0	Btu / kWh
Upgrade (After this installation)	4,875	Btu / kWh
Change (Upgrade - Base Case)	4875	Btu / kWh

CAPACITY DATA

Net Unit Capacity	Base	0	MWe
	Upgrade	0.727	MWe
	Change	0.727	MWe

Capacity Factor	Base	0.88	%
	Upgrade	0.88	%
	Change	0.00	%

Net Annual Generation Rate	Base	0	MWh / Year
	Upgrade	56.33691	MWh / Year
	Change	56	MWh / Year

Prepared by	Date
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3. CAPITAL EQUIPMENT COSTS - B&W and Vendors

Project Title: DOE PSOFC/MIDROTURBINE COMBINED CYCLE (727 kW)

Case No.:

Item No.	Description <small>[Note: If the material / installation breakout is not available enter as material]</small>	Cost		
		Material [\$ US]	Installation [\$ US]	Total [\$ US]
1		\$413,500	\$41,350	\$454,850
2		\$0	\$0	\$0
3		\$0	\$0	\$0
4		\$0	\$0	\$0
5		\$0	\$0	\$0
		\$413,500	\$41,350	\$454,850

Schedule

Project Conclusion
[In-Service Year] 1998

Project Duration
[Start to In-Service] 1 Months

Notes:

1) Project Duration refers to the period during which the customer will incur costs due to carrying the costs for work in progress during construction. Adjustments for these costs are made as the allowance for funds during construction in the Form 7, Total Capital Cost. If these costs are included in the project capital price entered above then enter 0 months.

2) If significant one-time costs, such as replacement catalysts, are required in out years (after the in-service year) those are addressed in Form 5, Fixed Operating Costs.

3) If a one-time upgrade can be expensed, it may be treated as a one-time- fixed cost in the year prior to the in-service year and entered as a one-time- fixed cost in that year in Form 5, Fixed Operating Costs.

4) Avoided capital costs may be treated as a negative value here or as an annual credit in Form 5, Fixed Operating Costs.

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Date

4. VARIABLE OPERATING COSTS REVENUES AND BENEFITS

Project Title: DOE PSOFC/MIDROTURBINE COMBINED CYCLE (727 kW)
Case No.:

FUEL COSTS

Base Case
(Before this Installation)

Fuel Type

Delivered Cost / Million Btu

Annual Escalation %

Full-Load Fuel Flow 0 Million Btu/hr

Fuel Cost \$0 / hr

Upgrade
(After this Installation)

Fuel Type

Delivered Cost / Million Btu

Annual Escalation %

Full-Load Fuel Flow 3.544125 Million Btu/hr

Fuel Cost \$11 / hr

Change
(Upgrade - Base Case)

Delivered Cost 1.7 / Million Btu

Fuel Cost \$11 / hr

UNBURNED CARBON LOSS (UBC) In Ash
[Expressed as % Carbon in Ash]

DO NOT USE IF UBC changes were
included in changes in the heat rate.

UBC
Base Case %

UBC
Upgrade %

UBC
Change
(Upgrade - Base Case) 0.00 %

Base Fuel Characteristics

Upgrade Fuel Characteristics

Higher Heating Value

Btu / lb

Ash Content

%

Higher Heating Value

Btu / lb

Ash Content

%

UBC Cost \$0 / hr

WASTE DISPOSAL - INCREASES IN QUANTITY

	Increase in Amount Generated	Unit Cost of Disposal	Cost / hr
Ash	<input type="text" value="0"/> Tons / hr	<input type="text" value="\$9.00"/> / Ton	\$0
Sludge	<input type="text" value="0"/> Tons / hr	<input type="text" value="\$12.00"/> / Ton	\$0
Other	<input type="text" value="0"/> Tons / hr	<input type="text" value="\$9.00"/> / Ton	\$0
Waste Disposal (Increase in quantity) Cost -Subtotal			\$0

WASTE DISPOSAL - INCREASES IN UNIT (\$ / Ton) DISPOSAL COSTS

	Total Amount of Waste Generated	Increase in Unit Cost of Disposal	Cost / hr
Ash	<input type="text" value="0"/> Tons / hr	<input type="text" value="\$0.00"/> / Ton	\$0
Sludge	<input type="text" value="0"/> Tons / hr	<input type="text" value="\$0.00"/> / Ton	\$0
Other	<input type="text" value="0"/> Tons / hr	<input type="text" value="\$0.00"/> / Ton	\$0
Waste Disposal (Increases in unit disposal price) Cost - Subtotal			\$0
Overall Change in Waste Disposal Cost -Subtotal			\$0

CONSUMABLES - INCREASES IN QUANTITY

	Increase in Amount Used	Cost / Unit Used	Cost / hr
Raw Water	<input type="text" value="0.08"/> 1000 Gals / hr	<input type="text" value="\$8.00"/> / 1000 Gals	\$1
Process Steam Consumed	<input type="text" value="0.00"/> 1000 Lbs / hr	<input type="text" value="\$3.50"/> / 1000 Lbs	\$0
Limestone	<input type="text" value="0.00"/> Tons / hr	<input type="text" value="\$18.00"/> / Ton	\$0
Other	----->		<input type="text" value="\$0"/>
Consumables Cost - Subtotal			\$1

CREDITS FOR BYPRODUCTS

	Increase in Amount Produced	Credit / Unit	Credit \$/ hr
Steam	<input type="text" value="0.00"/> 1000 Lbs / hr	<input type="text" value="\$0.00"/> / 1000 Lbs	\$0
Ash	<input type="text" value="0.00"/> Tons / hr	<input type="text" value="\$0.00"/> Tons	\$0
Gypsum	<input type="text" value="0.00"/> Tons / hr	<input type="text" value="\$0.00"/> Tons	\$0
Other	-----> <input type="text" value="\$0"/>		
Total Byproducts Credits [Negative cost]			\$0

ELECTRICITY PRICING

Electric Power Price / kWhr

Annual Escalation of Base Price %

Incremental Price (Power Price -(Fuel Cost / kWh x Variable Cost Factor)) \$0.0000 / kWh

Variable Cost Factor %
(Non-fuel variable cost shown as percentage of fuel cost)

NET INCREMENTAL REVENUE- Excluding Fuel Cost (Credit From Incremental Net Power Produced)

Incremental Power	0.727 MW	Net Incremental Revenue Except Fuel [Negative cost]	\$0 /hr
-------------------	---	---	--

Incremental power produced is automatically calculated from changes in capacity entered in the Unit Capacity and Operating Information. Increased production due to capacity factor changes are also included in the total Variable Operating Costs used in the Cash Flow and Revenue Requirements.

NET VARIABLE COSTS LESS BENEFITS AND LESS INCREMENTAL REVENUE

[Note that incremental revenue considered here is limited to that due to capacity increases only]

Net Variable Costs \$12 /hr

NET VARIABLE COSTS (COST LESS BENEFITS) EXCLUDING FUEL COSTS AND INCREMENTAL REVENUE

[Note that UBC & DGL costs and cost arising from changes in superheater and reheater temperatures and attemperator flow rates are treated under fuel cost considerations and are not included here]

Net Variable Costs (with adjustments noted above) \$1 /hr

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5. FIXED OPERATING COSTS

Project Title: DOE PSOFC/MIDROTURBINE COMBINED CYCLE (727 kW)

Case No.:

Sign Convention: Cost increases arising from this upgrade or enhancement are positive values (+), and cost reductions that result from this upgrade or enhancement are negative (-) values.

ANNUAL FIXED COSTS

OPERATING LABOR

Shift Head- Count Increase [8760 hrs / year]	Labor Cost	Supervisory, Overhead and Gen & Admin Cost	Labor + Supervisory Cost (Increase) Subtotal
<input type="text" value="0.33"/>	\$72,270 / Year	\$21,681 / Year	\$93,951 / Year
Labor Rate	<input type="text" value="\$25.00"/>	Supervisory Cost in Percent Labor Cost	<input type="text" value="30.00"/> %

MAINTENANCE

Labor Cost Increase	Materials Cost Increase	Labor + Materials Cost Subtotal
<input type="text" value="\$36,000"/> / Year	<input type="text" value="\$0"/> / Year	\$36,000 / Year
[Includes Overhead & G&A costs]		

PARTS INVENTORY

Inventory Increase	Annual Rate	Inventory Carrying Cost Increase	Administrative Cost Increase	Cost Subtotal
<input type="text" value="\$0"/>	<input type="text" value="20.00"/> %	\$0 / Year	<input type="text" value="\$0"/> / Year	\$0 / Year

STARTUP FUEL COSTS

Contact B&W Service Company for Input

 / Year

SUBTOTAL - Annual Cost (Increases) \$129,951 / Year

ONE-TIME (Non-Depreciable) FIXED COSTS - At time of performance

Sign Convention: One-time fixed costs related to this upgrade or enhancement are positive values (+),

Replacement Power Cost During Outage for Installation of Upgrade or Enhancement Net of Fuel
{Fuel Cost from VOC = \$ 0 /Hr form VOC}

Outage Duration	Replacement Power Cost	Capacity	Load Factor	Cost
0 days	\$0.060 /kWhr.	0 MW	80 %	\$0

OTHER ONE-TIME FIXED COSTS (Expensed Items that are not Depreciated - Example catalysts replacement) Note: these can be initial project expenses that are not capital expenses. If so, the year should be "year 0" -- the year preceding the "in-service year."

Item	Description	Year	Cost
1	Stack Replacement	2003	\$361,350
2	Stack Replacement/Microturbine Replacement	2008	\$454,850
3	Stack Replacement	2013	\$361,350
4	Stack Replacement/Microturbine Replacement	2018	\$682,275

Sign Convention: Added revenues arising from this upgrade or enhancement are positive values (+), and reduced revenues that result from this upgrade or enhancement are negative (-) values.

ANNUAL CREDITS for Premium on Power Sales and Other Benefits

Contact B&W Service Company for Input

System \$0 / Year

Plant \$0 / Year

SUBTOTAL \$0 / Year

TOTAL OF ANNUAL FIXED COST INCREASES Less ANNUAL CREDITS

Does not include "AVOIDED CAPITAL COST CREDIT SHOW BELOW."

TOTAL **\$129,951 / Year**

AVOIDED CAPITAL COST CREDIT

NOTE: Avoided capital costs may also be treated as a negative capital cost in Form 3. DO NOT APPLY THIS CREDIT IF A NEGATIVE COST WAS USED FOR THE SAME AVOIDED COST IN FORM 3.

USE CREDIT IN THIS SHEET?

ENTER "Y" for yes or "N" for no, here--->

N

Incremental Capacity

Avoided Capital Cost

Capital
Levelization
Factor

1

MWe

\$350 /kW

0.2381

ANNUAL CREDIT

\$0 / Year

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6. FINANCIAL PARAMETERS

Project Title: DOE PSOFC/MIDROTURBINE COMBINED CYCLE (727 kW)

Case No.:
Evaluation Period

Starting Date (In-service year) 1998

Evaluation Period (Typically book life) 25 Years

Tax Life

Years (with 1/2 year convention) 20

Book Life

Years 25

Income Tax Rates

Federal 34.000 %

State 4.000 %

Property Tax and Insurance as Percent of Capital Equipment Cost

2.000 %

Inflation Rate

4.000 %

Optional User Input Start Date If there is no depreciable capital cost

Tax Depreciation Rate Table

Year	Rate
1	3.750
2	7.219
3	6.677
4	6.177
5	5.713
6	5.285
7	4.888
8	4.522
9	4.462
10	4.461
11	4.462
12	4.461
13	4.462
14	4.461
15	4.462
16	4.461
17	4.462
18	4.461
19	4.462
20	4.461
21	2.231
22	0.000
23	0.000
24	0.000
25	0.000
26	0.000
27	0.000
28	0.000
29	0.000
30	0.000

Total 100.000

Discount Rate Calculation

Type of Security	Percent of Total Financing	Rate	Annual Return After Tax	Annual Return Before Tax
Debt	46.000 %	15.439 %	4.403 %	7.102 %
Preferred	8.000 %	14.939 %	1.195 %	1.195 %
Common	46.000 %	20.439 %	9.402 %	9.402 %
Discount Rate (after tax)			15 %	
<div> <div>Prepared by</div> <div>0</div> <div>Date</div> <div>0</div> </div>				

7. TOTAL CAPITAL COST

Project Title: DOE PSOFC/MIDROTURBINE COMBINED CYCLE (727 kW)
Case No.:

A	Total Material and Installation Capital			\$454,850
B	General Facilities	<input type="text" value="0.00"/>	% of A	\$0
C	A&E, Owner Engineering, Other Owner Costs	<input type="text" value="0.00"/>	% of A	\$0
D	Project Contingency	<input type="text" value="0.00"/>	% of A + B + C	\$0
E	Total Plant Cost		A + B + C + D	\$454,850
F	Allowance for Funds During Construction	0.000	% of E	\$0
	Project Duration (Years)	0.08		
G	Total Plant Investment		E + F	\$454,850
H	Pre-Production / Startup Cost	<input type="text" value="0.00"/>	% of G	\$0
I	Inventory			<input type="text" value="\$0"/>
J	Initial Catalyst and Chemicals			<input type="text" value="\$0"/>
K	Total Capital Requirement		G + H + I + J	\$454,850

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8. ECONOMIC EVALUATION SUMMARY

Project Title: DOE PSOFC/MICROTURBINE COMBINED CYCLE (727 kW)**Case No.:****CAPITAL COST SUMMARY (at Time of Performance)**

Total Depreciable Material and Installation Capital	\$454,850
Total Depreciable Capital Requirement	\$454,850
Total One-Time (Non-Depreciable) Fixed Costs	\$1,859,825

ANNUAL COSTS

	\$ / Year	mills/kWh
Levelized Capital (Depreciable)	\$108,313	1922.58927
Variable Cost (Benefit)	\$954	16.93280
Fixed Cost (Benefit)	\$218,830	3884.31285
Total	\$328,097	5823.83492

FINANCIAL RESULTS

Net Present Value at the Discount Rate of	15 %	(\$1,337,262)
Internal Rate of Return	Standard Calculation	ERR
	Alternate Calculation	10.00%

***If Standard Calculation and Alternate Calculation Are Not Equal or ERR Appears
Activate IRR Calculation button***

Payback (Years) Greater than Evaluation Period

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9. ANNUAL CASH FLOW

Change for the Evaluation Period

Project Title: DOE PSOFC/MICROTURBINE COMBINED CYCLE (727 kW)

Case No.:

Year No.	Year	Capital	Variable Operating Cost Revenues & Benefits	Fixed Costs	Income Tax		Property Tax and Insurance	Avoided Capital Cost Credit [levelized]	Total	Discount Factor	Discounted Cash Flow
					Savings due to Depreciation	Change due to Change in Income					
0	1997	(\$454,850)	\$0	\$0	\$0	\$0	\$0	\$0	(\$454,850)	1.0000	(\$454,850)
1	1998	\$0	(\$939)	(\$129,951)	\$6,482	\$49,738	(\$9,097)	\$0	(\$83,767)	0.8696	(\$72,841)
2	1999	\$0	(\$941)	(\$135,149)	\$12,478	\$51,714	(\$9,097)	\$0	(\$80,996)	0.7561	(\$61,244)
3	2000	\$0	(\$943)	(\$140,555)	\$11,541	\$53,769	(\$9,097)	\$0	(\$85,285)	0.6575	(\$56,076)
4	2001	\$0	(\$946)	(\$146,177)	\$10,677	\$55,907	(\$9,097)	\$0	(\$89,637)	0.5717	(\$51,250)
5	2002	\$0	(\$948)	(\$152,024)	\$9,875	\$58,129	(\$9,097)	\$0	(\$94,065)	0.4972	(\$46,767)
6	2003	\$0	(\$950)	(\$519,455)	\$9,135	\$197,754	(\$9,097)	\$0	(\$322,614)	0.4323	(\$139,473)
7	2004	\$0	(\$953)	(\$164,429)	\$8,449	\$62,845	(\$9,097)	\$0	(\$103,185)	0.3759	(\$38,791)
8	2005	\$0	(\$955)	(\$171,007)	\$7,816	\$65,345	(\$9,097)	\$0	(\$107,897)	0.3269	(\$35,271)
9	2006	\$0	(\$958)	(\$177,847)	\$7,712	\$67,946	(\$9,097)	\$0	(\$112,244)	0.2843	(\$31,906)
10	2007	\$0	(\$960)	(\$184,961)	\$7,711	\$70,650	(\$9,097)	\$0	(\$116,658)	0.2472	(\$28,835)
11	2008	\$0	(\$963)	(\$647,209)	\$7,712	\$246,306	(\$9,097)	\$0	(\$403,252)	0.2149	(\$86,674)
12	2009	\$0	(\$966)	(\$200,054)	\$7,711	\$76,388	(\$9,097)	\$0	(\$126,019)	0.1869	(\$23,553)
13	2010	\$0	(\$969)	(\$208,056)	\$7,712	\$79,429	(\$9,097)	\$0	(\$130,980)	0.1625	(\$21,287)
14	2011	\$0	(\$972)	(\$216,378)	\$7,711	\$82,593	(\$9,097)	\$0	(\$136,144)	0.1413	(\$19,240)
15	2012	\$0	(\$976)	(\$225,033)	\$7,712	\$85,883	(\$9,097)	\$0	(\$141,510)	0.1229	(\$17,390)
16	2013	\$0	(\$979)	(\$595,384)	\$7,711	\$226,618	(\$9,097)	\$0	(\$371,132)	0.1069	(\$39,659)
17	2014	\$0	(\$983)	(\$243,396)	\$7,712	\$92,864	(\$9,097)	\$0	(\$152,899)	0.0929	(\$14,208)
18	2015	\$0	(\$986)	(\$253,132)	\$7,711	\$96,565	(\$9,097)	\$0	(\$158,940)	0.0808	(\$12,843)
19	2016	\$0	(\$990)	(\$263,257)	\$7,712	\$100,414	(\$9,097)	\$0	(\$165,218)	0.0703	(\$11,609)
20	2017	\$0	(\$994)	(\$273,787)	\$7,711	\$104,417	(\$9,097)	\$0	(\$171,751)	0.0611	(\$10,494)
21	2018	\$0	(\$999)	(\$967,014)	\$3,856	\$367,845	(\$9,097)	\$0	(\$605,408)	0.0531	(\$32,164)
22	2019	\$0	(\$1,003)	(\$296,128)	\$0	\$112,910	(\$9,097)	\$0	(\$193,318)	0.0462	(\$8,931)
23	2020	\$0	(\$1,007)	(\$307,973)	\$0	\$117,413	(\$9,097)	\$0	(\$200,665)	0.0402	(\$8,061)
24	2021	\$0	(\$1,012)	(\$320,292)	\$0	\$122,096	(\$9,097)	\$0	(\$208,306)	0.0349	(\$7,277)
25	2022	\$0	(\$1,017)	(\$333,104)	\$0	\$126,966	(\$9,097)	\$0	(\$216,252)	0.0304	(\$6,569)
26	2023	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a
27	2024	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a
28	2025	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a
29	2026	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a
30	2027	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a
Totals		(\$454,850)	(\$24,311)	(\$7,271,752)	\$172,843	\$2,772,504	(\$227,425)	\$0	(\$5,032,992)		(\$1,337,262)

Prepared by

Date

10. UTILITY REVENUE REQUIREMENTS
For Book Life of Upgrade / Enhancement

Project Title: DOE PSDF CMC ROTORLINE COMBINED CYCLE (727 KW)
Case No.:

Year No.	Year	Undepreciated Balance	Tax Depreciation	Debt Return	Preferred Return	Common Return	Deferred Tax	Income Tax	Other Tax & Insurance	Total Carrying Cost		Fixed Operating Cost <small>(includes (booked) Ratchet Capital Cost/Credits from Sheet 9)</small>		Variable Operating Cost		Revenue Requirement	
											Discounted		Discounted		Discounted		Discounted
0	1997	\$454.85	\$0	\$1	\$1	\$0	\$0	\$0	\$1	\$1	\$0	\$1	\$0	\$0	\$0	\$1	\$0
1	1998	\$437.28	\$17.07	\$32.30	\$5.43	\$42.765	(\$4.33)	\$29.974	\$9.09	\$137.33	\$118.45	\$129.85	\$113.001	\$28	\$917	\$268.228	\$233.241
2	1999	\$413.33	\$32.836	\$31.04	\$5.22	\$41.095	\$5.564	\$22.825	\$9.09	\$133.04	\$100.59	\$135.14	\$102.192	\$941	\$712	\$269.130	\$203.508
3	2000	\$390.59	\$30.370	\$29.35	\$4.94	\$38.861	\$4.627	\$22.219	\$9.09	\$127.29	\$83.69	\$140.59	\$92.417	\$943	\$620	\$268.791	\$176.733
4	2001	\$368.55	\$28.096	\$27.73	\$4.66	\$36.715	\$3.763	\$21.601	\$9.09	\$121.77	\$69.62	\$146.17	\$83.577	\$946	\$541	\$268.894	\$153.739
5	2002	\$347.38	\$26.986	\$26.17	\$4.40	\$34.651	\$2.961	\$20.977	\$9.09	\$116.45	\$67.90	\$152.02	\$75.582	\$948	\$471	\$269.431	\$133.953
6	2003	\$326.98	\$24.039	\$24.67	\$4.15	\$32.662	\$2.221	\$20.342	\$9.09	\$111.34	\$48.13	\$159.45	\$224.572	\$950	\$411	\$631.746	\$273.117
7	2004	\$307.25	\$22.233	\$23.22	\$3.90	\$30.743	\$1.535	\$19.703	\$9.09	\$106.40	\$39.99	\$164.42	\$61.814	\$953	\$358	\$271.789	\$102.172
8	2005	\$289.15	\$20.566	\$21.82	\$3.67	\$28.868	\$902	\$19.054	\$9.09	\$101.62	\$33.22	\$171.80	\$55.901	\$955	\$312	\$273.598	\$89.435
9	2006	\$269.16	\$20.295	\$20.46	\$3.44	\$27.092	\$799	\$17.917	\$9.09	\$97.00	\$27.57	\$177.84	\$50.554	\$959	\$272	\$275.612	\$78.401
10	2007	\$250.17	\$20.291	\$19.11	\$3.21	\$25.307	\$797	\$16.685	\$9.09	\$92.41	\$22.84	\$184.96	\$45.718	\$960	\$237	\$278.334	\$68.788
11	2008	\$231.18	\$20.295	\$17.76	\$2.99	\$23.521	\$799	\$15.450	\$9.09	\$87.81	\$18.87	\$647.20	\$138.110	\$963	\$207	\$735.991	\$158.192
12	2009	\$212.19	\$20.291	\$16.41	\$2.76	\$21.736	\$797	\$14.218	\$9.09	\$83.22	\$15.55	\$200.05	\$37.390	\$966	\$181	\$284.243	\$53.126
13	2010	\$193.19	\$20.295	\$15.07	\$2.53	\$19.950	\$799	\$12.983	\$9.09	\$78.62	\$12.77	\$208.09	\$33.814	\$969	\$158	\$287.653	\$46.750
14	2011	\$174.20	\$20.291	\$13.72	\$2.30	\$18.164	\$797	\$11.751	\$9.09	\$74.03	\$10.46	\$216.37	\$30.579	\$972	\$137	\$291.384	\$41.185
15	2012	\$155.21	\$20.295	\$12.37	\$2.08	\$16.379	\$799	\$10.516	\$9.09	\$69.43	\$8.53	\$225.23	\$27.654	\$975	\$120	\$295.448	\$36.308
16	2013	\$136.22	\$20.291	\$11.02	\$1.85	\$14.593	\$797	\$9.284	\$9.09	\$64.84	\$6.92	\$595.39	\$63.623	\$979	\$105	\$661.207	\$70.657
17	2014	\$117.23	\$20.295	\$9.67	\$1.62	\$12.808	\$799	\$8.049	\$9.09	\$60.24	\$5.59	\$243.39	\$22.617	\$983	\$91	\$304.628	\$28.307
18	2015	\$98.24	\$20.291	\$8.32	\$1.40	\$11.022	\$797	\$6.817	\$9.09	\$55.65	\$4.40	\$253.13	\$20.453	\$986	\$80	\$309.772	\$25.030
19	2016	\$79.24	\$20.295	\$6.97	\$1.17	\$9.237	\$799	\$5.582	\$9.09	\$51.05	\$3.58	\$263.25	\$18.497	\$990	\$70	\$315.307	\$22.154
20	2017	\$60.25	\$20.291	\$5.62	\$94	\$7.451	\$797	\$4.350	\$9.09	\$46.46	\$2.83	\$273.78	\$16.728	\$994	\$61	\$321.246	\$19.627
21	2018	\$45.12	\$10.148	\$4.27	\$72	\$5.665	(\$3.058)	\$6.971	\$9.09	\$41.87	\$2.22	\$967.01	\$51.376	\$999	\$53	\$1,009.882	\$53.853
22	2019	\$33.84	\$0	\$3.20	\$53	\$4.242	(\$6.914)	\$9.844	\$9.09	\$38.20	\$1.78	\$296.18	\$13.661	\$1,003	\$46	\$335.336	\$15.492
23	2020	\$22.56	\$0	\$2.40	\$40	\$3.182	(\$6.914)	\$9.112	\$9.09	\$35.47	\$1.42	\$307.87	\$12.372	\$1,007	\$40	\$344.459	\$13.838
24	2021	\$11.28	\$0	\$1.60	\$27	\$2.121	(\$6.914)	\$8.379	\$9.09	\$32.74	\$1.14	\$320.29	\$11.189	\$1,012	\$35	\$354.054	\$12.368
25	2022	\$1	\$0	\$80	\$13	\$1.061	(\$6.914)	\$7.646	\$9.09	\$30.02	\$912	\$333.10	\$10.118	\$1,017	\$31	\$364.141	\$11.061
26	2023	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a
27	2024	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a
28	2025	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a
29	2026	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a
30	2027	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a
										total	\$1,994.425		\$7,271.752		\$54,311		\$5,290.480
											\$700.130		\$1,414.520		\$6,189		\$2,120.834
										equalized over Book Life (7 Year)	\$108.31		\$218.830		\$854		\$328.097

Prepared by: _____ Date: _____

Pressurized Solid Oxide Fuel Cell/ Gas Turbine Power System

Final Report

Contract Start Date: 29 May 1998
Contract End Date: 30 November 1999

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February 2000

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by

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ABSTRACT

Power systems based on the simplest direct integration of a pressurized solid oxide fuel cell (SOFC) generator and a gas turbine (GT) are capable of converting natural gas fuel energy to electric power with efficiencies of approximately 60% (net AC/LHV), and more complex SOFC and gas turbine arrangements can be devised for achieving even higher efficiencies. The results of a project are discussed that focused on the development of a conceptual design for a pressurized SOFC/GT power system that was intended to generate 20 MWe with at least 70% efficiency. The power system operates baseloaded in a distributed-generation application. To achieve high efficiency, the system integrates an intercooled, recuperated, reheated gas turbine with two SOFC generator stages – one operating at high pressure, and generating power, as well as providing all heat needed by the high-pressure turbine, while the second SOFC generator operates at a lower pressure, generates power, and provides all heat for the low-pressure reheat turbine. The system cycle is described, major system components are sized, the system installed-cost is estimated, and the physical arrangement of system components is discussed. Estimates of system power output, efficiency, and emissions at the design point are also presented, and the system cost of electricity estimate is developed.

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1. EXECUTIVE SUMMARY

1.1 Power System Design Description

Operating at atmospheric pressure, the efficiency horizon for an SOFC power system is 45% (net AC/LHV), while a gas turbine will typically convert to electric power 30% of the fuel energy supplied to the gas turbine (Brayton) cycle. When the SOFC generator and the gas turbine are integrated, as depicted in the hybrid cycle of Figure 1.1, system efficiency near 60% can be achieved. This is due to the enhanced performance of the SOFC generator at elevated pressure, and to the processing to power of SOFC exhaust heat by the gas turbine. Maximum system efficiencies are achieved when no fuel is fired at the GT combustor, which is possible since gas turbines can operate with turbine inlet temperatures in the 850°C to 870°C range, which are typical SOFC exhaust temperatures. With no combustor firing, all fuel enters the system via the SOFC generator, and the fuel energy will have two opportunities for conversion to power — first, by the SOFC electrochemical process, and second, by the Brayton-cycle conversion of SOFC exhaust heat.

Figure 1.1 — Simplified PSOFC/GT hybrid power system cycle.

The High Efficiency Fossil Power Plant (HEFPP) system concept developed by Siemens Westinghouse uses this cycle as its basis, but it incorporates additional features and components that boost the system efficiency to levels nearer the target 70%. The reference HEFPP system concept is depicted in Figure 1.2. Components from the basic hybrid cycle (Figure 1.1) are visible, as well as those added for increased system efficiency. They are the intercooler, the low-pressure (LP) reheat turbine section, and the LP SOFC generator. The intercooler reduces the compressor work input requirement, and reheating increases the turbine shaft power output. Both effects also act to increase the cycle efficiency, provided the cycle is recuperated. The LP generator provides another instance for the serial processing of power system fuel because, in addition to generating power, it provides the heat for the LP turbine, supplanting the reheat combustor. In addition, because of the reheat feature, and relative to the optimum pressure ratio for the basic PSOFC/GT hybrid cycle, the HEFPP system cycle optimizes at a higher compressor pressure ratio. As a result there is a stronger positive effect of elevated pressure on cell voltage at the HP SOFC generator. Figure 1.2 provides detail on the low-maintenance fuel desulfurization system, which is based on the processing of sulfur-bearing compounds to hydrogen sulfide in a cobalt-molybdenum catalyst bed, and the adsorption of the resulting H_2S on heated zinc oxide.

A pictorial view of the power system is presented in Figure 1.3. The dimensions of the rectangular plot plan depicted in the figure are 61 m x 41 m (200 ft x 135 ft), corresponding to a site footprint of approximately 0.6-acre. Visible in the figure are the SOFC generators at the HP and LP locations. Each generator consists of several SOFC modules that are arranged in flow parallel between process air and exhaust manifolds, and each module is a horizontal cylindrical pressure vessel, flanged at the middle, that houses an assembly of 11,520 SOFCs. (An individual fuel cell is tubular, having an active length of 1500 mm and a diameter of 22 mm.) The overall length of a module is approximately 11 m (36 ft), and its diameter is 3.5 m (11.5 ft). The HP SOFC generator consists of four modules, and the LP SOFC generator, five modules — a combination that was selected for maximum system efficiency. It is to be noted that while the power system design was developed for high system efficiency, the reference fuel cost of \$3.00/MMBtu does not permit the maximum efficiency and the minimum cost-of-electricity (COE) to occur at the same design point. This will happen most probably only if the power system is deployed in a region with a higher fuel cost. The reheat gas

turbine, with intercooler and recuperator, is installed between the HP and LP SOFC generators. Rolls Royce Allison specified the turbine and associated equipment.

Figure 1.2 — HEFPP system cycle.

In the SOFC generator design, the flows of air and reformed-fuel on the cathode and anode of each fuel cell occur cocurrently and parallel to the cell axis. This is the conventional configuration that has been used by Siemens Westinghouse in all demonstration SOFC generators designed and operated to date. It was selected for the application in the HEFPP system concept after consideration of the staged-cell SOFC generator concept. The staged-cell design, which retained the conventional air delivery design on the cell cathode side, but employed crossflow on the fuel (anode) side, was originally believed to enable the SOFC generator to operate at very high fuel utilizations, thereby contributing to the achievement of higher generator and system efficiencies.

- Key:
1. GT Skid	11. Propane Tank
2. Filter House	12. Auxiliary Air Compressor
3. HP Air Heater	13. Nat. Gas Compressor
4. LP Air Heater	14. Desulfurizer
5. HP SOFC Vessel	15. Elec. Cabinets
6. LP SOFC Vessel	16. UPS Shed
7. Storage Shed	17. H ₂ Generator/Compressor and Gas Mixer
8. 20 MW Substation	18. Water Storage Tank
9. N ₂ Tube Trailer	19. Startup Boiler
10. N ₂ Generator/Compressor	20. Control/Meeting Room

Figure 1.3 — Power system site arrangement.

Based upon analysis, the potential benefits of staging proved elusive for two major reasons. First, since the first cell stage is fed relatively cold fuel, the first few cell stages operate at substantially less than optimum temperature for yttria stabilized zirconia electrolyte cells.

Second, the last cell stage can not be operated at fuel utilization greater than that for the co-current flow stack because the approximately parabolic axial temperature distribution and concomitant non-uniform current density distribution place the cell hot spot

at hazard for anode oxidation at roughly the same average fuel utilization achievable in a non-staged stack.

1.2 Performance Estimates

The power system is designed for baseload operation at the peak-efficiency system design point. At that point, the estimated system net AC power output is 19.0 MWe, and its efficiency is 67.3% (net AC/LHV). Approximately 15 MWe are derived from the SOFC generator modules, and the remainder from the gas turbine. The estimated rates of CO₂ and NO_x emission at the system design point are 300 kg/MWh (5.7 vol %) and 0.006 kg/MWh (1 ppm_v), respectively; the exhaust flow rate and temperature are 19 kg/s (41 lb/s) and 225°C (437°F).

The PSOFC/GT power system conceptual design reported herein misses the target efficiency of 70% by 2.7 points. The desulfurization scheme used is a contributing factor to this efficiency shortfall because it employs an electrolyzer to provide the small stream of H₂ gas needed. If an ambient-temperature sorbent (as assumed in past studies) were practical, the resultant system efficiency would be 68% at a system net power output of 19.2 MWe. In addition, the SOFC power conditioning system (PCS) efficiency was set at 94% for the conceptual design. This value is two percentage points less than the more optimistic value applied in past HEFPP studies. Improvements in PCS performance, which may be possible, and should be evaluated, would translate directly to a higher HEFPP system efficiency. For example, at the current HEFPP system design point, a boost in PCS efficiency to 97% would result in an increase in system power output to 19.5 MWe and in efficiency to 69%. Finally, the two to five point efficiency gain believed possible for the proposed electrochemical staging of the tubular SOFC was determined to be unobtainable in practice for fuel cells operating at economically meaningful current density. When the projected effects of the passive ambient-temperature desulfurizer sorbent and the improved PCS performance are combined, the estimated system power output and efficiency are 19.7 MWe and 69.6%, and the addition of a small low-pressure steam turbine cycle would result in system net AC output of 20 MWe, at an efficiency of 71%. Thus, it appears that the target efficiency of 70% may be achievable through improvements in PCS efficiency and the desulfurization technique, and through the addition of the steam turbine generator, without electrochemical SOFC staging.

1.3 System Cost and Economics

The system installed-cost estimate is \$1431/kWe. It includes costs related to site preparation, equipment procurement, shipping, and installation, as well as allowances for G&A, sales and marketing, and profit. Mature technologies and products were assumed. The distribution of the power system installed cost is provided in Table 1.1.

Table 1.1 — HEFPP System Installed Cost Summary

	Cost, \$/kWe
SOFC Generator Equipment	471
SOFC Power Conditioning Equipment	107
Gas Turbine Equipment	211
Balance of Plant Equipment	259
Subtotal	1048
Site Preparation	22
Project Management and Engineering	48
Overhead and Profit	291
Total Plant Cost	1409
Spare Parts, Startup, and Land Allowance	22
Total Capital Requirement	1431

COE estimates have been developed for the HEFPP system, operating at its design point, and also for a competing technology, which was assumed to be a 20 MWe-class gas turbine combined cycle power system. Table 1.2 summarizes input used in the analysis, and also the results.

The projected COE is approximately 3% higher than conventional-technology COE. The conventional-technology power system will emit more NO_x and SO_x (see emission estimate summary in Table 1.2), and the cost of equipment to reduce those emissions to HEFPP levels would improve the relative COE performance of HEFPP system. Comparing the systems in a higher fuel cost environment would also improve the COE attractiveness of the HEFPP system. For example, with \$6 fuel, the HEFPP system COE estimate is 7% less than the conventional-system COE (66 mills/kWh, vs. 71 mills/kWh), and with \$9 fuel it is 13% less (83 mills/kWh, vs. 95 mills/kWh).

Table 1.2 — Economic Analysis Summary of Results

	HEFPP system (PSOFC/GT)	Conventional-Technology Power System (Gas Turbine/Steam Turbine Combined Cycle)
No. of round-the-clock operators	1	1
No. systems in operation	5	5
Labor cost components	System operation & house-keeping maintenance	System operation & house-keeping maintenance
Gas turbine/steam turbine system maintenance	\$0.01/GT kWe	\$0.007/system kWe
Power system capacity factor	0.92	0.92
Capital charge rate	15%	15%
Fuel cost	\$3/MMBtu	\$3/MMBtu
Power system capital cost	\$27.3M	\$14.9M
Power output	19.0 MWe	17.9 MWe
Emissions estimates		
CO ₂	300 kg/MWh	420 kg/MWh
NO _x	0.006 kg/MWh (1 ppm _v)	0.380 kg/MWh (25 ppm _v)
SO _x	Virtually zero	2.5 g/MWh (4 ppm _v S in fuel)
Power system efficiency (net AC/LHV), %	67.3	47.9
Cost of electricity, mills/kWh	49.1	47.6

* Source: Gas Turbine World 1997 Handbook, turnkey power generation projects, p. 24.

1.4 Conclusions and Recommendations

Study conclusions can be summarized as follows:

- A PSOFC/GT system concept of near 20 MW capacity has been devised that is conservatively capable of 67% efficiency, a value ten points greater than that achievable with the best available large-plant conventional power generation technology, and twenty points above the efficiency achieved by a conventional 20 MW-class gas turbine combined cycle power system.
- The specific power system concept developed during this study, integrating HP and LP SOFC generators with an intercooled, SOFC-reheated gas turbine, achieves an estimated power output of 19 MWe at an efficiency of 67.3% (net AC/LHV). Improvements in the performance of major system components, particularly in the SOFC PCS, for which there was no study design task, and employment of an ambient-temperature passive sorbent technology for fuel desulfurization would cause the system efficiency estimate to approach very closely the 70% efficiency target. The addition of the steam turbine cycle, which could have a marginal practicality due to its small output and the increased power system complexity and maintenance, could boost the power system efficiency to slightly above 70%.

- The staged-cell SOFC stack design does not offer the large SOFC efficiency gain (over the standard cocurrent axial flow stack design) that was projected originally. Cell cooling in the fuel-entry cell rows reduces the average cell voltage while there is little increase in average fuel utilization at the last cell row at meaningful current densities because of the hazard of anode oxidation.
- For the reference fuel cost of \$3.00/MMBtu, the estimated COE for the HEFPP system is 3% higher than the COE estimate for a conventional 20 MW-class gas turbine/steam turbine power system. Leveraged by its significantly higher efficiency, the HEFPP system would have a COE advantage in a higher fuel cost environment. For example, with \$6 fuel, the HEFPP COE would be 7% less than the conventional-system COE.

Recommendations:

- Desulfurization technologies not requiring a source of hydrogen, and capable of operation at ambient-temperature levels, should be developed.
- Power conditioning topologies with greater than 95% efficiency should be developed.
- For deployment in SOFC/GT hybrid cycle power systems, small, efficient, highly-reliable, recuperated gas turbines with turbine inlet temperature commensurate with SOFC exhaust gas exit temperatures (870°C) should be developed.
- A PSOFC/GT power system of 70% efficiency potential should be developed and demonstrated at the smallest capacity class practical for proof-of-concept.
- SOFC development should be pursued to improve intrinsic power density and to ensure operational feasibility at elevated pressures beyond 3 atmospheres.

2. INTRODUCTION

A simplified cycle diagram for a simple-cycle, atmospheric pressure SOFC power system is provided in Figure 2.1. Fuel cell process air is supplied by the air blower, and the air is preheated as needed for SOFC thermal management using heat recovered at the recuperator from the SOFC exhaust gas. The SOFC generator operating pressure is near atmospheric, typically being in the 30 to 50 mbarg (10 to 20 in. H₂O) range. SOFC power systems based upon this cycle are capable of electric generating efficiencies in the 45% to 50% (net AC/LHV) range. The 100 kWe SOFC combined heat and power (CHP) demonstration power system operating in the Netherlands is based on this cycle. That system, designed and fabricated by Siemens Westinghouse and EDB/ELSAM, a team of Dutch and Danish generating and distribution companies, and sponsored by EDB/ELSAM, is installed at a utility site near Arnhem. To date, the unit has logged over 10,000 operating hours, it is generating approximately 110 kWe net AC power at 46% efficiency (net AC/LHV) for the utility grid, and it also produces hot water for the local district heating system. The demonstrated system energy efficiency is nearly 75%.

Figure 2.1 — Atmospheric pressure SOFC power system cycle.

Power system efficiency can be increased by integrating the SOFC generator with a gas turbine in the cycle depicted in Figure 2.2. The SOFC generator in this cycle is pressurized (PSOFC), operating on air coming directly from the compressor discharge, via the turbine exhaust recuperator. The system based upon this cycle achieves high efficiencies due to the utilization by the gas turbine generator of thermal energy in the pressurized SOFC exhaust stream. Power system performance is also enhanced by

SOFC generator operation at elevated pressure. For a given cell operating current, cell voltage, and hence cell power output and efficiency, increase with pressure. Assuming PSOFC operation at 50% efficiency and the conversion of SOFC exhaust heat to power by the gas turbine at 30% efficiency, it is clear that the cycle efficiency can be made to approach 60%. This is predicated on the restriction that the gas turbine combustor is not fired during normal, high-efficiency power operation. It could, however, be fired for peak-power purposes. While combustor firing would indeed result in more power from the system, it would not increase the system efficiency because the incremental fuel is converted to power at the relatively low gas turbine efficiency level. The SOFC generator is ideally suited for deployment in a power system of this type since the SOFC exhaust gas has a temperature of about 870°C. Gas turbines can operate with turbine inlet temperatures of 870°C, and therefore the two technologies are directly integrable, with the SOFC generator effectively supplanting the turbine combustor.

Figure 2.2 — Simplified PSOFC/GT hybrid power system cycle.

The purpose of the present project, undertaken by Siemens Westinghouse Power Corporation, in conjunction with Rolls Royce Allison, was to develop a conceptual design and feasibility study for a power system that utilizes the PSOFC/GT cycle to achieve a system electric generation efficiency of 70%. An adjunct objective is that the system yield a reduced cost-of-electricity (COE) relative to the COE for a conventional power system, which, for this study, was assumed to be a gas turbine combined cycle. The power system capacity was to be 20 MWe. The reference power system in the Siemens Westinghouse/Rolls Royce Allison study is based upon a recuperated reheat

gas turbine cycle. In the reheat turbine cycle, the gas turbine has two expander stages, each stage is equipped with a combustor, and after the hot gas produced at the HP combustor expands partially across the HP turbine, it is reheated at the LP combustor before the expansion process is completed across the LP turbine stage. The recuperated reheat turbine cycle will have a higher efficiency than the recuperated single-stage cycle because reheating increases the cycle's average heat reception temperature, without affecting the temperature at which heat is rejected. Then, applying that cycle, and supplanting both combustors with SOFC generators, efficiencies well above the 60% level forecast for the simpler PSOFC/GT cycle of Figure 2.2 should be achieved. The efficiency improvement is due to the implementation of the more efficient gas turbine cycle (intercooled, reheated) and to the cycle's optimization at higher compressor pressure ratios. As the result of the higher pressure ratio, there is a stronger positive effect of elevated pressure on cell voltage at the HP SOFC generator.

The conceptual design of a high-efficiency power system based upon the reheat turbine cycle is developed and discussed in this report. Tradeoff studies are discussed, the features of the main components in the reference power system design are described, and power system performance and COE estimates are presented.

3. RESULTS AND DISCUSSION

3.1 Power Plant Conceptual Design

3.1.1 Design Requirements and Objectives

3.1.1.1 Introduction

This document establishes requirements and specifications for use by the Siemens Westinghouse Power Corporation in developing the conceptual design for a high-efficiency fossil-fueled power plant. The plant concept to be developed by Siemens Westinghouse is based on the integration of solid oxide fuel cell (SOFC) and gas turbine technologies.

3.1.1.2 Design Basis

Power Plant Application	Commercial distributed-power generation
Output Power Specification	60 Hz AC, at utility-grid voltage
Utility AC Grid Connection	The power plant will be connected to the utility grid, and all net plant power will be exported to the grid.
Dispatch Mode	Base load
Power Plant Heat Recovery	Heat will be recovered for plant power generation support; no heat will be recovered for site thermal application.
Power Plant Startup	The utility AC grid will be available for plant startup operations.
Installation	Outdoors.
Conceptual Design Scope	Power plant equipment between the site fuel supply point and the AC grid interface will be included in the design. The equipment considered will be essential for plant startup, operation, control, shutdown, and maintenance.

3.1.1.3 Performance Requirements

Power Plant Design-Point Capacity	20 MW net AC, +/- 2 MW
Design-Point Efficiency	70% (net AC/LHV)
Power Plant Normal-Operation Turndown Requirement	None
Power Plant Overpower Requirement	None
Output Power Conditions	60 Hz Utility-grid quality
Acoustic Noise Control	Consistent with typical gas turbine practice.

3.1.1.4 Fuel and Oxidant Specifications at the Power Plant Design Point

Power Plant Fuel	Natural gas
Fuel composition	<u>(mol percentages)</u>
Methane	96
Nitrogen	2
Carbon dioxide	2
Sulfur bearing compounds	Mercaptans
Sulfur concentration	4 ppmv
Fuel supply pressure	1.034 barg (15 psig)
Fuel supply temperature	15°C (59°F)
Oxidant	Air (ISO conditions)
Air composition	<u>(mol percentages)</u>
Oxygen	20.7
Nitrogen	78.0
Carbon dioxide	0.3
Water	1.0 (60% relative humidity)
Ambient air temperature	15°C (59°F)
Ambient air pressure	1.014 barg (14.7 psia)
3.1.1.5 Ambient Air Ranges	
Pressure	Sea level to 1500 m (5000 ft)
Temperature	-29°C to 49°C (-20°F to 120°F)
Relative Humidity	0 to 100%

3.1.1.6 Power Plant Physical Design Objectives

Major Subsystem Fabrication	Maximum practicable skid mounting at the factory.
Transportation Options	Truck, sea, air, rail
Site Installation Operations	Minimum component/skid assembly at installation site. Interconnect factory-assembled skids at the site. Interface the power plant with the site.
Plant Design Lifetime	Conventional power plant equipment - 25 years.

3.1.1.7 Power Plant Operation

Normal Power Operation Mode	Automatic, unattended, remotely monitored
Power Plant Startup	Attended
Annual Operating Time	Fifty weeks
Annual Planned Shutdown	Two weeks

3.1.1.8 System/Site Interface Requirements

Appropriate interface points will be available at the installation site for:

- Fuel supply
- Obtaining utility AC power during plant startup operations.
- Connecting the plant power output with the utility grid.

3.1.1.9 Economic Evaluation Parameters

Power System Cost Estimation Basis	Costs will be based on the projected needs of mature SOFC/gas turbine technologies and commercial power plant operation, not first-of-a-kind.
Cost Basis	Mid-1998 US dollars

3.1.1.10 Economic Evaluation Parameters (continued)

Fuel Cost	\$3.00/MMBtu (HHV)
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Capital Charge Rate	15%
Availability	92%
COE Evaluation Method	Constant dollars
Power Plant Optimization Basis	Consistent with >70% (LHV) efficiency, achieve a design-point COE that is 10-20% below the COE of today's conventional plants.
Transportation Cost Basis	800 km (500 miles) - factory to installation site.
Conventional Power Plant COE Basis	Gas turbine combined cycle

3.1.2 Cycle and Power System Description

The direct integration of a pressurized SOFC generator and a gas turbine in the basic PSOFC/GT hybrid cycle, Figure 2.2, enables the generation of electric power at high efficiencies - typically in the 55% to 60% (net AC/LHV) range. As explained above, this is due to the extended processing by the gas turbine of system fuel energy that is not converted electrochemically to power by the fuel cell, and to the operation of the SOFC generator at elevated pressure. With a peak cycle temperature of 870°C (the SOFC generator exhaust temperature), the optimum compressor pressure ratio is 2.5:1 to 3.0:1, and the SOFC generator therefore operates at a pressure in the 2.5 bar(abs) to 3.0 bar(abs) range. The optimum pressure ratio is determined from peak-efficiency considerations, and is found by trading the positive effect of increasing pressure on SOFC efficiency against the negative effect of operating the SOFC generator at higher, less efficient, cell currents. As the design pressure ratio is increased, while the turbine inlet temperature is fixed at 870°C, the turbine exit temperature drops, cooling the recuperator and requiring the SOFC generator to operate at the high currents.

The advanced power system cycle upon which the 20 MWe PSOFC/GT power plant design is based is depicted in Figure 3.1. It builds on the basic PSOFC/GT hybrid cycle, but it provides for further increases in the system efficiency by three mechanisms – higher SOFC operating pressure [>3 bar(abs)], compressor intercooling, and turbine reheat. Since the peak cycle temperature, occurring at the two turbine inlets, is 870°C, the optimum expansion across each turbine will again be in the 2.5:1 to 3.0:1 range. Thus, the LP SOFC generator will operate, as does the generator in the basic

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to-air heat exchanger. Reheating, again in combination with recuperation, increases the average Brayton cycle heat reception temperature, without changing the average heat rejection temperature, and this also translates directly to a higher Brayton cycle efficiency. The system components are flow matched, the fuel cell operating points are chosen such that the SOFC generator exhaust temperatures are approximately 870°C, and there is no firing of fuel at the gas turbine combustor and air heaters. The combustor and heaters will typically function only during system startup operations, although it is conceivable that the combustor could be fired to achieve peak power output.

The system fuel is pipeline natural gas, assumed in this study to consist of 96 vol % methane, 2% nitrogen, and 2% carbon dioxide. The gas also contains sulfur-bearing compounds, occurring naturally, or added to enable leak detection. The sulfur concentration in the raw fuel, per the design requirements, is 4 ppmV, and it must be reduced to the 0.1 ppmV level prior to SOFC generator entry to preclude the adsorption (reversible) of sulfur on SOFC nickel components. After the fuel has been hydrogenated, the desulfurizer in this power system concept processes the sulfur to hydrogen sulfide in a cobalt-molybdenum catalyst bed, and the H₂S is adsorbed on a bed of hot zinc oxide that operates optimally at 350°C to 400°C (660°F to 750°F). This temperature level is achieved recuperatively and by electric heat addition. The power required for fuel heating is small, and the system is simple and requires low maintenance. Alternatively, the heat for this process could be derived from the turbine exhaust, at the expense of complicating the turbine exhaust design. Hydrogen for the fuel desulfurization process is generated on site.

The SOFC generators produce DC power, which is prepared for export to the utility AC grid by the power conditioning systems. AC power is also produced for export by the gas turbine.

The gas turbine compressor is composed of two radial stages separated by the intercooler. The compressor air intake rate is approximately 18 kg/s (40 lb/s), and the design compressor pressure ratio is 7:1. The stage pressure ratio, allowing for intercooler pressure drop, is 2.73:1. The HP and LP turbine sections each consist of a single axial wheel.

As Figure 3.1 indicates, the rotating gas turbine components are installed on a single shaft, and it is noted that the electric load on the gas turbine generator will be modu-

lated to maintain set-point shaft speed. This feature will assure a steady flow of air to the HP SOFC generator, an important function of the gas turbine, and a necessity to provide for SOFC thermal management.

If the LP turbine were a free power turbine, the expansion ratio across the HP turbine would be set by work-balancing the HP turbine with the compressor. Given a compressor pressure ratio, this would determine the gas temperatures at the HP and LP SOFC inlets, the cell current levels, and the cell operating efficiencies. However, with the single-shaft arrangement, the HP turbine expansion ratio can be an independent variable, and since it does influence the performance of both the HP and LP SOFC generators, an optimum value for the ratio can be determined for maximum system efficiency. It is found that the optimum HP turbine expansion ratio is a function of the compressor ratio, ranging between 0.55 for a pressure ratio of 5:1, and 0.35 at 12:1; for the design pressure ratio of 7:1, the optimum expansion ratio is 0.5. Additional discussion of the optimum expansion ratio, and of the optimum compressor pressure ratio for the high-efficiency power system, is presented in Section 3.4.1. A discussion of compressor intercooling is provided in Section 3.4.2. As noted there, intercooling does increase system efficiency and power output, but due to increased capital cost and the added maintenance requirements, a reduced COE does not necessarily follow. The intercooling feature was selected for the HEFPP cycle due to the project's high-efficiency focus.

A pictorial view of the power system is presented in Figure 3.2. Visible are the SOFC generators at the HP and LP locations. Each generator consists of several SOFC modules that are arranged in flow parallel between process air and exhaust manifolds. Each module is a horizontal cylindrical pressure vessel, flanged at the middle, that houses a cell stack assembly. The basic stack building block is the 576-cell substack pictured in Figure 3.3. The individual cells are tubular, with active lengths of 1500 mm and diameters of 22 mm. They are arranged in the substack in the vertical orientation, with closed ends at the bottom. Shown at the top of the substack are air distribution plenums through which process air is admitted to the cell air injection tube inlets, and on the side of the substack, shown are the depleted-fuel recirculation plenum, the ejector that drives the depleted-fuel recirculation, the fuel prereformer, and the ducting for distributing the fresh-fuel/recirculated-fuel mixture to the underside of the substack. Within the substack, methane reformation occurs in the in-stack reformers located between

the cell bundles, and the reformed fuel mixture is then distributed to the individual cells. At the cells, the fuel mixture and air flow cocurrently from the cell closed ends, and a fraction of the CO and H₂, typically 85% to 90%, is processed electrochemically. For the high-efficiency power system, twenty 576-cell substacks compose the cell stack assembly in a single generator module. The HP SOFC generator consists of four modules, manifolded as indicated in Figure 3.2, and the LP SOFC generator is composed of five modules. It is to be noted that this power system design has been developed for maximum system efficiency, realizing that the main objective of the project was the development of a power system concept that could reach the 70% efficiency (net AC/LHV) level.

- Key:
1. GT Skid	11. Propane Tank
2. Filter House	12. Auxiliary Air Compressor
3. HP Air Heater	13. Nat. Gas Compressor
4. LP Air Heater	14. Desulfurizer
5. HP SOFC Vessel	15. Elec. Cabinets
6. LP SOFC Vessel	16. UPS Shed
7. Storage Shed	17. H ₂ Generator/Compressor and Gas Mixer
8. 20 MW Substation	18. Water Storage Tank
9. N ₂ Tube Trailer	19. Startup Boiler
10. N ₂ Generator/Compressor	20. Control/Meeting Room

Figure 3.2 — Power system site arrangement.

Figure 3.3 — Basic stack building block is the 576-cell substack.

3.1.3 Power System Performance Analysis

3.1.3.1 Analysis Basis

Basic input for the analysis such as fuel composition, fuel supply conditions, and ambient-air conditions were taken from the design requirements, Section 3.1. Additional information on key input is provided in the following:

- Cell V-I characteristic – the V-I characteristic is graphed in Figure 3.4. It is a projected characteristic for the mature SOFC product that will be available in 2005 to 2010. The V-I characteristic applies to operation at 1 atm (abs), 85% fuel utilization, and to a peak cell temperature of 1020°C (1870°F).

Figure 3.4 — Cell V-I characteristic.

- Cell voltage correction for SOFC generator operation at elevated pressure – A cell voltage adjustment for operation at pressures above 1 atm (abs) is presented in Figure 3.5. Given an operating pressure, the corresponding voltage adjustment is added to the base cell voltage from Figure 3.4. Data for Figure 3.5 were obtained from cell testing (Test No. 503) performed at Ontario Hydro Technologies (OHT) in Toronto, Ontario, Canada by OHT personnel. The test article was designed and supplied by Siemens Westinghouse. The tests covered the pressure range from 1 atm (abs) to 15 atm (abs). Over the cell current density range of interest in this conceptual design study, the adjustment for pressure is essentially independent of current density.

Figure 3.5 — Cell voltage adjustment for pressure.

- Stoichs profile – air flow to the HP SOFC generator is determined by the stoichs profile that is graphed in Figure 3.6. Its application results in cell operation with a peak cell temperature of 1020°C (1870°F) and a generator combustion zone exhaust temperature of approximately 870°C (1600°F). One stoich provides the normal-air flow needed to supply oxygen for the cell electrochemical process. A generator air flow based upon multiple stoichs provides the required amount of oxygen for that process, and it flattens the cell axial temperature distribution, thereby raising the cell average temperature. From energy balance considerations, and given a set combustion zone exhaust temperature, the generator air inlet temperature increases with decreasing cell current.
- Electrochemical fuel utilization – generator fuel consumption was set at 90%, meaning 90% of the fuel admitted to the SOFC generator was consumed electrochemically on the cell active surface and by the ionic and molecular leakage of oxygen from the cathode side of the cell to the anode. Approximately 89% of the generator fuel is consumed electrochemically.

Figure 3.6 — Stoichs profile.

- SOFC power conditioning system efficiency – the development of a power-conditioning concept was not a task in this project. For the performance estimates, the power conditioning efficiency was assumed to be 94%. This value covers system losses from the SOFC DC terminals to the utility AC grid.
- Gas turbine generator efficiency – 96%.
- Gas turbine compressor isentropic efficiency – 86.4%.
- HP turbine isentropic efficiency – 90.7%.
- LP turbine isentropic efficiency – 91.3%
- Auxiliary power losses – 125 kW_e [I&C, cabinet ventilation, and intercooler heat rejection (water circulation and forced-air fan power)].

3.1.3.2 Power System Performance Estimates

Power system design-point performance has been analyzed; results are summarized in Table 3.1.

Table 3.1 — Power System Design-Point Performance Estimates

Compressor air intake rate	18.1 kg/s
Compressor pressure ratio	7:1
HP SOFC generator DC power	9.0 MWe
LP SOFC generator DC power	7.5 MWe
SOFC gross AC power	15.6 MWe
Compressor shaft power	4.1 MW
HP turbine shaft power	3.4 MW
LP turbine shaft power	4.9 MW
Gas turbine gross AC power	4.1 MWe
Power system net AC power	19.0 MWe
Fuel flow rate to power system	0.62 kg/s
Efficiency (net AC/LHV)	67.3%
Carbon dioxide emission	300 kg/MWh
Nitrogen oxide emission (based on 1.0 ppm _v)	0.006 kg/MWh
Exhaust flow rate	18.7 kg/s

Corresponding to the operating point represented in Table 3.1, and for the points identified in Figure 3.7, estimates of cycle statepoint parameters are provided in Table 3.2.

Table 3.2 — Power System Statepoint Parameter Estimates

Statepoint ID	Mass Flow, kg/h	Temperature, °C	Pressure, bara	Molecular Weight	Enthalpy Flux, MW
1	2,224	15	2.05	16.8	-0.014
2	2,224	187	10.14	16.8	0.249
3	3	15	10.14	2.0	
4	2,226	187		16.7	0.249
5	2,226	378		16.7	0.609
6	2,226	399		16.7	0.653
7	2,226	399		16.7	0.653
8	2,226	208		16.7	-0.058
9	0				0.000
10	1,248	93		16.7	0.056
11	0				0.000
12	978	93		16.7	0.044
13	0				0.000
14	65,285	15	1.01	28.8	-0.182
15	65,285	15	1.01	28.8	-0.182
16	65,285	125	2.75	28.8	1.834
17	64,959	24	2.59	28.8	-0.021
18	64,959	137	7.06	28.8	2.048
19	63,980	608		28.8	11.056
20	63,980	603		28.8	10.956
21	63,980	603		28.8	10.956
22	65,228	876		28.4	17.578
23	65,228	871	6.34	28.4	17.466
24	65,718	703	3.17	28.4	13.862
25	65,718	703		28.4	13.862
26	66,695	875		28.1	18.465
27	66,695	870	2.82	28.1	18.347
28	67,185	647	1.07	28.1	13.281
29	67,185	225	1.01	28.1	4.029

3.1.3.3 Discussion

Power System Efficiency. The predicted power system efficiency (67.3%) is 2.7 points less than the 70% project target. However, it appears the target could be reached through improvements in the performance of the fuel desulfurization and SOFC power conditioning systems, combined with the generation of additional power by a small steam turbine generator; steam for the turbine would be produced via the recovery of heat at the recuperator exhaust. In the reference HEFPP power system design, the fuel desulfurization system uses hot catalysis and sorbent beds in which temperature is

maintained with the aid of a parasitic electric heat input, and for which a small hydrogen stream produced by the electrolysis of water (also needing power input) is required. Eliminating both parasitics through the application of an effective ambient-temperature fuel desulfurization process, and one needing no hydrogen input, would increase the system net AC power output by 200 kWe, and the system net AC efficiency by one-half of one percentage point.

Concerning the SOFC power conditioning system (PCS), and as noted in Section 3.1.3.1, the power system performance estimates assumed a PCS DC-to-AC-grid efficiency of 94%. A one-point increase in that efficiency, achieved through engineering development of the PCS technology, would increase the power system output by approximately 160 kWe, producing an increase in system efficiency of approximately six-tenths of one percentage point.

Relative to steam turbine power, the heat available for steam generation at the recuperator exhaust is of low grade, and the steam pressure that could be developed for effective use would be low, probably 5.5 barg to 7.0 barg. Nonetheless, if it were done, and assuming a 25% steam turbine cycle efficiency, it would be possible to generate 360 kWe of additional AC power output.

The individual and combined effects of these component improvements and cycle modifications (i.e., application of low-temperature, zero-H₂ desulfurization technology; use of a higher-efficiency PCS; and the generation of steam turbine power) are graphed in Figure 3.8. Combining them, it appears that the 70% efficiency goal is achievable. A complete evaluation of the effects must be done that considers cost and maintenance effects as well as the performance benefits. This has not been done in the present study for these particular system refinements, and it is an area for further work. It is noted in particular that the steam cycle would add complexity and maintenance needs to the power system, and these will have to be weighed carefully against the benefit gained. A distinct advantage of the reference system concept is that it is dry; as a result, it involves none of the maintenance demands related to water treatment, corrosion, etc, and it imposes no on-site operator requirements that are frequently associated with pressurized, steam-related systems.

Figure 3.8 — Power system performance sensitivities.

Elevated SOFC Generator Operating Pressure. As indicated earlier, the HEFPP SOFC generator design utilizes the conventional generator configuration, and in addition, the design assumes the uniform distribution of air flow between cells, and the uniform distribution of reformed-fuel flow on the cell anode side. In the conventional generator configuration, fuel flows upwards along the cell exterior surfaces (anodes) in the system of communicating parallel flow channels that is defined by the anodes. That this configuration is suitable for atmospheric-pressure generators has been confirmed via current and past field unit demonstrations, and a pressurized bundle at three atmospheres pressure has also exhibited good, stable performance. However, it is not known with certainty that this parallel flow channel configuration is suitable for higher pressure SOFC generators. Flow channel pressure drop is determined by friction and buoyancy components. At low pressures, the friction component, varying inversely with gas density, and hence with pressure, is dominant, and the buoyancy effects of modest channel-to-channel temperature or composition variations are of little conse-

quence. However, at elevated operating pressures, the buoyancy effect, being proportional to density differences, has more influence, while the friction pressure drop component, with its inverse pressure relationship, is diminished. At higher pressure ($P > 3$ atm) transverse channel-to-channel parameter variations could affect parallel-channel flow stability, and this is of special concern considering the density of the oxidation product stream that exits the channels at the top will be larger than the density of the reformed fuel stream that enters at the bottom. It is therefore clear that cell-stack thermal and hydraulic interactions need to be well understood, particularly relative to stack operation at elevated pressure levels. A very reasonable first step in developing this understanding would be to undertake a detailed analytical evaluation of the anode-side flow field and of the effect of operating pressure on that field.

3.1.4 Power Plant and Site Arrangement

Power system pictorial and plan views are presented in Figure 3.9 and Figure 3.10.

The arrangement includes a trailer mounted control/meeting room with sanitary facilities and four-space parking area, a storage shed, a 20 MW substation/switchyard, two fire hydrants, intake air filter house and the BOP equipment block. The BOP equipment block is shown between the control trailer and switchyard. The BOP equipment block is comprised of several equipment, component and electrical skids. Each skid is mounted to a dedicated concrete pad or foundation. The various skids are in weather-proof enclosures:

- Electrical cabinets
- Water storage tank/pump skid
- Packaged startup boiler
- Auxiliary air compressor with dual fuel engine
- Propane tank
- Natural gas compressor and accumulator

- Key:
1. GT Skid	11. Propane Tank
2. Filter House	12. Auxiliary Air Compressor
3. HP Air Heater	13. Nat. Gas Compressor
4. LP Air Heater	14. Desulfurizer
5. HP SOFC Vessel	15. Elec. Cabinets
6. LP SOFC Vessel	16. UPS Shed
7. Storage Shed	17. H ₂ Generator/Compressor and Gas Mixer
8. 20 MW Substation	18. Water Storage Tank
9. N ₂ Tube Trailer	19. Startup Boiler
10. N ₂ Generator/Compressor	20. Control/Meeting Room

Figure 3.9 — Power System Arrangement — Isometric View.

Figure 3.10 — Power System Arrangement — plan view.

- Natural gas desulfurizer
- Pressurized nitrogen storage cylinders
- Nitrogen generator and compressor
- Hydrogen generator, compressor and gas mixing equipment

Not shown are the security fence and its gates around the perimeter or any storm water drainage features. These features will generally be site specific. The overhead high voltage power lines and their poles from the SOFC vessels and GT skid are also not shown.

For this study it is presumed that direct access to the site is via paved roadway capable of bearing the loads from the heaviest items [GT skid weight is 29,000 Kg (64,000 lb), and the SOFC vessel half weight with internals is 25,400 Kg (56,000 lb)]. The site must also have direct connections to natural gas, city water, sanitary sewer, telephone and electric power. The sites, as shown, do not allow for sanitary sewer drain fields or wells for water.

To minimize field construction, each SOFC vessel half with SOFC stacks installed is factory fabricated on a transportable trailer/skid. This foundation trailer is field mated with the trailer bearing the other half of the SOFC vessel. After proper alignment of the vessel/trailer halves, the trailers are blocked and the wheels removed. In the event that a vessel half is to be returned to the factory for maintenance or repair, the wheels can be reinstalled on the foundation trailer, the blocks removed and the trailer towed away.

Access to the switchyard is available on three sides and the BOP equipment block also has fork truck clearance between skids. The process piping is factory-fabricated, internally-insulated, flanged sections. Where possible, externally-jacketed thermal insulation is applied at the factory to minimize field construction. These sections are relatively large and heavy, varying from about 1000 kg per meter (750 pounds per linear foot) to more than 2500 kg per meter (1700 pounds per foot). They are assembled on site by landing them on their pipe supports and bolting together the gasketed, flanged joints. Thermal expansion joints are provided between significant runs of piping as needed. The plant arrangement drawings do not show the detail of the flanged joints in the process piping or the pipe supports.

The arrangement uses SOFC pressure vessels with process air entry and exhaust exit at opposite ends. This vessel and internal SOFC stack is the easiest design. The pneumatic balancing for equal distribution of air to the cells is straightforward. The significant consequence of the vessel with nozzles on both ends is the need to use a relatively large quantity of large diameter, internally and externally insulated piping. This plant occupies the largest amount of ground of the three arrangements studied (see Section 3.4.3). The plot shown is 61 m (200 ft) by 41 m (134 ft), or 0.6 acres.

The SOFC generator modules are accessible from each end, and there is space for a vehicle access road at each end. Access for inspection, repair or maintenance is good. Routing of high voltage electric power lines from the SOFC vessels to the switchyard is efficient. The DC power generated in each vessel is routed through thermally and electrically insulated feedthroughs directly into a top-mounted package containing a DC to AC inverter with step-up transformer. The inverter/transformer package is mounted onto each pressure vessel at the site. Shipping restrictions prevent installation of this electrical equipment at the factory. The top of the inverter/transformer package terminates with three standard high voltage insulators. The three phase 13.8 kV high voltage lines rise from the inverter/transformer package and connect to their respective phase feeder lines. These feeder lines run overhead to the switchyard.

3.2 Systems and Components

A simplified process diagram that identifies systems and components is presented in Figure 3.11. In this process diagram only one vessel of four is shown on the HP side of the plant and one vessel of five is shown on the LP side. The distribution of air and fuel within the vessels has not been shown, but is alluded to by the flow arrows branching from lines inside the vessels.

20 MWe PSOFC / GT Hybrid - Allison Turbine
SIMPLIFIED PROCESS FLOW DIAGRAM

Figure 3.11 — 20 MWe PSOFC/GT Hybrid Simplified Process Flow Diagram.

3.2.1 SOFC Generator

3.2.1.1 Staged-Cell Generator Concept Evaluation

Two separate analyses were made to evaluate the staged fuel cross-flow stack concept. The original analysis was the coupled thermal/electrochemical model and applied specifically to the tubular SOFC geometry. This analysis and its results are summarized below. The isothermal model was created to independently verify that the benefits of staging were much smaller than had been anticipated. The isothermal model is general and is not limited to a particular cell or stack geometry. The isothermal model is presented in the Appendix. The results of the two analyses are combined in the summary at the end of this section. The analytical models consider an SOFC bundle row consisting of four bundles, each bundle consists of a three-cell-in-parallel by eight-cells-in-series array. The bundle row is flanked by in-stack reformers.

3.2.1.1.1 Coupled Analysis of the Staged Fuel Cross-flow Stack

This section gives an overview of the coupled cell/stack analysis model used to evaluate the staged fuel cross-flow stack. The model was based on cross-flow of fuel through four bundles in series. Thus, fuel passes sequentially across a total of 32 cells. The cross flow stack analysis model was developed for a stack that would replace a conventional stack that has axial fuel flow. The staged-cell stack has recirculated fuel and is coupled to the in-stack reformer board (SRB) assembly positioned between each pair of bundle rows. Thus the cell configuration of the stack remains essentially unchanged, with the exception that fuel flows in at one side, through the stack, and exits on the other side. From there the fuel passes back through the recirculation zone to the inlet side of the stack where it is mixed with fuel feed and pre-reformed. The exhaust portion of the spent fuel exits to the combustion zone as it passes through the recirculation zone. A side view of the fuel flow through the stack and SRBs is shown in Figure 3.12.

Figure 3.12 — Staged-Cell Generator Concept.

The model is a fully coupled heat transfer, mass transfer, and electrochemical analysis. Local temperatures are used to evaluate all local properties including thermal conductivity, mass diffusion coefficients, electrical resistivity, and gas transport properties. Local temperatures and concentrations are used to calculate the Nernst potential, cell resistance, polarization losses, and the local current density. Fuel and oxidant streams are depleted based on local current density. Local heat generation includes the excess heat of reaction, Joule and polarization losses, and heating due to O₂ ionic leakage through the cell. Heat transfer coefficients are also based on properties evaluated at local temperatures. Radiation heat transfer is calculated based on the fourth power law. For all models that include an SRB, the temperature and the heat flux from the cell are matched with the temperature and the heat flux to the SRB. The reformation model of the SRB satisfies heat transfer, mass diffusion, reaction rate and chemical equilibrium conditions.

The development of the analysis model for the cross flow stack is based on a similar model that has been used for the conventional stack. To include a number of stages, the number of axial segments per stage between the closed end of the cell and the fuel recirculation zone was reduced to 20. Further, it was decided to divide the 32 cell stages into groups of cells. The cells in each group would be represented by the condi-

tions at the center cell of the group. The group arrangement selected is shown in Figure 3.12. To provide detail at the inlet fuel side of the stack, the first two groups include only one cell each. The third group includes 2 cells, the fourth group includes 4 cells, and the fifth and the sixth groups include 8 cells each. The seventh and eighth groups include seven cells and one cell, respectively. Note that the bundle and SRB boundaries coincide with group boundaries. This was required to simplify the coupling of the SRBs and the cells.

The initial results from the staged configuration analysis were disappointing. The problems can be described as follows.

1. The first cell row (at the fuel inlet) was too cold and final cell rows were too hot. This is a consequence of the large flow of relatively cold fuel that impacts directly onto the first stage or the first cell row. The latter cell rows become hotter as the fuel temperature increases and cell losses increase as fuel is depleted.
2. The overall temperature uniformity was poor. This is a result of the combination of the superposition of the row to row temperature distribution on the axial temperature distribution. The overall non-uniformity is the sum of the two.
3. The fuel concentration becomes non-uniform in the latter cell rows. The fuel is consumed at a higher rate at the hotter portions of the cell due to the lower local cell resistance. After passing a number of cell rows, the fuel becomes depleted in the central region of the cells. To preclude anode oxidation, the average fuel utilization at bundle row exit must be comparable to that for an axial-flow cell stack — a higher fuel utilization is not feasible.
4. The average cell voltage was lower than for an axial flow stack due to the lower overall average temperature.

Several modifications were made to the model stack to address the issues of temperature uniformity.

1. The inserts normally used to enhance heat transfer within the combustion zone region were removed from the air feed tubes in the last three cell bundles (24 cell rows). Removal of the inserts results in reduced pressure drop which acts to increase the cooling flow of air to the last 24 cell rows relative to the cooling flow to the first 8 cell rows. Removal of inserts also reduces the pre-heating of cooling air to the last 24 cell rows, thereby lowering their temperature relative to the inlet cell rows.
2. The fuel flow to the SRB adjacent to the first cell bundle was restricted by an orifice. This reduces the cooling of the first 8 cell rows at the inlet side of the

stack. It also increases cooling of the last cell rows by shifting the reformation heat load toward the exit side of the stack.

3.2.1.1.2 Results of Coupled Analysis of Staged-Cell Stack

Analysis results for the staged-cell stack with the modifications listed above are shown in Figure 3.13, Figure 3.14, and Figure 3.15. The cell current was 253 A/cell corresponding to a current density of 304 mA/cm². Air inlet temperature was selected to give a maximum local cell temperature of 1020°C (1870°F). The temperature distributions of the cells at the center of each of the 8 cell groups are shown in Figure 3.13. Figure 3.14 displays the average cell temperature and the terminal voltage as a function of the cell row position. The fuel concentration profiles at the exit of the 8 cell groups are shown in Figure 3.15.

The stack model modifications made significant improvements in the stack temperature uniformity, but the predicted performance of the cross-flow stack was still lower than for a conventional stack with the same maximum temperature limit. This result was obtained although several idealistic assumptions were made for the analysis which enhance performance relative to what would actually be expected. These assumptions included

1. The lateral heat losses at fuel inlet and exit were neglected. This would further reduce temperature of the first cell rows.
2. It was assumed that the fuel mass flow rate was uniform per unit axial length of the cell and bypass of fuel at the inactive ends of the cells was neglected. Actual fuel flow through the central portions of the cells would be reduced due to the effect of higher temperature by the combined effects of low density and high viscosity. This is the same elevation where the current density — hence, fuel consumption, is higher than the average rate.
3. The influence of dimensional tolerances on fuel flow distributions was ignored. The actual fuel flow distribution would be sensitive to the lateral spacing between cells and to the cell-to-SRB gap.

TCELL, Cell temperature distribution for each Group.

Figure 3.13 — Temperature Profiles in Staged-Cell Generator.

Figure 3.14 — Cell Voltage and Average Temperature in a Staged-Cell Generator.

Figure 3.15 — Fuel Mole Fraction Distribution in a Staged-Cell Generator.

Even with these assumptions, the fuel utilization in the staged cell design is limited due to the non-uniform consumption of fuel as it passes each cell. The fuel concentration at the exits of each stage at 92% fuel utilization is shown in Figure 3.16. In Figure 3.17, the partial pressure of O_2 at the fuel electrode is compared to the equilibrium partial pressure of O_2 for the $Ni/NiO/O_2$ reaction. The comparison is made for the exit stage cell. Oxidation of the fuel electrode is likely whenever the O_2 pressure is within two orders of magnitude of the equilibrium pressure. It is seen that this limit is exceeded for much of the cell length (data points are equally spaced).

Figure 3.16 — Fuel Concentration at the exit of each group of cells.

Figure 3.17 — Nickel Fuel Electrode Oxidation Limit – Cross Flow Stack at FU = 92%.

3.2.1.1.3 Summary of Staged-Cell Generator Concept Evaluation

The conclusion from the isothermal model analysis in the Appendix is that there is only a small improvement in stack efficiency available due to cell staging at current densities near the point of maximum power. At the maximum power point, the improvement in power output due to staging is less than 0.6% for 85% fuel utilization, and less than 1.2% for 95% fuel utilization.

In the coupled model analysis the small benefits of staging were eroded by the effects of fuel cross flow on the overall stack temperature distribution. The inlet fuel for 32 cells impinges on each of the cells in sequence. This high flow of relatively cold fuel removes several times more heat from the first cell row than is generated. Radiation from the neighboring rows provides a large amount of heat but it is insufficient to achieve satisfactory temperatures in the first rows. A way to improve the effects of cold fuel would be to heat the fuel exiting from the SRBs. We were unable to find a reasonable way to effect heat exchange between the inlet fuel and a hotter exhaust stream without incurring significant pressure drops and complicating the stack design.

The previously assumed benefit of higher fuel utilization with a staged-cell concept was not available due to uneven consumption of fuel in the direction perpendicular to the fuel flow. Fuel was locally depleted to the point of fuel electrode oxidation before the fuel utilization reached 95% even with the optimistic fuel flow distribution assumptions used in the model. Due to this effect and to the cool-cell effect in the fuel-entry rows, it was concluded that the SOFC generator module design should retain the conventional co-current fuel flow and air flow configuration.

3.2.1.2 SOFC Generator Design Description

Much of the HEFPP/SOFC Generator system design philosophy and implementation is influenced by technology derived from two sources: the conventional 100 kWe SOFC generator operating at atmospheric pressure and the pressurized SOFC generator in the 220 kWe hybrid cycle power system which is designed to operate at 3 atmospheres.

The generator submodule design philosophy is an extrapolation of the current 200 kW SOFC Generator configuration with the exception that the stacks are housed in a hori-

zontal pressure vessel. Each stack assembly is installed on a common removable support structure lined with high performance thermal insulation.

This configuration greatly simplifies assembly of internal components by utilizing common parts and a modular, reusable insulation system.

The design basis for the HEFPP SOFC Power Generation System is the air electrode supported (AES) fuel cell derived from the porous support tube (PST) cell. To date, PST test cells have achieved over 65,000 hours of operation. The AES fuel cell operating and durability history of over 32,000 hours reflects an improved design with significant advancements in material composition and processing.

Figure 3.18 shows a simplified array of closed tubular cells composed of concentric electrodes separated by a layer of solid electrolyte. Fuel for the electrochemical reaction flows upward between the tube exteriors, while process air flows upwards in the annular space between the air feed tube and the cell inner surface. Unreacted or spent fuel is burned above the bundles of tubes to preheat the incoming air. Nickel felt contacts provide flexible electrical connections between cells.

Figure 3.18 — Schematic fuel cell array.

Each substack contains 576, 22 mm OD, 1500 mm active length tubular fuel cells, each generating over 200 watts, arranged in 6 bundle rows. The cells are arranged in three parallel paths for improved reliability.

The submodule configuration is depicted in Figure 3.19. As shown, it includes ten fuel cell substacks, each fueled at the base by a fuel prereformer with integral fuel distribution manifolds and recirculation plenum, air manifold, DC power leads and stack internal insulation.

Figure 3.19 — Ten substacks isometric view.

Each substack is fed from a fuel supply system including a recirculation loop, ejectors, a prereformer and a fuel manifold with riser tubes. The recirculation plenum is used to mix the depleted fuel extracted from the stack with the fresh incoming fuel injected through an ejector nozzle. The mixture is then directed to a prereformer chamber where higher hydrocarbons are reformed to prevent carbon deposition in the stack reformers where full reformation of methane occurs. From the prereformer exit, the fuel mixture is distributed to each stack through a series of 7 bottom manifolds.

Process air is introduced into the submodule through one inlet nozzle connected to a centrally located air duct branching to individual smaller ducts feeding a large air plenum located on top of each stack. Each air plenum supplies air to 576 adjacent cells through ceramic air feed tubes. Air flows from the air plenum into the air feed tubes, which convey the oxidant to the lower, closed-end of each fuel cell.

Exhaust from the substack is collected from side manifolds and ducted through larger side ducts terminating in a common transition fitting connected to the pressure vessel exhaust nozzle.

Fuel Cell Substack. The tubular SOFC cell features a porous air electrode made of doped lanthanum manganite ceramic material. A gas-tight electrolyte layer of yttria-stabilized zirconia covers the air electrode, with the exception of an axial strip along the entire active length of the cell. This strip of bare air electrode material is covered by a thin, dense layer of doped lanthanum chromite. This layer, known as the cell interconnection, serves as the electric contact area to an adjacent cell or to a power contact. A top layer, the fuel electrode, is a nickel-zirconia cermet and covers the electrolyte surface except in the vicinity of the interconnection.

To construct an electric SOFC generator, individual cells are bundled into an array of electrically connected fuel cells, forming a monolithic structure that constitutes the basic generator stack.

The proposed fuel cell substack consists of a number of bundle arrays of 3x8 cells with an active (interconnection) length of 150 cm (59 in.). The cells are electrically interconnected in three parallel paths, each with 192 cells connected in series.

The three series-connected paths are intermittently connected by parallel shunts to improve reliability and gross electrical performance of the cell substack. An exploded view of the substack, as shown in Figure 3.20, is composed of 6 groups of cells, referred to as bundle rows, separated by electrically insulating in-stack reformers. Each bundle row includes 96 cells (three parallel paths of 32 series-connected cells).

These bundle rows are series-connected in a serpentine configuration and terminate at power take-offs that transfer current to the two opposite power leads. The metallic power leads may be air-cooled to withstand the high temperature environment surrounding the cell substack.

Each fuel cell substack is assembled externally on a common support structure prior to final installation into the pressure vessel.

Figure 3.20 — Exploded view of SOFC substack basic building block.

Stack Liner. Each cell substack is enclosed with a liner assembly composed by vertical panels made of Haynes® 214 alloy foil, 0.25 mm (0.010 in.) thick, and each panel is joined full length by resistance or fusion welding process to form four individual subassemblies which are welded at the four corner seams.

The liner assembly incorporates a bottom metallic sealing with the lifting tray and the fuel manifold penetration sealing typical of the 220kW PSOFC design. This type of design minimizes fuel leakage and it provides an effective barrier between the cell substack and the outer insulation made of low purity material.

Each panel includes a layer of “puff” board tiles bonded to the surface facing the substack. “Puff” boards are manufactured from layers of Saffil blanket material saturated with an organic binder and compressed to a 3.2 mm (0.125 in.) thickness to obtain the required density. During generator start up, the organic binder burns off and the Saffil layer is free to expand and fill the volume between the cell stack outer wall and the liner, thus preventing fuel bypass which may occur as a result of foil warpage under high temperature conditions. Each panel includes a series of “puff” board tiles, 305 mm (12 in.) wide, 406 mm (16 in.) high and 3.2 mm (0.125 in.) thick, glued to the stack with an organic adhesive.

Each liner panel includes two edge flexural flanges that are formed to a special shape designed to absorb thermal differential expansion of the panel in the direction normal to the flange length. The contour of the flexural flange is designed so that it achieves

maximum displacement, minimum stress when the panels are welded side by side to form a complete subassembly; these multiple ribs produce a substantial stiffness and load capacity improvement in the direction of the ribs. The ribbed or corrugated sheet basically exhibits one way flexural behavior, since the perpendicular direction retains the properties typical of a flat plate.

Fuel Distribution System. The Fuel Distribution System includes fuel nozzle housing, a recirculation plenum system, ejector, prereformer and an array of tubular manifolds connected to vertical risers penetrating the SRB's assemblies. An isometric view of the Fuel Distribution System is shown in Figure 3.21.

Figure 3.21 — Fuel Distribution System.

Fresh fuel is injected through an ejector nozzle that entrains depleted gas extracted from the upper zone of the fuel cell substack. This fuel mixture is directed to a pre-reforming section where partial reformation occurs within a catalytic bed.

The fuel stream then exits the pre-reformer and is manifolded through a piping network connected to horizontal manifolds installed at the base of the substack. These manifolds are coupled at the periphery of the stack by a quick connect flange assembly.

Each manifold includes a number of vertical risers that direct the fuel to the top of the in-stack reformer internal cavity. The preponderance of fuel reformation occurs in this area and a hydrogen-rich stream is fed at the base of the stack to the exterior of the tubular fuel cells. Complete reformation is finally achieved at the closed end of the fuel cell.

In-Stack Reformers. The in-stack reformer concept was developed to improve the SOFC system efficiency by reducing the airflow pumping requirements necessary to maintain cell axial temperature distributions within an acceptable range. Additionally, lower airflows reduce the required effectiveness (surface area) of the internal and external recuperators.

The SRB concept provides a large surface area for catalytic activity and heat transfer required for hydrocarbon reformation. It removes thermal energy uniformly along the entire cell length thereby resulting in sufficiently flat axial temperature distribution which helps maintaining acceptable axial temperature gradients within the substack. Moreover it has a much lower pressure drop than a conventional packed bed type external reformer.

The in-stack reformers each consist of a catalytically active inner assembly surrounded by a nickel foil liner, an outer board assembly and an embedded riser tube. Heat is radiated from the adjacent cell bundle region through the porous alumina outer board to supply the energy required for the endothermic reformation reaction.

As explained in the previous section, the in-stack reformers accomplish 85 to 100% of the reformation depending on the generator operating point.

The following is a summary of functional requirements of the in-stack fuel reformation system:

- control the cell axial temperature distribution to minimize voltage loss and maintain an acceptable axial temperature gradient.
- minimize the impact of prereformer catalyst replacement on system availability.
- provide adequate structural support for the upper air plenum and air feed tubes.
- assure electrical isolation to avoid short circuits within the stack or flash-over at the maximum voltage potential in open circuit.
- maintain reformer pressure drop consistent with the capability of the recirculation ejector.
- maintain adequate fuel flow distribution uniformity within the cell stack.
- minimize stack plus reformer size and weight.
- minimize reformation cost.

Air Supply System. Process air is introduced at the top of the generator submodule into an array of air plenums. Air flows from the air plenums into the air feed tubes, which are coaxial with the fuel cells and convey the air to the bottom of the cell. The air subsequently flows upward through an annulus between the feed tube and the cell inner surface.

Spent air exiting from the open end of the cells enters the combustion zone and reacts with spent fuel to completely consume the remaining fuel. Because the air feed tubes cross the combustion zone in a manner similar to a gas-to-air heat exchanger, incoming air is heated by the exhaust gas exiting the cell stack. The exhaust is directed to an upper dome prior to being manifolded to an external common exhaust duct.

High-pressure exhaust gas is collected from all submodules and is conveyed to the gas turbine through insulated piping. The turbine drives the compressor for compressing air that is delivered to the fuel cells.

Internal Thermal Insulation Package. To minimize heat losses from the SOFC stacks and at the same time maintain the pressure vessel wall temperature within reasonable limits, it is necessary to embed all the internals in thermal insulation without impairing the capability to easily service and replace internal components including the generator stack.

Typically, stack insulation within the stack liner is constituted by high purity alumina which exhibits excellent chemical stability at high temperature in both reducing and oxidizing atmosphere.

All elements of the stack insulation structure are typically vacuum formed by utilizing alumina fibers and binders in specific ratios to yield the desired thermal and mechanical properties. These elements include fuel manifold support boards, fuel distribution boards, in-stack reformer outer boards, stack end boards, lower/upper-positioning boards, recirculation and combustor boards.

The insulation package external to the stack liner consists of a number of highly efficient thermal insulation blocks and panels interlocked to form an effective thermal barrier system between the cell stack operating at 1000°C (1800°F) and the pressure vessel at ambient air. This insulation material is fitted between each substack and around all main manifolds and piping spool pieces.

The insulation blocks and panels are constructed with silica material that has a microporous structure and is reflective to thermal radiation.

An additional layer of low density fiber material is installed between the ceramic cell stack and the composite blocks to form a compliant layer capable of accommodating cyclic thermal differential expansion while offering effective bypass gas sealing and adequate lateral load support.

PSOFC Module Description. The pressurized SOFC generator module includes twenty substacks (576 cells) electrically connected in series/parallel to the module terminals. Advantages of this type of construction include ease of assembly, flexibility in power output requirements, better serviceability and substack replacement capability.

The substacks are installed within a pair of horizontal pressure vessels containing a supporting structure designed to provide easy access to the internals and easy assembling and replacement of the submodules without interference with the installed internal components.

Pressure Vessel. The pressure vessel, as shown in Figure 3.22, is a horizontal cylindrical shell and is supported by two saddles anchored to the shipping container structural frame.

Figure 3.22 — Pressure vessel isometric views.

Pressure bearing components are required to meet construction codes such as the American Society of Mechanical Engineers (ASME) B31.1 piping code or the ASME Boiler and Pressure Vessel (BPV) Code. The allowable stresses for materials currently approved for construction under the rules of Sect. VIII, Div.1 of the ASME BPV Code are provided in various Code Cases or in Sect. II, Part D.

The design of the components is in compliance with the ASME BPV Code, Section VIII, Division 1. The cylindrical shell material is SA-516, Grade 70 carbon steel which provides a minimum tensile strength 480 MPa (70,000 psi). The cylindrical shell thickness is determined by the tangential stress due to the design pressure. Since the maximum longitudinal stress ($PR/2t$) is only half of the maximum tangential stress, one-half the shell thickness is available for the longitudinal bending stress due to weight at the mid-span or in the plane of the saddles, assuming the vessel to behave as its own carrying beam.

As shown in Figure 3.22, two saddles including four legs and stiffener plates support the pressure vessel. The saddles are welded to the outer shell of the vessel with a 120-degree contact angle, and they could be anchored to a concrete slab in the field. The saddle reactions are highly concentrated and induce localized stresses in the shell, which are within the allowable stress conditions specified by the ASME Code.

The ASME flanged and dished head is cold formed from the same material as the pressure vessel. Each head will include one large central air inlet or exhaust outlet nozzle and smaller penetrations for the electrical DC power leads or for internal instrumentation cables.

The shell flange as shown in Figure 3.22, has an outer diameter of approximately 3.5 m (137 in.) and a length through the hub of 280 mm (11 in.) and is forged from SA-105 carbon steel material. Seventy-two 38 mm (1.5 in.) diameter radial holes are drilled through the flange for the bolts. The overall weight of each flange is approximately 2450 kg (5400 lb).

Bolting material for pressure connections must conform to the specifications listed in the ASME Code. Specifically, for this design, SA-193 Grade B7 (1 Cr-1/5 Mo) ferritic steel bolts have been selected. To minimize galling when the bolts are tightened, fasteners are made up with a thread lubricant such as Molykote paste. The proposed

bolting configuration includes 72 bolts with a 35 mm (1.4 in.) bolt diameter in a bolt circle of 3.4 m (134 in.).

Two of these pressure vessels bolted together at their flanges, as shown in Figure 3.23, form the basic building block of the power plant.

Figure 3.23 — PSOFC generator module isometric view.

A detail drawing of the pressure vessel and associated Bill of Materials is shown in Figure 3.24.

Internal Sub-module Supporting Structure. As shown in Figure 3.25, the pressure vessel incorporates internally a supporting structure that is embedded in high temperature insulation material. This supporting structure is composed by two tubular steel side members separated by a series of equally spaced high density ceramic boards over which a steel deck is finally installed.

Figure 3.25 — Stack support car.

The tubular members include a series of high temperature roller assemblies including a wheel, axle box; die steel caged roller bearing and high load rating axle. These wheel assemblies are typically used in furnace and kiln cars for a variety of heat treatment processes.

This rolling structure provides a convenient open deck for stack assembly, support for piping, ducting and electrical instrumentation without interfering with the assembly/disassembly of the stacks. Once the internals are fully assembled on the support structure, the entire assembly is rolled into the pressure vessel over two rails welded to the shell wall. Once in position, the wheels are locked and secured.

This method of supporting and transporting the stack submodules is particularly advantageous for in-situ stack replacements or quick field installation in the event that the pressure vessels are procured locally or are already installed at the power plant.

A complete assembly of the Module internals, including the SOFC stacks, ducting and supporting structure is shown in Figure 3.26.

EXPLD STATE: STACK_OUT

Figure 3.26 — Isometric view of Module with internal components.

Internal Air/Exhaust Ducting. When the substacks are inserted and positioned within the pressure vessel, they are connected to a common central air feed manifold by utilizing individual spool pieces connected to the inlet nozzle of each substack. The central manifold uniformly distributes the incoming pressurized process air to the individual substacks. The air manifold is positioned longitudinally within the pressure vessel, is supported by the ceramic frames and is embedded in the same insulation material surrounding the supporting structure. The final connection of the air manifold is made in correspondence to the side flanged cover nozzle through an expansion joint/adaptor.

The exhaust gas flow from each substack is directed to an exhaust collecting plenum overhanging each substack and is subsequently manifolded into a central duct exiting through the exhaust outlet nozzle.

The air ducting operates at a maximum temperature of 760°C (1400°F) and the exhaust ducting temperature does not exceed 870°C (1600°F). Because the internal differential pressure between process air and exhaust is less than 60 mbar (0.9 psi) as a result of only frictional losses and generator pressure drop, ducting rather than piping is utilized internally. Externally, all piping is fabricated to withstand a 10 barg (150 psig) design pressure.

Fuel Inlet Piping. The fuel inlet piping will operate at a pressure of 3.5 to 5 bar (50 to 70 psi) differential inside the pressure vessel and up to 12 to 14 bar (180 to 200 psi) externally. The final pressure will be determined on the basis of the selected generator operating range, fuel composition and gas turbine selection.

Electrical Interconnections. Internal electrical interconnections between stacks are accomplished through utilization of a main DC bus bar and flexible cable connectors clamped to each stack power lead.

High reliability electrical feed-throughs are utilized on the pressure boundary of the module in order to guarantee sound electrical connection between the internal stacks and the external power conditioning system.

Generator Instrumentation. The SOFC generator module is equipped with a variety of instrumentation providing for automatic control with manual capability for plant operation, monitoring and diagnostics.

The generator stack instrumentation consists primarily of DC voltage taps and cell stack thermocouples. Within the cell stack, there are a number of voltage taps monitoring the progressive buildup of accumulated cell voltages with the first tap near ground potential and the last at the maximum DC voltage. The stack external terminal voltage is also monitored as well as the main generator module terminals.

Several thermocouples are embedded within each cell stack at different elevations to monitor and control the temperature of the generator stack.

A number of pressure taps may be included into each stack fuel supply system to monitor differential pressures around the fuel ejector systems and to provide gas sampling as required.

3.2.2 Gas Turbine System

3.2.2.1 Introduction

Rolls-Royce Allison (RRA) subcontracted to Siemens-Westinghouse to examine the design of components relating to a turbogenerator for the bottoming cycle of a 20 MW SOFC power generation plant. RRA issued a preliminary report defining the geometry and performance of the intercooled compressor system and gasifier (HP turbine) and free power turbines (LP turbine) required. Subsequently, Siemens Westinghouse has since determined that a better system would be rendered by placing all the turbo machinery on one shaft.

In the original free power turbine configuration, the free power turbine speed was in no way compromised by matching to compressor requirements and was allowed to rotate at about 20% lower speed than the compressors, yielding an optimum low stress design. Rolls-Royce Allison has examined the possibilities for the single shaft machine and finds that it is appropriate to leave the compressor and HP turbine stage designs unchanged and use an unshrouded low aspect ratio second stage rotor blade in order to preserve conservative stressing.

To provide a single reference document for the RRA subcontract results, the preliminary compressor and turbine data, issued in October 1998, are reproduced herein with corrections and updates where appropriate. Other topics briefly discussed include the startup combustor, which energizes the turbines whilst the fuel cell stack exit temperature is too low, electrical system efficiency and general arrangement of the “turbogenerator.”

3.2.2.2 Plant Design Change

The original Siemens Westinghouse configuration called for an intercooled compressor driven by a gasifier turbine, followed by a power turbine, on its own shaft, driving a generator. A fuel cell stack added the heat necessary to drive the first turbine and a second stack reheated the working fluid between the two turbines. Siemens Westinghouse subsequently requested that the design be reconsidered, and adjusted where necessary, to be appropriate to placing all the turbomachinery on one shaft, without cycle change. Advantages of this configuration potentially include:

- Simpler mechanics (fewer total bearings)
- Better plant control due to ability to force the system
 - flow level to the optimum value by adjusting generator loading
- More freedom in cycle choice because the pressure level
 - level between stacks is no longer dictated by balancing
 - the first turbine work with total compressor work

A disadvantage of the system arises from the fact that the flow from the first turbine exit does not pass straight to the second turbine inlet, because it has to be diverted to be reheated in the second stack. The challenge for this system is it tends to increase the shaft length required to accommodate low loss ducting, hazarding shaft dynamic characteristics.

3.2.2.3 Design Approach

The Siemens Westinghouse cycle uses two fuel cell stacks in flow series in which hot air from the first stack powers the HP turbine and the second stack reheats the working fluid prior to entry to the LP turbine. The compressor system is intercooled at mid-point, enabling a slightly higher-pressure ratio to be used whilst retaining the correct heat balance around the cycle.

Though axial compressors are often more efficient than radial stages, this ceases to be true when the pressure ratio is low and the exit flow has to be diffused to near-zero velocity prior to the next flow process, as is the case for both compressors in this application (the first stage feeds to the intercooler, the second stage feeds to the recuperator). For the defined duties, it is clear that two low cost radial compressor will serve the purpose well. Where radial compressor stages are used in aero engines, they are of much higher stage pressure ratio.

There is no strong merit in considering a radial turbine because these really only come into their own as a means of reducing cost if one stage can replace two axial stages. In the proposed cycle, flow is reheated between turbine stages, requiring them to be separate, and each has a duty easily accomplished in a single uncooled axial stage.

The HP turbine stage geometry is somewhat like an aero HP turbine stage but runs at much lower temperature and is therefore not cooled and does not use the same high cost super-alloy. The LP turbine stage is determined to be best satisfied by a low aspect ratio unshrouded rotor blade that is not like a lightweight aero stage design.

During steady state running, the electrochemical fuel cell process keeps the stack and its throughflow at high temperature. Also, non-utilized fuel is reacted with residual oxidant in the exit section of the stack, raising gas stream temperature high enough to drive the turbine, approximately 870°C (1600°F). When the system is being started however, and the fuel cell stack is at ambient temperature, no galvanic action can take place and thus no heat is added or fuel burned within the stack. During this period, though the fuel cell is fed with hot pressurized air at inlet during this period, its exhaust is at low temperature and too cool to drive the turbine. A combustor burning natural gas is therefore required for system startup and is placed in the exhaust stream just before entry to the HP turbine.

3.2.2.4 Compressor Selection

The two compressor stages are very similar in duty, substantially differing only in required corrected flow size. In radial stages, the pressure ratio dominates the level of impeller rim speed required for optimum loading and efficiency. The axial inlet diameter is determined by:

- Good level of inlet air velocity/blade speed (for favorable aerodynamics)
- Avoiding excessive inlet tip relative Mach number
- Keeping the entry axial Mach number low enough to limit duct losses

For a given rim speed and initial area allowance for bearings, too high a value for RPM raises the impeller inlet tip relative Mach number and may lead to a need to reduce inlet radius, cramping the bearing design and increasing axial Mach number.

Noting that both stages have the same pressure ratio and similar inlet temperature is, it follows that they have similar impeller rim diameters. Inducer inlet inner radii have been held equal, which is compatible with most mechanical arrangements. Diffusion to low velocity is required at exit from both stages and so the performance is quoted at a nominal 0.15 outlet Mach number, i.e. including diffuser losses.

The rim speed of the two impellers is approximately 365 m/s (1200 ft/s), which is compatible with the use of aluminum or steel rather than relatively expensive titanium alloy. Compressor design point performance is summarized in Table 3.3 below, and compressor performance maps are provided in Figure 3.27 to Figure 3.30. Note that RC denotes pressure ratio and the curve parameter is shaft speed in percent.

Table 3.3 — Radial Compressor Stage Design Point Data

Stage	Inlet Temp °C	Pressure Ratio	Airflow kg/s (corrected)	Adiabatic Efficiency	Speed RPM
1	16	2.47	18.5	0.864	10,600
2	23	2.48	7.7	0.864	10,600

* Compressor corrected flow = $\frac{W \sqrt{(T/T_{ISA})}}{(P/P_{ISA})}$

** Definition of efficiency

Adiabatic efficiency = (Ideal rise of enthalpy from inlet to outlet pressure)/Unit work done

Outlet is defined as total pressure after diffusion to 0.15 Mach number

Working fluid is air with zero relative humidity

Windage is included in radial stages. Bearings not included.

Figure 3.27 — Compressor performance map, first stage.

Figure 3.28 — Compressor efficiency, first stage.

Figure 3.29 — Compressor performance map, second stage.

Figure 3.30 — Compressor efficiency, second stage.

In a high-pressure ratio multistage compressor, part speed operation is compromised because the working fluid is no longer fully compressed. This leads to the situation where the last stage is now too small to accept all the flow the first stage would pass when working at its design incidence. In consequence, the last stage works at negative incidence relative to design and the front stage works at positive incidence, tending towards stall. In a normal compressor, this mismatch is ameliorated by the fact that the temperature at entry to the last stage is falling with pressure ratio and its capacity therefore increases a little. As a result, for a simple $2.5 \times 2.5 = 6.25$ nominal design pressure ratio, given the relative flexibility of radial stages, it might be possible to handle the whole range of operation without variables.

In our case, intercooling prevents this capacity increase and the part-speed movement towards stall is much more rapid. The consequence is shown in Figure 3.31, which is the stacked map for the two-stage compressor system with the intercooler working at fixed effectiveness. It will be seen that the stall line is very steep such that stall will be encountered with very little speed reduction (operating line will pass through design point and run nearly parallel to last stage stall line).

Figure 3.31 — Overall compressor performance map.

To preserve part speed stability, interstage bleed will be required. Figure 3.32 shows that the stall line can be recovered to a good level using an interstage bleed scheduled to pass the amount of flow shown in Figure 3.33. This bleed impacts on plant overall efficiency at part power.

Figure 3.32 — Overall compressor performance map.

Figure 3.33 — Compressor bleed schedule.

3.2.2.5 Intercooler Selection

There are several possible configurations for the intercooling system. Air liquefaction plant intercooled compressor systems generally use air-to-water tube-in-shell heat exchangers adjacent to the compressors and then the heated water is passed to a large natural convection-cooling tower, or to a dry, forced-convection heat rejection unit, dumping the heat to atmosphere (in some plants this heat could gainfully be transferred to boiler feed water). This scheme is compatible with our observation that the potentially lower cost approach of using water as the final heat sink is usually rejected by customers who, almost universally, insist on an atmospheric sink, due mainly to permitting requirements.

Rolls-Royce Allison have studied intercooling systems for gas turbines and found that the ducting and heat exchanger pressure losses for the above systems tend to be 5%-8% in addition to the compressor exit diffusion loss. The parent company, Rolls-Royce plc, has developed a Naval engine (the WR21) which uses a close mounted, near axis-symmetric, heat exchanger array with about half this loss. The geometry is well suited to the gas path of radial stages. The drawback is that the heat exchanger modules have to be super-compact, with very small matrix passages. This results in their being expensive, sole-source and more subject to blockage. They are probably not cost-effective in this application.

3.2.2.6 Turbine Selection

In most cases, the highest efficiency configuration is served by the use of tip sealing shrouds on the rotor blades to limit tip clearance losses. The presence of a tip shroud implies a maximum tip velocity of about 460 m/s (1500 ft/s) to limit shroud cantilever and blade airfoil direct stresses. When the LP turbine was a separate stage, it was possible to run it a little slower than the other stages to keep tip speed down for comfortable mechanical loading whilst having the speed high enough to keep turbine aerodynamic loading low enough for high efficiency.

Full use of the highest permissible tip speed tends to lead to high hub/tip radius ratio that is undesirable in being very sensitive to tip clearance and is also inclined to reduce efficiency due to more dominant wall effects. The first stage was restrained not to ex-

ceed hub/tip radius ratio of 0.875. To ensure favorable design geometry, the flow coefficient (axial velocity/blade speed) was also kept above 0.3.

The original free second stage, which was a normal shrouded design, has been forced up by about 20% RPM for the single shaft. Blade hub and fixing stresses relate to the term AN^2 , chosen originally to be to be 4×10^{10} . "A" represents the axially resolved annulus area in square inches at exit from the last rotating row and "N" is the rotational speed in RPM. The product, AN^2 , has the dimensions of velocity squared and is a good measure relating to the maximum turbine tensile stress. "A" is set by the requirement to minimize outlet losses by reducing exit axial Mach number. Increasing the RPM from 8185 to 10,600 would have brought AN^2 up from 4 to 6.7×10^{10} .

This is too high for uncompromised aerodynamics and mechanics together and it was chosen to change to an unshrouded rotor blade to keep stresses low and to decrease the tip radius. The tip radius is reduced to bring the stage flow coefficient into an optimal range while preserving a low exit axial Mach number but the tip speed is 490 m/s (1600 ft/s), i.e. still in the range mechanically requiring an unshrouded design. By itself, the removal of the shroud reduces efficiency due to tip leakage effects, but the increased blade speed serves to decrease aerodynamic loading that has a slight positive effect on efficiency. Though the Total-Total efficiency actually improves a little (Table 3.4), the effective value after diffusion decreases is almost unchanged due to a small increase in exit dynamic head and consequent diffusion loss. The LP turbine stage exit diffuser is good enough to assume a 50% dynamic head loss.

Table 3.4 — Axial Turbine Design Point Data Summary, 850°C Entry Temperature

Press Ratio	RPM	Airflow kg/s (corrected)*	Adiabatic Efficiency**	No. of Stages
1st stage 2.066	10,600	7.1	90.7	1
2nd stage 2.096	10,600	17.5	92.3	1

$$* \text{ Turbine corrected flow} = \frac{W \sqrt{(T/T_{ISA})} \times F}{(P/P_{ISA})}$$

where

$$F = \sqrt{(\theta') / \delta'}$$

$$\theta' = \frac{2 \times R_{gas}}{62.23} \times \gamma / (\gamma + 1)$$

$$\delta' = 0.3698 A \times \gamma / (\gamma + 1)$$

$$A = (2 / (\gamma + 1))^{\gamma / (\gamma - 1)}$$

$$R_{gas} \text{ typical} = 53.3 \text{ ft-lbf/lb-}^{\circ}\text{R}$$

$$F \text{ typical value is } 1.00 \text{ to } 1.04$$

** *Adiabatic Efficiency = Work done/(Ideal fall of enthalpy from inlet to outlet pressure)*

*Outlet is defined as delivery at 0.2 Mach axial number exit
Working fluid as defined for cycle
No windage or bearing loss*

Turbine performance maps are provided in Figure 3.34 to Figure 3.37, and the single-shaft arrangement of the major gas turbine components is depicted in Figure 3.38. Note that the curve parameter in Figure 3.34 to Figure 3.37 is shaft speed in percent.

Figure 3.34 — HP turbine stage performance map.

Figure 3.35 — HP turbine stage efficiency.

Figure 3.36 — LP turbine stage performance map.

Figure 3.37 — LP turbine stage efficiency.

Rolls-Royce

Not to accurate scale

Figure 3.38 — Turbomachinery arrangement for Siemens Westinghouse 20 MW HEFPP Fuel Cell Plant.

More detailed views of the compressor air inlet sections are provided in Figure 3.39 and Figure 3.40.

To reduce weight in aero engines, and to minimize the amount of high cost alloy used in large turbines, it is frequent practice to use high aspect ratio turbine blades. Tip shrouds are essential to such blades because, without them, they are liable to a wide range of resonant vibration and flutter conditions, leading to fatigue failure. It is therefore important to ensure that the unshrouded blades are sufficiently low aspect ratio to have good vibration characteristics.

Taken too far, this can result in very pronounced individual blade pressure fields that also require large axial gaps between airfoil rows to prevent blade passing order vibration. Excess axial gaps increase wall friction losses and may compromise shaft dynamics, and the turbines have been laid out with this requirement. Unshrouding the LP turbine stage has decreased its total blade numbers and simplified manufacture, leading to a cost saving, which is incorporated in Table 3.4.

STAGE 1, 2.47:1 RC CENTRIFUGAL

Figure 3.39 — Compressor inlet configuration, first stage.

Figure 3.40 — Compressor inlet configuration, second stage.

The geometry proposed is based on general experience, from which we recommend a compromise 40 to 50 airfoils in a row for weight insensitive applications. These relatively low numbers favor cost minimization for the relevant size of turbine. There is insufficient time in this feasibility study to create a model of these blades and demonstrate analytically that the proposed geometry would be correct.

The exit axial Mach number from both turbines at design point is only about 0.2 corresponding to an exit dynamic head about 3% for the HP turbine and 4% for the LP turbine stages. The HP turbine stage has a compromised diffuser and a pressure loss of 3% should be assumed.

Note: Continued cycle optimization caused some final parameter values to differ somewhat from those reported in the above narrative. For example, the final compressor ratio is 7:1, vs. 6:1, which increased slightly the pressure ratio across each compressor stage. In addition, the LP turbine is loaded more heavily as a result of the HP turbine expansion ratio optimization. This resulted in a reduction of the LP turbine efficiency from 92.3% to 91.3%, the value used in the final power system performance estimates.

3.2.2.7 Combustor Selection

While the stack exit temperature rises from ambient to its full steady state value, the delivery temperature from the combustor is held constant, requiring progressively less fuel to be burned. Combustor stability is therefore required over a substantial range of fuel flow. The simplest, low cost, way to enable wide stable range for a gas combustor is to create a primary burning zone in the combustor in which the fuel richness varies from lean to greater than stoichiometric. This ensures that the gas flow can be turned down substantially before the whole volume capable of supporting a stable flame disappears.

The disadvantage of this approach is that it produces locally very high temperature and starts to generate NO_x. However, as the combustor is only fueled for a short time, it is probably not required to have ultra-low emissions. Drawing a balance between emissions and turndown, typically leads to a stable turndown range of about 1:4. If a greater range of turndown is required (to smooth the transition from combustor heating to galvanic heating) a relatively convenient arrangement is to select a combustor with

multiple liners and progressively shut them down to extend range, e.g. 3 liners would give 1:12 turndown. It would be convenient to use an existing developed gas turbine combustion system. To avoid complicating the hot ducting, the combustor is always in the main flow path even when it is not fired so material changes would probably be necessary for non-liner parts to be compatible with the high inlet temperature relevant to normal plant running. Durability of the system will be aided by the fact that the combustor will use the same desulphurized natural gas fuel that is consumed in the fuel cell.

Sizing the combustors is affected by the fact that their highest-pressure loss condition is when they are unfired and the system is up to full temperature. Clearly the pressure loss should be small at this condition to favor high plant efficiency and so the combustors should be sized to the full 870°C (1600°F) flow of the HP turbine stage. This corrected combustor flow level is about 7.3 kg/s (16 lb/sec) [compare with 6.8 kg/s (15 lb/sec) nominal for the 9-liner system of the Pratt & Whitney FT8 or 12.7 kg/s (28 lb/sec) for the 8-liner Rolls-Royce Avon industrial engines. Note that these figures are around 20% above their effective value because turbine cooling bypasses the engine liners].

In turbine engines, an important function of the combustor pressure loss is to give a positive pressure difference to turbine blade cooling film feeds and this is consistent with a typical design loss level of about 4% total pressure. In our application, no turbine blade cooling is required and the design pressure loss should be lower. Noting that pressure loss is proportional to velocity squared, the Avon system would give about

$$4 \times \{16 / (0.8 \times 28)\}^2 = 2\% \text{ pressure loss, which is a good level.}$$

The design of a combustion system can also be approached more fundamentally. To have good control of the combustion zones by conventional jet penetration, it is appropriate to design for a pressure loss function

$$PLF = 15$$

i.e., a total pressure loss of 15 x the dynamic head based on total chamber flow area. We therefore require an entry dynamic head only 1/1000th of the total pressure to have a 1½% loss. Loss variation away from design point will be in proportion to this entry dynamic head squared. If the combustor liner system is contained within a single cy-

lindrical outer casing, the radius required to achieve this dynamic head level is approximately 380 mm (15 in.).

The length of combustor is dictated by the uniformity of temperature required at delivery (presumed close to turbine inlet). Fair practice is to aim for the value of

$$\frac{T_{\text{max local}} - T_{\text{average}}}{\text{Combustor temp rise}} = 0.2$$

if cylindrical combustor liners are used, this requires 2½ diameters for burning and mixing.

If only a single liner was used, the required length would be 1.5 to 1.8 m (5 to 6 ft) which seems a bit bulky. The length can be reduced by filling the same outer case with a group of smaller diameter cylindrical liners having the same total area, e.g. three 432 mm (17 in.) diameter liners, which would require a length of only 1.1 m (3.6 ft). If the flow is delivered to an annulus rather than to cylindrical liners, the length can be further reduced to 0.6 m (2 ft). This compactness of the arrangements again points to the use of existing gas turbine systems.

The fuel nozzles can be of simple “showerhead” form with low-pressure drop. Natural gas is much easier to feed than kerosene because it does not require atomization. Premixing to attenuate emissions would detract from turndown capability.

Running at 870°C (1600°F), all the combustor and associated ducting will have to be contained in a thermal blanket, mainly to prevent overheating adjacent plant items.

3.2.2.8 Turbomachinery General Arrangement

Intercooled radial compressor sets, which feature 2 to 6 stages, with the relevant 1 to 5 intercoolers, are offered industrially by such manufacturers as Atlas-Copco. The cost of these units varies with detailed specification, including the issue of whether the compressors have to be customized. Referring to Atlas-Copco sales literature indicates that they might require a 3-stage unit or possibly a customized 2-stage unit to meet the required pressure ratio. An intercooled, motor-driven Atlas-Copco compressor system is depicted in Figure 3.41.

Rolls-Royce

Figure 3.41 — Example of multistage intercooled compressor set industrial process equipment by Atlas-Copco.

If this approach were adopted for the proposed HEFFP plant, a generator would replace the motor and the two turbine stages would be required to drive direct into the bull gear through a small pinion. In the long term, this is a rather expensive system without the possibility of eliminating lubricating oil. The mechanics of this arrangement would be slightly less satisfactory than the balanced multiple compressor system because the compressor and turbine gear tangential loads work in the same direction when their radial loads are opposed.

The aerodynamic designs put forward in this report are directed towards the single shaft configuration illustrated in Figure 3.38. Briefly, the features of this system are:

- The high-pressure compressor stage, being smaller, is placed on the end of the single system shaft. This aids the achievement of good shaft dynamics and permits the use of a simple cylindrical intake for this stage.
- The intake for the second compressor, having only a small internal pressure difference relative to ambient, can be a fabricated sheet metal box.
- Both compressors use radial vanes and an exit volute to diffuse the delivery flow into a pipe.

- So that both compressor impellers may be run at minimum clearance, they are both axially located by ball bearings a no-compromise high efficiency arrangement. To avoid conflict between the two, the compressor stages are joined with a piloted diaphragm coupling. This is a rather complex and expensive arrangement and more detailed study could show that relaxing to one ball bearing and a solid drive could be more cost-effective.
- The exhaust from the HP turbine stage must be turned for radial removal in a short axial length in order to minimize shaft length. The form of diffuser illustrated has been used in RRA small gas turbine practice for many years. It can be expected to lose nearly one dynamic pressure head (5%).
- The exhaust from the LP turbine stage is handled in a very conventional low loss diffuser, probably experiencing no more than 3% pressure head loss.
- The intercooler is of the tube-in-shell variety used on industrial process compressor sets.
- The combustor outline is based on the use of a Rolls-Royce Avon outline. Material for the outer casing and some other parts would have to be upgraded to Inco 718 or similar.

Keeping an eye on the future content of the powerplant, it is considered quite likely that the most attractive and cost-effective plant will be based on

- Direct drive alternator (5 MW @ 10,600 rpm). Motors of similar speed and power ratings are entering service.
- Power conditioning to 60 Hz grid connected output (in distributed generation).
- Oil-free bearings. Magnetic bearings are considered to be the most suitable. Their use will often be incompatible with using aeroderivative parts due to their rather large working area requirements.

It is expected that magnetic bearings direct drive alternators and inverters will all be introduced in more conventional gas turbine gensets. Part of the development process will be to bring the integrated system cost to a level competitive with, and therefore similar to, the current oil-lube bearing, gearbox and low speed synchronous alternator systems.

There are three recuperator types that can be expected to give good service

1. The traditional robust “chemical industry” tube-in shell design. For this flow class, typical practice has been of cylindrical units about 9 m (30 ft) long. They have normally been sized to considerably less than the 90% effectiveness level intended here.
2. Compact plate-fin designs, very much smaller than the tube-in-shell units, even when configured for 90% effectiveness. Past experience with this class of unit has been unsatisfactory due to unsubtle mechanical designs that permitted severe thermal stresses to prevail, leading to early failure in cyclic use. Northern Research Engineering Corporation (NREC) has recently reengineered this type and offers a unit that may well be satisfactory.
3. Primary surface recuperators have been developed over a very long period of time by Solar. These are even smaller than the compact plate-fin designs and have many features specifically included to avoid cyclic thermal stress problems. Solar have succeeded to the point where they offer these recuperators in their Mercury 50 product. In fuel cell plant application, relatively few temperature cycles are expected and the greater concern with these thin foil based units is long term creep closing down pathways and increasing flow resistance. The fuel cell plant temperatures are close to normal for these units.

Given that a recuperator effectiveness of at least 90% is specified for this application, it is suggested that the unit needs to be of compact form. We have examined an NREC-type arrangement based on individual cores of the dimensions shown below.

In order to reach 90% effectiveness for the 18 kg/s (40 lb/s) flow, we estimate that 5 of these cores will be required. On the exhaust side, the five cores are arranged in flow series, but on the air side a parallel configuration is employed, with the cores receiving cool air from and delivering hot air to common manifolds.

3.2.2.9 Gas Turbine Equipment Cost Considerations

The cost of turbomachinery depends strongly on its detailed design, particularly whether it is prepared for purely industrial purposes or for multipurpose duty including commercial aviation. In the most active sections of the industrial market, tight competition and purpose-designed industrial machinery tends to set the acceptable price level for units produced in substantial volume. At the flow required for the 20 MW fuel cell plant 18 kg/s (40 lb/s), a representative price for a complete simple cycle gas turbine is about \$150/kW at 1090°C (2000°F) turbine inlet temperature. Total generating set price is \$400/kW; i.e. the balance of plant costs \$250/kW, which is more than the engine.

The special configuration turbomachinery required for the fuel cell plant will be more expensive than normal competitive industrial turbines because

- The low turbine inlet temperature of 870°C (1600°F) reduces output to only about 70% of “normal” power. This alone scales turbomachinery cost to \$214/kW.
- Extra duct features required for intercooling and stack feed and return increase cost.
- The relatively small production volume does not justify so much production tooling. This and the extra duct features are assessed as doubling the cost to \$428/kW.
- In practice, program development cost recovery may also place a significant charge on each unit. This term has been ignored.

The balance of plant cost should largely remain in proportion to power output, thus the generating set skid is expected to cost $(428 + 250) = \$678/\text{kW}$. Given an output of 3.8 MW, the skid cost is \$2.58 M.

The turbine-based system cost in this plant includes two other substantial items, the intercooler and the recuperator.

We have examined intercoolers in other programs and consider that for this radial compressor application, only requiring moderate exchanger effectiveness, the most practical and quite satisfactory arrangement is a conventional water cooled tube-in-shell heat exchanger. Though the primary intercooler is not very expensive, in most arrangements a large water-air secondary heat exchanger is required. This can be forced or natural ventilation and, in combination with suitable ducting and plumbing, tends raise the price of a suitably sized system to around \$0.2 M.

There is little data available to us defining the price of modern compact recuperators but what we have suggests an allowance of \$0.83 M for the heat exchanging cores and \$0.2 M for associated ducting.

The cost of the turbine skid plus intercooling and recuperation is thus \$ 2.58 M + \$ 0.2 M + \$1.03 M = \$ 3.81 M or \$ 1000 / kW.

Intercooling and recuperation introduce additional maintenance requirements and it is considered that this will double running maintenance cost from a representative industry value of \$ 0.005/kWh to \$0.01/kWh.

3.2.3 SOFC Power Conditioning System

This project involved no task for the development of a power conditioning system (PCS) concept. For the power system performance estimates, the PCS efficiency, pertaining to the process between the SOFC DC terminals and the utility AC grid, was assumed to be 94%. This is consistent with current Siemens Westinghouse studies of mature power generation products to be offered circa 2010. The PCS equipment was costed at \$120/kW DC.

The 20 MW PSOFC generator plant will be made up of nine electrically independent operating PSOFC generator modules rated at approximately 2 MW, and one 4 MW Gas Turbine generator. Each 2 MW PSOFC module should have its own power conditioning system (PCS) to process the dc power. The power conditioning systems should be located immediately outside each of the nine pressure vessels. By converting the dc power to ac at the PSOFC pressure vessel, the length of the high current dc bus duct, and the number of high current dc electrical components (breakers, etc.) can be minimized. Medium voltage ac components are more readily available, smaller, and less costly than low voltage, high current dc components.

One gas turbine alternator, nominally rated at 4 MW, will provide the additional power to the system. The turbine system supplies compressed air to all nine PSOFC modules. The GT unit will be supplied with the gas turbine, the alternator, and the power conditioning system as a single unit. As such, no other power conditioning will be required. The unit will export power directly to the 13.8kV bus.

Operator control of the PCS systems will be provided through standardized interface ports. Interface ports for operation and system status/diagnostics will be made available at the control panel provided. Each PCS system should provide status readouts at the control center for operator review and intervention, if necessary.

The PCS should be configured to supply continuously adjustable current between 0 and 100%. The output power factor will also be adjustable from leading to lagging power factor. The PCS should be designed to tolerate some level of phase imbalance. The PCS will manage the export power based on the set points transmitted from the PSOFC control. In the event of a complete utility disconnect, one or more of the PCS systems will enter a stand alone operating mode. This mode of operation is required to maintain PSOFC auxiliary equipment in the event of loss of utility power, to avoid complete plant shutdown, until utility power is restored. However, the PCS units should be configured for bi-directional power flow, should utility power be required to support the PSOFC auxiliary system power bus.

Figure 3.42 is a simplified block diagram of the major electrical components of the 20 MW power plant, and shows the PCS system configuration. Only two of the nine parallel modules are shown. Included in the PCS system is the DC to AC inverter, and a step-up transformer. The DC to AC inverter converts the high current DC power into 480V, three phase ac for distribution. The transformer boosts the voltage for greater distribution efficiency and reduced bus conductor requirements. The nine parallel connected PSOFC PCS systems feed into a common ac bus for electrical distribution.

3.2.4 Instrumentation and Controls System

The PSOFC modules and auxiliary equipment will be equipped with instrumentation and controls that provide for automatic operational control, and with manual capabilities for plant operation. The instrumentation and controls (I & C) systems also provide for monitoring, data collection, and diagnostics. The I & C systems will be categorized into generator module instrumentation and auxiliary system instrumentation and control. Each of the nine PSOFC modules will have complete and independent I & C systems for individual control of each sub-system. Each I & C system will allow for remote operation with the operating data consolidated and provided with data displays at the main operator interface panel.

Figure 3.42 — Power Connection Block Diagram.

3.2.4.1 Hardware

PLC. The heart of the instrumentation and controls system is the programmable logic controller (PLC). The PLC provides safe, reliable, and steady operation of the PSOFC system over the entire operating range of the PSOFC generators. The controllers will scan inputs, execute preprogrammed logic, and set the state of outputs based on the control strategy. Depending on the final control architecture, one PLC could serve each PSOFC sub-system, with an additional PLC serving as a master controller.

Input/Output Modules. The I/O modules will link the instrumentation signals to the PLC. Thermocouples, voltage sensors, and current sensors will provide raw analog data through the I/O modules. The PLC converts the signals to digitized engineering units for display, or it provides an appropriate output response to various system actuators.

Communication Modules. The communication modules will link the master PLC to one or more PSOFC Control/Data computers. In addition, the communication modules will provide the control signals to the PCS for power control.

Electrical Control Hardware. The electrical control hardware is composed of those components that gather the input and output signals and respond to the needs of the main PLC process controller. Power supplies, signal conditioners, relays, current and voltage sensors, etc., comprise the electrical control hardware. These components form the control circuits that implement the commands from the PLC. The electrical control hardware also provides for continuous control of auxiliary devices such as blowers, valves, and the PSOFC generator/GT electrical loading. The components will be packaged for efficient space utilization and convenient maintenance access.

Operator Interface. An operator interface computer will be supplied as part of the I & C system, to provide a display of the critical PSOFC generator parameters, and to allow for operator input of some control parameter setpoints. The operator interface will consist of a computer with keyboard. A graphics display monitor, a large capacity hard drive, and a printer. Provisions for alarm monitoring/logging, operator input for state transition, manipulation of remote alarm action and control parameter setpoints will be incorporated into the operator interface.

3.2.4.2 Software

The instrumentation and controls system is a PLC based system that stores and uses information about the status of the PSOFC generator, the gas turbine systems, and the support systems. Data is read continuously and updated by the control software. Some of this data is also directed to, or taken from, the operator interface. The system software responds to the data, and controls the state of the generator. The system detects alarm conditions, and controls both modulating and discrete devices.

The PSOFC generating system will have a table of pre-defined operating states. For example, Pre-Op, Purge, Heat, Load, Run, Cool, Stop, and Sstop are typical PSOFC system states. The control system software contains the logic that determines the operating state of the PSOFC generators and all valid transition states. The control system will allow the operator to transition from the present state to a valid transition state only.

Discrete control logic will be used to change the states of control solenoid valves, and to provide valve status information to the operator.

Analog control logic will be set site alarm limits, and detect out of limit alarm signals. Response to the alarm signal will depend on the severity of the alarm. All alarms warn the operator. Others may result in a shutdown of the system.

Operator interface software package will be provided for supervisory control and data acquisition.

3.2.5 Electrical Distribution System (EDS)

The PSOFC/GT electrical distribution system links the PSOFC modules and the gas turbine systems to the PCS, and the PCS to the utility ac power grid. Included in this link are the bus leads, all of the power monitoring equipment, disconnect switches, and protective devices. A step up transformer is supplied as part of the PCS to elevate the output voltage before it is routed to the switchyard. At this switchyard additional step-up transformers raise the voltage as necessary for export to the utility grid. The disconnect switches will be strategically located for safe operation and maintenance of the PSOFC generating plant. Fault detection equipment will be provided, to sense utility grid under voltage, over voltage, and off frequency conditions.

There is also a requirement in the electrical distribution system for a 480 vac, 60 Hz bus. The bus will be fed by a step down transformer from the primary bus. The 480 vac bus is required to supply three-phase power for the PSOFC system auxiliary equipment, and for backup (UPS) power supplies. The electrical distribution system will have service breakers for individual sub-systems, for maintenance and for electrical protection capability in the event of a sub-system fault. A power distribution panel will be located within the PSOFC system to distribute the power to the internal system loads.

The electrical power from each of the PSOFC modules can be exported to the utility grid via a 13.8 kV bus if the adjacent grid lines are at this voltage. Otherwise, step-up transformers in the switchyard are used to match voltages. A static isolator will be provided between the high voltage bus and the grid interconnection to allow for quick disconnect, in the event of a fault, either on the utility grid, or on the PSOFC generating system. Individual PSOFC sub-modules will be protected by three phase circuit breakers.

The performance of the electrical distribution system is closely monitored and controlled by the instrumentation and control system. The I&C system provides the supervisory functions for power flow and fault conditions for each PSOFC sub-system and the gas turbine systems. The major power components of the electrical distribution system are shown in Figure 3.42.

3.2.6 Fuel Processing System

The supply of natural gas is assumed to be within the conventions of the U.S. distribution system. In the U.S., the distribution mains range in size from 0.508 m. (20 in.) to 1.07 m. (42 in.) and contain gas at 13.8 barg (200 psig) to 103.4 barg (1500 psig). Compressor stations are normally located 80 km (50 miles) to 100 km (60 miles) apart and occasionally as far as 290 km (100 miles). Gas flows in the mains at 24 km/hr (15 mph or 22 fps). The natural gas is reduced in pressure at the local natural gas supply utilities. These utilities add an odorant and distribute the gas through lines at about 4.1 barg (60 psig). The pressure at residential customers varies from 4 barg to 17 mbarg (60 psig to ¼ psig). The most commonly used gas piping is rated to operate at 6.9 barg (100 psig). For this study, it is assumed that the supply of natural gas at the plant site

is at 1.03 barg (15 psig) and contains a sulfur-based odorant. The gas line size needed for a 20 MW plant is 30.5 cm (12 in.) at 1.03 barg (15 psig) supply pressure.

On the plant site, the natural gas is filtered in an automatic duplex inlet filter. One side of the filter is on standby while the other side performs the filtering. When the differential pressure across the filter reaches the set pressure corresponding to a dirty filter element which needs to be changed, the automatic duplex filter uses the gas supply pressure for the motive force to actuate the three way valves and divert the flow through the standby side. A dirty filter indication is sent to the control system and filter element changing is scheduled. Filter element changing is performed with the unit on line and fuel flow is not interrupted.

The natural gas enters the inlet of one 100% capacity natural gas compressor. The compressor discharges to an accumulator which provides a pulsation dampening function as well as a reservoir of natural gas at compressor discharge pressure to ride through supply interruptions when the compressor stops and goes off line when the accumulator high pressure set point is attained. The compressor operates on accumulator pressure control, starting on low pressure and stopping on high pressure. The compressor design flow rate is set slightly higher than the maximum plant fuel consumption rate. During normal plant operation, the compressor will be idle about one fourth of the time.

From the natural gas accumulator, the fuel passes through a gas recuperator. Hot natural gas from the outlet of the desulfurizer system exchanges heat with the cooler natural gas leaving the natural gas accumulator. This preheated natural gas then leaves the gas recuperator at a temperature near 378°C (710°F).

The natural gas desulfurizer includes two 100% capacity each adsorbent containing vessels operating in series. At the entrance of the first vessel, a very small quantity of hydrogen, about 1% (by volume) of the fuel flow, is mixed with the natural gas. The gas mixture is electrically heated to raise its temperature to near 400°C (750°F). Each vessel includes an electrical gas heater under automatic temperature control to ensure that the zinc oxide bed and natural gas are at suitable temperature for effective sulfur removal.

The gas mixture first comes into contact with a cobalt-molybdenum catalyst where conversion of the complex sulfur compounds into H_2S occurs. The gas then passes into a bed of zinc oxide adsorbent. The adsorbent is ineffective for sulfur removal at ambient temperatures, but is very effective at temperatures above $277^{\circ}C$ ($530^{\circ}F$). Pre-heating of the natural gas/hydrogen mixture in the gas recuperator followed by supplemental electrical heating in a gas heater to the optimum operating temperature of $400^{\circ}C$ ($750^{\circ}F$) permits use of minimum sized adsorbent beds.

Sulfur removal effectiveness is periodically (monthly) checked after drawing a sample from the line between the two-desulfurization vessels. When the sample shows detectable sulfur, the valves are realigned to isolate the first vessel and pass the flow through the second vessel. After the zinc oxide adsorbent is replaced, the vessel is put back into service and aligned as the second in series. In this manner the adsorbent is fully consumed while ensuring that continuous desulfurization of the fuel occurs.

3.2.7 Gas Supply Systems

The plant includes both nitrogen and hydrogen generating equipment. The nitrogen is stored in pressurized cylinders until needed. The hydrogen is produced as demanded and not stored, thereby reducing fire protection concerns.

3.2.7.1 Nitrogen Supply System

The inlet to the natural gas compressor accumulator has a connection to a nitrogen purge system. The nitrogen supply system is normally isolated from the natural gas supply and is activated under two conditions:

- Emergency or safety stop (SSTOP)
- Maintenance of natural gas system components

When activated, the nitrogen purge system connection isolation valve and gas vent valve both automatically open. Nitrogen gas from a pressurized tank flows through the accumulator, gas recuperator, desulfurizer vessels and outlet duplex filter to the vent.

This dilutes and forces the natural gas out of the system until a nonflammable concentration of gas remains.

SOFC Cover Gas. The SOFC stacks require an inert or reducing environment on the fuel side of the cells when the temperature approaches and exceeds 600°C (1100°F). During normal operation, a reducing environment is maintained by reformation of the natural gas fuel by recirculation of a portion of the spent fuel back into the fuel ejector where it passes through the pre-reformer and is distributed into the in-stack reformers. Here steam reformation converts the methane and higher hydrocarbons to hydrogen and carbon monoxide.

During startup when the stack temperature is low and fuel recirculation will not support methane reformation, a nonflammable, inert nitrogen cover gas is used to provide the required atmosphere. A commercially available nitrogen generating system produces this cover gas. This system uses a selective membrane to separate nitrogen from the other constituents of air. This nitrogen cover gas system operates at plant startup and also when the plant is shut down.

A pre-packaged, skid mounted nitrogen generation system produces nitrogen gas whenever demanded by the low-pressure control in the nitrogen storage accumulator. The system stops when high pressure in the accumulator is attained. During normal plant operation there is no need for nitrogen. The nitrogen is available upon demand for plant start or cool down. The nitrogen generating equipment may occasionally operate to refill the accumulator if there are any leaks. The nitrogen system is sized to produce in seven days a sufficient quantity of nitrogen to support an unplanned emergency plant shut down. The storage accumulator has provisions to be filled with nitrogen from an external source, if necessary.

The nitrogen production is in commercially available equipment that consumes electricity and compressed air to generate nitrogen gas. The nitrogen is separated from the compressed air through a selective membrane. The nitrogen is 98% pure. Nitrogen flows under its own pressure from the accumulator without the need for additional pumping. When used as a cover gas for the SOFCs, some excess hydrogen (about 2% by volume) is added to scavenge the residual oxygen which remains in the nitrogen.

3.2.7.2 Hydrogen Supply System

A pre-packaged, skid mounted hydrogen generation system produces hydrogen continuously during plant start and normal operation. During normal operation the hydrogen is mixed into the natural gas to form a mixture which is about 1% hydrogen. This mixture is heated and delivered to the cobalt-molybdenum catalyst in the fuel desulfurization vessels. During plant start when fuel is supplied to the SOFC generators the hydrogen is also needed for catalysis of the natural gas.

The hydrogen production is in commercially available equipment that consumes electricity and electrolyzes water to generate hydrogen gas. This gas is drawn from the electrolysis chamber and immediately compressed. Hydrogen is produced as needed for desulfurization of natural gas. The hydrogen generator is sized to provide 100% of the hydrogen needed for catalysis of sulfur compounds in natural gas on the cobalt molybdenum catalyst.

3.2.8 Balance of Plant (BOP) Equipment

3.2.8.1 Startup Boiler

A natural gas heater boiler is supplied to provide a source of steam for plant start up. This commercially available, prepackaged, skid mounted boiler is a self-contained complete system with all-auxiliary components and instrumentation. Integration into the SOFC plant requires connection of natural gas and demineralized water piping and electrical power cable. The steam outlet is piped to the SOFC vessels and the boiler blow-down line is piped to a drain.

Boiler operation is manually controlled for the short time that it is needed. The boiler is used only for plant startup and is drained and laid up dry for a prolonged period of inactivity during normal plant operation.

3.2.8.2 Auxiliary Air Compressor

The auxiliary air compressor system includes an internal combustion engine driven compressor, inlet air filter, compressed air filter/dryer and air accumulator plus the necessary engine starter, instrumentation and controls. The dual fuel engine can use either natural gas or propane gas. The primary engine fuel is the normal natural gas supply.

In the event of loss of natural gas supply, the engine automatically switches over to use propane from a pressurized tank that contains liquid propane.

The compressor operates to fill an air accumulator. The compressor engine starts and stops based upon pressure control signals from the air accumulator. The auxiliary air compressor is sized to continuously provide all of the necessary airflow required to protect the airside of the SOFCs after a shutdown. This flow of air-cools the stacks from normal operating temperature to less than 260°C (500°F) in about three hours.

The compressed air accumulator provides a pulsation dampening function at the compressor outlet and permits supply of an uninterrupted source of auxiliary air when the engine has an interruption due to automatic fuel supply changeover.

3.2.8.3 Water Treatment and Storage

Potable water is taken from a local source, filtered, deionized and stored in a day tank. The day tank is sized to provide the water needed by the hydrogen generating system and startup boiler system during a normal plant start. The day tank is automatically filled on level control. Municipal water supply pressure is used to fill the tank.

The potable water system includes a pump to transfer the deionized water from the day tank to the hydrogen generating system and startup boiler.

3.2.8.4 Process Air Piping

The design of the process air piping from the main compressor to the SOFC vessels and back to the turbines has to satisfy several criteria. To satisfy these criteria, the process air piping is internally and externally insulated.

- The piping wall temperature should be less than or equal to 371°C (700°F). This criterion allows the use of standard piping material instead of higher cost alloys rated for high temperatures. All standard piping steels suffer no loss of strength until the working temperature exceeds 400°C (750°F).
- The external insulation jacket surface temperature is less than the OSHA 29CFR1910.107(l)(7) limit of 65.6°C (150°F). Conventional jacketed, mineral wool is suitable for this application.

- The speed of the air inside the inner insulation is less than 30 m/s (100 ft/s). This air speed does not require an internal protective liner for the insulation and permits the use of less expensive, commonly available insulating materials without concern for erosion of the surface. Vacuum formed alumina-silica fiber insulation was found to have good thermal resistance and appropriate abrasion resistance.

To satisfy the piping criteria, the internal insulation is approximately 19.4 cm (7.63 in.) thick and the external insulation is 7.6 cm (3 in.) thick.

3.2.9 Plant Operation

Plant operations are summarized in the following sections. These summaries include normal plant startup, normal operation, normal shutdown and emergency shutdown.

3.2.9.1 Preparations for Normal System Start

The following services are available:

- Natural gas compressor is operational and gas supply pressure is in range
- Auxiliary air compressor is operational and compressed air accumulator pressure is in range
- Demineralized water day tank is filled to required level with deionized water
- Nitrogen system accumulator pressure is in range
- Hydrogen system is operational

System Configuration:

- Initial system status — unpressurized at ambient temperature
- Fuel piping — fuel supply control valves for all SOFC modules are closed. Fuel supply control valves for both air heaters are closed. Combustor fuel supply is off. Desulfurizer vessels' vent to atmosphere is closed.
- Vessel fuel vent system piping — fuel vent system isolation valves from each SOFC module are closed.
- Auxiliary air compressor piping — Isolation valves to SOFC module air supply headers are closed.

- Cover gas piping — Isolation valves to both SOFC module fuel supply headers are closed.

Operator Actions Sequence. Start turbine air compressor motor. When airflow through turbine is established, fire the combustor to bring the turbine/compressor assembly up to normal speed. Allow the turbine/generator to free wheel as the flow of air passes through the LP SOFC modules to the LP turbine inlet.

Fire the air heater on the HP SOFC module inlet header. Control firing to limit the air heater exit air temperature to 500°C above the lowest temperature HP SOFC vessel stack.

Fire the air heater on the LP SOFC module inlet header. Control firing to limit the air heater exit air temperature to 500°C above the lowest temperature LP SOFC vessel stack.

When the highest temperature HP SOFC module stack reaches 500°C, open the cover gas system supply isolation valves to admit inert gas into the header which feeds the HP SOFC vessels.

When the highest temperature LP SOFC module stack reaches 500°C, open the cover gas system supply isolation valve to admit inert gas into the header that feeds the LP SOFC vessels.

Begin adding steam to the SOFC generators, via the fuel supply system, when combustion zone temperatures reach 600°C.

When the SOFC vessels reach 675°C stack temperature, begin to draw current. Start low fuel flow. Slowly increase the power being drawn from the stack. Fuel flow increase will automatically follow current increase.

Isolate cover gas system from the SOFC module headers when the current from each stack reaches 300 A. Fuel reformation and recirculation is sufficient to maintain a self-sustaining reducing atmosphere in the fuel region.

Begin to load the turbine generator. Increase the power drawn as the temperature increases in order to maintain airflow (shaft speed) at design value(s).

Turn off air heaters when the current from each stack is approximately 460 A. At this time the recuperated air temperature is self-sustaining.

Turn off the turbine combustor.

3.2.9.2 Normal Operation

The plant run state is entered when the stack temperature achieves 850°C and the current draw is 300 A from each SOFC module. During normal operation, the operator has control of:

- Fuel utilization
- Current
- Stack temperature

Setting the control system set points for any two of these parameters determines the third.

The plant is expected to operate unattended for 50 weeks with periodic remote monitoring and control adjustments made via telephone modem access for a remote location. At the end of the scheduled normal time, operators return to the site to supervise a normal shutdown.

While most maintenance will be performed during the annual scheduled plant outage, some activities may need to be performed during normal operation. The replacement of natural gas desulfurizer adsorbent will normally be performed during the annual maintenance period. In the event that the first bed in series does not last until the end of the 50 weeks of operation, it is isolated and all flow diverted through the remaining bed. After the bed has cooled and the adsorbent replaced, it is put back on line as the second in series. Similarly, the duplex filters on the inlet and outlet of the desulfurizers will normally last for years before developing enough filtrate cake to produce the pressure drop to trigger automatic change over. The duplex filter changes position to pass flow through the standby filter when across the filter media increases to the dirty filter set point. Spent filter elements will be changed as the need arises.

3.2.9.3 Normal Shutdown

Start Cover Gas System. When cover gas system is fully operational, the turbine AC generator is unloaded.

Shutting down the SOFC modules is accomplished with simultaneous control actions to:

- Stop natural gas flow to the modules
- Start cover gas system flow to the modules
- Unload current to ~35 A per stack and transfer power to stack energy dissipaters

The turbine will continue to rotate and produce compressed air flow, running on the stored heat from the SOFC modules. After system cooling has proceeded and the turbine/ compressor approaches stall, the generator will be used as a motor to drive the compressor. An alternate, but slower, way to cooling the plant is to open the auxiliary air supply isolation valve to permit flow of air into the HP SOFC modules from the auxiliary air system.

Stop flow of cover gas system when the stack temperatures are lowered to 500°C.

Continue supply of auxiliary air until stack temperature is reduced to 100°C.

Automatic Stop Conditions. The plant will automatically enter STOP when any of the following conditions are sensed:

- Low generator voltage
- Low generator current
- High stack temperature for 10 minutes
- Low UPS battery voltage

The sequence of automatic shutdown actions is the same as if the shutdown were initiated by manual control action.

3.2.9.4 Normal Maintenance

At the end of a normal run time, the plant is shut down for normal maintenance. The maintenance period is two weeks. During this period the activities include:

- inspection of turbine and generator rotating parts
- lubricating oil changes
- air filter element replacement
- instrumentation and control system calibrations
- repair or replacement of failed or improperly operating components
- fire protection system inspection and testing

3.2.9.5 Emergency Operations

To accommodate the occurrence of abnormal situations, the plant control system is equipped to automatically take control actions. The priorities for automatic emergency control actions are 1) personnel protection and 2) property protection. The extent to which emergency conditions are accommodated is indicated in Table 3.5.

It is assumed that no two failures occur simultaneously.

Table 3.5 — Plant Emergency Situations

Plant Mode	Control Sequence			
	Loss of Grid	SOFC Failure	Turbine or Generator Failure	Cover Gas System Failure
Startup	Start auxiliary air compressor Enter aux. air stack cooling sequence Maintain cover gas flow until stacks are less than 500°C	Enter normal stop sequence	Start auxiliary air compressor Enter aux. air stack cooling sequence Maintain cover gas flow until stacks are less than 500°C	Depressurize Vent Fuel Side Start primary air cooling sequence Reestablish purge with nitrogen
Normal Operation	Divert turbine/generator power to resistance banks, SOFC to open circuit.	Enter normal stop sequence	Start auxiliary air compressor Enter aux. air stack cooling sequence Start and maintain cover gas flow until stacks are less than 500°C	N/A
Normal Shutdown	Start auxiliary air compressor Enter aux. air stack cooling sequence Maintain cover gas flow until stacks are less than 500°C	Continue normal shutdown sequence	Start auxiliary air compressor Enter aux. air stack cooling sequence Maintain cover gas flow until stacks are less than 500°C	Depressurize Vent Fuel Side Start auxiliary air compressor Purge with nitrogen

3.3 Power System Installed Cost and Cost of Electricity Estimates

Table 3.6 summarizes the development of the power system installed cost, and in Table 3.7, the system COE estimate is developed. Mature-product conditions (circa 2008) were assumed. SOFC power conditioning equipment was costed at \$120/kW. Balance-of-plant equipment costs were obtained from potential suppliers; supply rates consistent with 10 to 100 power systems/year were assumed. The transportation cost estimates were developed consistent with Siemens Westinghouse transport experience with recent SOFC demonstration units; the transport distance was 500 miles per the design requirements. The site preparation, project management, and equipment installation cost estimates were developed by ICF Kaiser Engineers, Pittsburgh, PA, under subcontract. Input for the COE analysis is identified in Table 3.7.

Table 3.8 provides the COE estimate for the conventional power generation technology, a 20 MWe-class gas turbine/steam turbine combined cycle. For the reference fuel cost, \$3.00/MMBtu, the HEFPP COE estimate is higher than the conventional-technology

COE estimate by approximately 3%. Better COE performance is achieved when the HEFPP system operates in a higher-fuel-cost environment. For example, with \$6 fuel, the HEFPP COE estimate is less than the conventional-system COE by approximately 7% (66 mills/kWh vs. 71 mills/kWh), and for \$9 fuel, it is 13% less (83 mills/kWh vs. 95.0 mills/kWh).

Table 3.6 — Power System Installed-Cost Estimate

Installed Equipment Costs						
	Equipment	Freight	Installation	Totals		
SOFC Generator	8,890,422	31,500	47,365	8,969,287		
Gas Turbine System	3,960,682	3,500	59,347	4,023,529		
SOFC Power Conditioning System	1,988,520	15,750	24,374	2,028,644		
Instrumentation, Controls, and Electrical Cabinets	877,542	7,000	199,520	1,084,062		
Switchyard and Electrical Distribution	959,600		237,980	1,197,580		
Fuel Supply System	167,091	1,750	10,000	178,841		
Hydrogen Supply System	89,779	1,750	10,000	101,529		
Purge Gas Supply System	120,520	1,750	10,000	132,270		
Auxiliary Air Supply System	179,723	1,750	6,510	187,983		
Startup Boiler System	74,884	1,750	1,316	77,950		
Piping and Insulation	1,608,054	15,750	317,649	1,941,453		
Site Buildings				36,159		
Totals	18,916,818	82,250	924,061	19,959,288		
Project Cost Summary						
Installed Equipment						19,959,288
Project Management, Engineering, and Permitting						919,369
Site Preparation						412,994
Grading, utilities installation					145,744	
Foundations installation					217,519	
Structural steel installation					49,731	
G&A, R&D, Sales & Marketing, Profit Allowance						5,544,303
Total Plant Cost						26,835,954
Spare Parts Allowance						246,914
Startup						150,000
Land						20,000
Total Capital Requirement						27,252,868

As indicated in the trade-study discussion, Section 3.4.1, a hybrid power system with a lower installed cost and lower efficiency could be configured that would, in the \$3 fuel case, have a lower COE than the high-efficiency power system because the system capital cost is the dominant COE parameter. With higher fuel costs, that would not be the case, and the higher-cost, higher-efficiency power systems would become more attractive.

Table 3.7 — Power System Cost of Electricity Estimate

COE Calculation Basis		Cost Summary	
No. round-the-clock power system operators	1		
No. plants on system	5	Fixed O&M, \$/kW-year	
Operator labor cost, \$/man-hour	75	Plant operation & control	6.9
Housekeeping maintenance, man-hours/week/system	20	Housekeeping maintenance	1.4
Housekeeping labor cost, \$/man-hour	25	Administration (30% of operation & maintenance labor)	2.5
System rating, MW net ac	19.0	Total Fixed O&M, \$/kW-year	10.7
Gas turbine rating, MW ac	4.1		
SOFC module rating, MW dc	16.571	Variable O&M, mills/kWh	
Gas turbine maintenance cost, \$/GT kWh	0.01	SOFC replacement	2.1
SOFC replacement cost, \$/SOFC generator module	468,920	Gas turbine maintenance	2.1
SOFC replacement interval, years	10	Desulfurizer adsorbent/catalyst replacement	0.1
Power plant capacity factor	0.92	Total Variable O&M, mills/kWh	4.3
Desulfurizer adsorbent & catalyst, \$/year	8,919		
Interest rate (SOFC replacement cost calculation), %	6.0	Total O&M COE, mills/kWh	5.6
Power system efficiency (net AC/LHV), %	67.3	Capital COE, mills/kWh	26.6
Capital charge rate, %/year	15	Fuel COE, mills/kWh	16.9
Power system capital cost, \$/kW	1,431	Total COE, mills/kWh	49.1
Fuel Cost, \$/MMBtu	3.0		
		COE, Relative to Conventional Technology	1.032

Table 3.8 — Conventional-Technology Power System Cost of Electricity Estimate

Information Source: Gas Turbine World 1997 Handbook, p. 24			
System Classification: Gas Turbine/Steam Turbine Combined Cycle			
Gas Turbine: General Electric LM 1600			
Steam Turbine Cycle: Two-pressure, 4.8 MWe			
Net System Power Output: 17.9 MWe			
System Efficiency: 47.9% (net AC/LHV)			
Turnkey Price: \$14.9M (\$830/kWe)			
Basis		COE Summary	
No. round-the-clock power system operators	1	Fixed O&M, \$/kW-year	
No. plants on system	5	Plant operation & control	7.3
Operator labor cost, \$/man-hour	75	Housekeeping maintenance	1.5
Housekeeping maintenance, man-hours/week/system	20	Administration (30% of operation & maintenance labor)	2.6
Housekeeping labor cost, \$/man-hour	25	Total	11.4
Power system maintenance, \$/kWh	0.007		
Power system capacity factor	0.92	Variable O&M, mills/kWh	7.0
Power system capacity, net AC kWe	17900		
Power system efficiency (net AC/LHV), %	47.9	Total O&M COE, mills/kWh	8.4
Capital charge rate, %/year	15		
Power system capital cost, \$/kW	830	Capital COE, mills/kWh	15.4
Fuel cost, \$/MMBtu (HHV)	3.00		
		Fuel COE, mills/kWh	23.7
		Total	47.6

3.4 Conceptual Design Trade-Off Studies

3.4.1 SOFC Generator Sizing and Pressure Ratio Selection

The PSOFC/GT hybrid-cycle power system in this analysis uses the intercooled recuperated Rolls Royce Allison reheat gas turbine. Earlier analyses to size the SOFC generator for this 20 MWe system, and also to select a compressor pressure ratio, assumed the staged-cell, crossflow stack configuration, and they applied the preliminary estimate of the staged-cell voltage effect that was used in the proposal effort. That estimate took the form of an adder on the conventional cocurrent-flow cell V-I characteristic, with the adder increasing as the cell current decreased. As a result, that model tended to favor design for low-pressure ratios. Low ratios are needed to support system operation at the low current densities that are necessary for high-efficiency operation, and with the crossflow voltage adder model, the cell efficiency gain that occurred as the current was reduced more than compensated for the loss in turbine power that occurred due to the pressure ratio reduction. Thus, from the efficiency standpoint, the staged-cell voltage model tended to drive the system design point to the very lowest pressure ratios. Without the crossflow adder, the increase in cell voltage as the cell current is reduced is not strong enough to compensate for the drop in turbine power, and after some minimum cell current is reached, there is no incentive to go to lower currents and pressure ratios. The crossflow analysis indicates that the performance gains by employing the staged-cell would actually be less than those predicted originally, and it was therefore concluded that the cocurrent-flow stack configuration should be retained. As a result, the generator sizing analysis and the calculations to select a design pressure ratio needed to be re-visited.

The earlier generator sizing analyses also assumed that the power turbine and generator were mounted on a shaft separate from the gasifier shaft, and that the gasifier produced no net power. This constrained the analysis in that the expansion across the gasifier turbine was therefore set to work-balance the turbine and compressor, and the power turbine was then required to accept the remaining expansion. To better control the hybrid system, it is now believed that the turbine should be a single-shaft machine. This means that the expansion across the gasifier turbine could be an independent variable, and that it could be selected for optimum system efficiency performance.

Considering both system performance and the cost of electricity (COE), an analysis has been done to final-size the SOFC generator for the 20 MWe high-efficiency power sys-

tem, and also to evaluate the effect of pressure ratio. The SOFC generator cell stacks were of the conventional cocurrent flow configuration, and the compressor and expander wheels in the reheat gas turbine were all assumed to be installed on a single shaft. Results of the analysis are summarized in this note

3.4.1.1 Analysis Basis

- Cycle configuration – Figure 3.43.
- SOFC generator configuration – the generator is composed of pressurized generator modules. Each module is a horizontal pressure vessel housing twenty 576-cell substacks, or the equivalent of ten EDB/ELSAM 100 kW cell stacks. In the reheat power system, there are high pressure (HP) and low pressure (LP) generator sections. The number of modules in each section is to be determined by this analysis. For cost-estimating purposes, it has been assumed that the same module design is applied to both sections.
- Cell V-I characteristic – the mature-product characteristic.
- SOFC stoichs model – the stoichs profile determined by thermal-hydraulic analysis for the 1020°C peak cell temperature and 870°C combustion zone exhaust temperature (air feed tube heat transfer enhancers out) was used. The HP generator modules generally operate at higher cell currents than do the LP modules, and the air flow in the cycle is therefore set by the air flow requirement of the HP modules.
- Gas turbine – the turbine is intercooled and recuperated, with all rotating components mounted on a single shaft. The compressor pressure ratio is to be determined by this analysis. The turbine has two compressor stages, separated by the intercooler. The intercooler air exit temperature is 23°C (73°F), 8°C (14°F) above the ambient air temperature. The pressure ratio across each stage is the square root of the compressor pressure ratio, with an allowance for a 6% intercooler pressure drop.
- The gas turbine combustors and the air heaters are not fired during normal steady-state power operations.
- SOFC fuel consumption – 90%
- SOFC cell stack fuel bypass leakage – 1%
- Power lead cooling air flow – 0
- Power lead and stack internals power loss – assumed 1% of cell power. For comparison, at 500 generator amps, the EDB/ELSAM 100 kW terminal voltage is 0.5% less than the sum of the bundle voltages, based on 2/5/98 data sheet analysis. Therefore, the 1% loss assumption is reasonable, given the uncertainty in the power takeoff design.
- SOFC power conditioning system overall efficiency – 94%, mature-product value.

3.4.1.2 Discussion

System Efficiency - Effects of Pressure Ratio, Gasifier Expansion Ratio, and the Number of SOFC Modules. Results of calculations to evaluate the sensitivity of system performance to variations in compressor pressure ratio (PRAT) and gasifier turbine expansion ratio (turbine outlet pressure/inlet pressure) are shown in Figure 3.44. For these particular calculations, the power system included four HP SOFC generator modules, and four LP modules. Given a PRAT value, a curve of system efficiency can be drawn vs. power output, with each point on the curve applying to a different value of the turbine expansion ratio. In Figure 3.44, three such points, applying to expansion ratios of 0.4, 0.5, and 0.6, are identified on the PRAT = 6 curve. As the expansion ratio is increased, there is less pressure drop across the gasifier turbine, and the inlet temperature at the LP SOFC modules increases, meaning those modules will have to operate at lower cell current to maintain the required combustion zone exhaust temperature. With less expansion across the gasifier turbine, there must be more across the power turbine. This will cause lower gas temperatures at the HP SOFC module inlets, and the cell currents in those modules will therefore have to increase. Thus, with an increasing gasifier turbine expansion ratio, the tendency is for more power to be produced in the HP modules, and less in the LP modules, and there will be an optimum expansion ratio value at which the system efficiency is maximized. For PRAT = 6, that ratio is 0.50. As PRAT increases, the optimum expansion ratio decreases, because that will tend to result in less expansion across the power turbine, and higher inlet temperatures at the HP SOFC module inlets, which supports HP SOFC operation at lower, higher-efficiency cell currents.

There is also an optimum PRAT value that results in maximum system efficiency. That value weighs the positive efficiency effect of increasing pressure on cell voltage against the negative effect that is caused by the tendency to balance at higher cell currents as PRAT increases.

Figure 3.44 — Effect of pressure ratio and gasifier expansion ratio for 4/4 module configuration.

Figure 3.45 shows the same curves for the case in which the power system is equipped with an additional LP SOFC module, and higher system efficiencies result. Adding LP SOFC surface area will generally have a positive effect on efficiency because the LP cells will now be able to operate at lower, higher-efficiency currents, the gas temperature at the HP SOFC inlet will tend to increase, meaning the HP SOFC modules can be operated at higher-efficiency cell currents as well. Similarly, adding an HP SOFC module will generally result in lower system efficiencies - compare Figure 3.44 and Figure 3.46.

Figure 3.45 — Effect of pressure ratio and gasifier expansion ratio for 4/5 module configuration.

Figure 3.46 — Effect of pressure ratio and gasifier expansion ratio for 5/4 module configuration.

Performance curves for four HP modules and three LP modules are presented in Figure 3.47. The efficiencies are similar to the values reported in the Figure 3.46 for five HP and four LP modules, but the predicted power outputs are generally lower due to the overall smaller SOFC system.

Figure 3.47 — Effect of pressure ratio and gasifier expansion ratio for 4/3 module configuration.

The peak efficiency points from Figure 3.44 to Figure 3.47 for the four combinations (4/4, 4/5, 5/4, and 4/3) of HP and LP SOFC modules are graphed in Figure 3.48. The highest efficiencies occur for the 4/5 configuration. For that configuration, 20 MWe are generated with a compressor pressure ratio in the 7:1 to 8:1 range. The possibility of further unbalancing the module distribution further in favor of the LP modules (e.g., 4/6 or 3/5) to achieve higher efficiencies has not been analyzed. Large flow imbalances between the HP and LP modules can result, and their effects on module temperatures and cell voltages would need to be considered.

Figure 3.48 — Peak-performance estimates vs. module configuration and pressure ratio.

3.4.1.3 Cost of Electricity

Corresponding to the peak-efficiency curves of Figure 3.48, and based upon this study's reference fuel cost of \$3/MMBtu, relative cost-of-electricity (COE) curves for the power system are plotted in

Figure 3.49. The figure shows that with this fuel cost, the minimum COE value, at near 20 MWe, occurs with a 4/3 SOFC module combination and a high compressor pressure ratio (lowest SOFC capital cost, and maximum power from the relatively low cost gas turbine), and Figure 6 shows that the minimum COE point coincides with system operation at the minimum efficiency value. This simply confirms that with low-cost fuel, the system capital cost can be the major COE driver, and the lowest COE and the highest fuel efficiency may not occur at the same system design point.

Figure 3.49 — Relative COE estimates vs. module configuration and pressure ratio (Fuel cost = \$3.00/MMBtu).

Table 3.9 provides a COE estimate for the conventional competing technology, which for this study, is assumed to be a 20 MWe-class gas turbine combined cycle power system. Using base assumptions similar to those used in calculating the COE for the SOFC/GT hybrid power system; the estimated COE for the competing-technology system is 47.6 mills/kWh. For the high-efficiency system study, the target COE for the PSOFC/GT system is to be 10% to 20% below the competing-technology value. With the current system cost calculations, the 10% goal will possibly be reached at a relative COE in the 0.85 to 0.87 range (see

Figure 3.49).

Table 3.9 — Conventional Technology Power System COE Estimate

Information Source: Gas Turbine World 1997 Handbook, p. 24			
System Classification: Gas Turbine/Steam Turbine Combined Cycle			
Gas Turbine: General Electric LM 1600			
Steam Turbine Cycle: Two-pressure, 4.8 MWe			
Net System Power Output: 17.9 MWe			
System Efficiency: 47.9% (net AC/LHV)			
Turnkey Price: \$14.9M (\$830/kWe)			
Basis		COE Summary	
No. round-the-clock power system operators	1	Fixed O&M, \$/kW-year	
No. plants on system	5	Plant operation & control	7.3
Operator labor cost, \$/man-hour	75	Housekeeping maintenance	1.5
Housekeeping maintenance, man-hours/week/system	20	Administration (30% of operation & maintenance labor)	2.6
Housekeeping labor cost, \$/man-hour	25	Total	11.4
Power system maintenance, \$/kWh	0.007		
Power system capacity factor	0.92	Variable O&M, mills/kWh	7.0
Power system capacity, net AC kWe	17900		
Power system efficiency (net AC/LHV), %	47.9	Total O&M COE, mills/kWh	8.4
Capital charge rate, %/year	15		
Power system capital cost, \$/kW	830	Capital COE, mills/kWh	15.4
Fuel cost, \$/MMBtu (HHV)	3.00		
Fuel higher heating value, Btu/lb	21873	Fuel COE, mills/kWh	23.7
Fuel lower heating value, Btu/lb	19711	Total	47.6

3.4.1.4 Reference Power System Specifications

If we place the emphasis on system efficiency, the reference power system should have the following characteristics:

No. HP SOFC modules – 4 ($I_{\text{cell}} = 290\text{A}$, $V_{\text{cell}} = 0.681\text{V}$)

No. LP SOFC modules – 5 ($I_{\text{cell}} = 181\text{A}$, $V_{\text{cell}} = 0.726\text{V}$)

Gas turbine pressure ratio – 7

Gasifier turbine expansion ratio – 0.50

Compressor air intake rate – 18 Kg/s

Gas turbine AC power capacity – 4 MWe

System capacity – 19.0 MWe net AC

System efficiency – 67.6%

The gas turbine specs are fairly close to those already developed by RR Allison, except for the pressure ratio that is 7:1 vs. 6:1 in RR Allison design.

3.4.2 Effect of Compressor Intercooling on Power System Efficiency and Cost of Electricity

Intercooling the compressor increases the gas turbine power output, and in a recuperated cycle, it increases the power system efficiency as well. Both effects are seen in the efficiency and power curves of Figure 3.50. The curves apply to a high-efficiency PSOFC/GT hybrid-cycle power system with fixed SOFC generator design (four HP SOFC generator modules, five LP SOFC modules), but the gas turbine design varies from point to point with the changing compressor pressure ratio. The curves indicate that with intercooling, the peak-efficiency operating point occurs at a higher pressure ratio, and that intercooling adds approximately one-half of one percentage point to the peak power system efficiency, and two to three megawatts to the peak-efficiency power output. The analysis considered both the heat-rejection parasitic-power requirement that intercooling imposes on the power system, as well as the direct effect of intercooling on compressor and system power output.

The predicted effect of intercooling on COE is indicated in Figure 3.51. The estimates consider the capital cost of the intercooling equipment, and also the increased maintenance requirements that intercooling imposes on the power system. Figure 3.51 indicates that with the project reference fuel cost (\$3/MMBtu), and for a desired system net AC power output, intercooling will add to the system COE – even though it increases both power and efficiency. For the reference power system design, and consistent with the project high-efficiency focus, the intercooling feature was adopted because of its positive effect on system efficiency.

Figure 3.50 — Effect of compressor intercooling on hybrid system efficiency and power output.

Figure 3.51 — Effect of compressor intercooling on power system cost of electricity.

3.4.3 Power System Arrangement Studies

The power plant arrangement studies investigated the differences resulting from use of SOFC vessels that had process air inlet and exhaust nozzles on opposite ends versus the nozzles on the same end. Since the process air piping will be large, heavy, and expensive, arrangements were examined to minimize the quantity of process air piping.

The arrangement that minimizes the process air piping for SOFC vessels with nozzles on opposite ends is the reference arrangement that is discussed in Section 3.1.4. Although the amount of piping is greater, it has been selected for ease of installation and continuing inspection and maintenance activities. It occupies a slightly larger amount of real estate (0.6 acres vs. 0.5 acres) than the plant arrangements that support SOFC vessels with inlet and exhaust nozzles on the same end.

Two alternative plant arrangements that use SOFC vessels with process air connections on the same end were also considered. The alternative 1 system is depicted in Figure 3.52 and Figure 3.53 and alternative 2 is shown in Figure 3.54 and Figure 3.55.

Figure 3.52 — Alternative 1 System Arrangement — Isometric View.

Figure 3.53 — Alternative 1 System Arrangement — Plan View.

Figure 3.54 — Alternative 2 System Arrangement — Isometric View.

Figure 3.55 — Alternative 2 System Arrangement — Plan View.

3.4.3.1 Alternative 1 System Arrangement

This arrangement is shown on a site 51 m (167 ft) by 41 m (134 ft), or 0.5 acres. The vessels are set like the spokes of a wheel on radii from a central air distribution plenum. This plant uses a vessel design where the process air inlet and outlet nozzles are on the same end of the vessel. The vessel and SOFC stack design are relatively more difficult due to the need to ensure uniform air distribution within the stack.

Its advantage is the minimization of the quantity of large bore, internally and externally insulated piping. The cost savings for elimination of the process air piping is partially offset by the cost of the air distribution plenum vessels. This arrangement was investigated for its potential to yield lowest cost. It was hoped that the reduction in land area and process piping length would override the disadvantages. This was not the result. The advantage that was hoped to be gained by the alternative 1 configuration in its use of less land area did not materialize.

The primary disadvantages of the arrangement are inefficient routing of high voltage lines, difficult initial installation, and poor access to the SOFC vessels for maintenance.

3.4.3.2 Alternative 2 System Arrangement

This system would be installed on a site 51 m (167 ft) by 41 m (134 ft), or 0.5 acres. The plan dimensions are the same as those for the alternative 1 system. The vessels are set in rows receiving and returning process air to air distribution headers at one end of the vessels. The vessel and SOFC stack designs are relatively more difficult due to the need to ensure uniform air distribution within the stack.

The alternative 2 site has vehicle access roadways on both ends of the vessels. Routing of high voltage electric power lines from the SOFC vessels to the switchyard is efficient.

The primary disadvantages of the arrangement are difficult initial installation and poor access to the SOFC vessel halves that have the process air nozzles.

3.4.4 Desulfurization System Cost Study

This study compared an ambient temperature activated carbon desulfurization system with a heated catalytic hydrolization and adsorption system. The results of the study shows the heated system to be advantageous as discussed below.

The demonstration SOFC plants have used activated carbon beds contained within carbon steel pressure vessels to desulfurize natural gas. The significant advantage of the activated carbon is that the process works at room temperature. The main disadvantage is that the adsorption process does not produce a strong chemical bonding on the carbon and consequently the amount of sulfur adsorbed per unit amount of carbon is relatively low. As a result, the activated carbon vessels are large and the adsorbent must be replaced frequently.

A study was performed to assess the cost of using alternate sulfur adsorbents. The use of zinc oxide was found to be very attractive. The volumetric consumption of zinc oxide is about two orders of magnitude less than the consumption of activated carbon for the same duty. In order for zinc oxide to be effective, however, the natural gas temperature must be above 277°C (530°F). The optimum desulfurization temperature is at about 400°C (750°F). Operation at elevated pressures also produces a substantial improvement in sulfur removal.

If using activated carbon as an adsorbent, a 20 MW_e plant is projected to consume 173 m³ (6100 ft³) of activated carbon per year. The material replacement cost is \$578,000 at \$2.50 per pound. Additionally, the disposal cost is \$145,000 at \$175 per 55-gallon drum. The annual cost for activated carbon is \$723,000.

This is compared to the annual consumption of only 0.85 m³ (30 ft³) of zinc oxide which operates at 12.4 barg (180 psig) and 400°C (750°F). The material replacement cost is \$8100 at \$270 per cubic foot. The disposal cost is estimated as \$715 per year for four 55-gallon drums. The desulfurization process does not consume the cobalt molybdenum catalyst. It will slowly become poisoned over a long period of time. It will probably require replacement after some time greater than ten years. Assuming the catalyst is replaced after ten years, the annual replacement cost is \$920 for two 0.57 m³ (20 ft³) Co-Mo catalyst beds. The spent cobalt-molybdenum can be sold to a reclaiming company for about \$0.50 per pound. This gain is offset by the shipping cost with the result being no net gain or loss to dispose of the Co-Mo. The annual cost for desulfurization

adsorbent and catalyst cost is \$9735. Combined with the cost of consuming about 180 kW for electrolysis of water to produce hydrogen, the total annual desulfurization cost is about \$90,000. This is considerably less than the cost to use activated carbon.

The first cost of the large vessels for activated carbon would be higher than the combined cost for the much smaller vessels for heated zinc oxide, the H₂ generator, the natural gas recuperator and electric gas heater, even with the added complexity of supporting separate adsorbent and catalyst beds. The study concluded that the preferred fuel desulfurization technique employs the cobalt molybdenum catalyst and zinc oxide adsorbent.

3.4.5 Cover Gas System Cost Study

This study compared three alternatives for providing a protective mixture of nitrogen cover gas to the SOFC stacks. The alternatives were buying bottled nitrogen gas, buying liquefied nitrogen and generating nitrogen on site and storing it in pressurized bottles. The most cost-effective alternative is to generate nitrogen on site and store it for use as discussed below.

The demonstration SOFC plants have relied upon supplies of NHMIX cover gas stored in pressurized bottles. As the numbers of cells increased, the quantities of NHMIX increased proportionally. It became a source of concern that the large number of SOFCs in a 20 MW_e plant would require a large allocation of real estate for storing NHMIX in addition to the continuing cost of replenishing the cover gas.

Limited testing of SOFCs has shown that under some conditions it is not necessary to provide a reducing atmosphere on the fuel side of the SOFCs. It is expected that by the time that a 20 MWe plant is supplied, the process conditions and possibly even the SOFC may modify chemistry such that NHMIX is not needed. This study has proceeded upon this assumption.

In the current commodity gas market, 28.3 m³ (1000 ft³) of bottled nitrogen costs \$12.03.

The required quantity of nitrogen which must be stored at the plant are based on the following assumptions:

- The initial stack temperature is 566°C (1050°F). The stack-cooling rate after shutdown is -3°C/minute. Cover gas is needed until the stack temperature is reduced to 260°C (500°F). Thus, the minimum supply of cover gas must be sufficient for 183 minutes.
- The amount of cover gas needed for starting up the plant is the same, as that needed for cooling the plant after shut down.
- An additional 33% is added to the gas quantities for uncertainty and conservatism.
- 242.5 cc/minute of nitrogen is needed for each SOFC. To support the 73,728 SOFCs in the plant, this results in 17.9 m³/min (632 scfm) of nitrogen.

Each plant start up or shut down consumes 4376 m³ (154,525 scf) of nitrogen. The gas consumption cost is \$1859 for nitrogen for a single shut down. For one plant start and stop per year, the direct annual gas usage costs are \$3718. In addition to the cost of the gas, the demurrage for a large nitrogen tube trailer is \$950/month. The delivery fee for a trailer is built into the demurrage. Total costs for use of purchased bottled nitrogen gas is \$15,108 per year.

The costs for using purchased bottled gas are considered constant because the compressed gas industry is mature and gas is supplied as a commodity. Over a twenty year plant life, the cover gas costs based upon using bottled gases sum to \$302,000.

The amount of nitrogen merits consideration for buying and storing in the liquid form. The dewar for storing 4376 m³ (11234 pounds or 154,525 scf) of liquid nitrogen must be at least 70.8 m³ (250 ft³). Liquid nitrogen costs \$3.40 per 28.3 m³ (1000 scf). The delivery to fill the dewar will include an excess of 5% (225 lbm) to allow for vaporization. This cost is \$552 for 4594 m³ (162251 scf) of liquid nitrogen plus a \$125 delivery fee. Every month the dewar must be topped off with an additional 5% to maintain the minimum required inventory. This is an additional \$28 for the nitrogen plus a \$125 delivery fee. The dewar and instrumentation rental is a monthly cost of \$750. Total costs for two fills of purchased liquid nitrogen, monthly replenishment and equipment rental is \$11,388 per year. Over a twenty year plant life, the cover gas costs based upon using bottled hydrogen and liquid nitrogen sum to \$228,000.

As a basis of comparison, the installation of nitrogen generating equipment was assessed and resulted in the preferred (lowest cost) option. A prepackaged nitrogen generator system with accumulator, compressor, vent, pressure relief and pressure control is available for about \$20,500. The high-pressure gas storage bottles cost about \$60,000. The unit has been sized to fill the high-pressure storage accumulator in seven days. While operating, the unit will consume electrical power and compressed air. Once the accumulator is filled, the unit will normally be idle. It will start upon low-pressure demand signal from the accumulator and operate until the accumulator is re-filled. The functional specifications for the unit are shown in Table 3.10.

Table 3.10 — Nitrogen Generator/Compressor Characteristics

Nitrogen production rate	28.3 m ³ /hr (1000 scf/hr)
Nitrogen pressure before compression	6.9 barg (100 psig)
Nitrogen pressure after compression	165 barg (2400 psig)
Nitrogen purity	>98%
Process used	Selective Membrane
Package size	1.52 m x 1.83 m x 1.93 m (60" x 72" x 76")
Nitrogen separator storage tank	1.9 m ³ (500 gallons), ASME tank
Nitrogen compressor accumulator	twelve – 2.19 m ³ (77.5 ft ³) ASME/CGA bottles
Electrical power requirements	110 vac/60 Hz/1 phase for separator 440 vac/60 Hz/3 phase, 5 kW for compressor
Compressed air supply required	1.9 m ³ /min (66 scfm) @ 175 psig

The equipment costs for the nitrogen generating and storage system sum to about \$80,500. The initial expenditure of this amount is justified compared to the continuing annual cost of purchasing bottled or liquid nitrogen.

Table 3.11 summarizes the costs for the various options investigated for supply of a nitrogen system.

Table 3.11 — Costs for Options Investigated for Supply of Nitrogen System

	Bottled Nitrogen	Liquid Nitrogen	Nitrogen Generating System
	\$3,718 Annual Gas Cost (Delivery Included)	\$2,388 Annual Gas Cost & Delivery Fees	\$20,500 Initial N ₂ Equipment Cost (one time cost)
	\$11,400 Annual Bottle Trailer Demurrage	\$9000 Storage Dewar & Instrumentation Annual Rental	\$60,000 Bottle Trailer (one time cost)
Annual Cost	\$15,108	\$11,388	\$4025 pro-rated
20 Year Cost	\$302,000.	\$228,000	\$80,500

3.4.6 Hydrogen Gas Generation Cost Study

The use of a hydrogen generating system has already been justified due to the value of adding hydrogen to the natural gas to support catalyst assisted desulfurization using heated zinc oxide.

There are several suppliers of commercially available hydrogen generating equipment both in the USA and Canada. A prepackaged hydrogen generator/compressor system with accumulator and pressure control is currently available for \$245,000, but can be purchased in quantity (>100 unit per year) for \$61,250 to \$122,500 (Supplier's estimates). The functional specifications for the unit are shown in Table 3.12.

Table 3.12 — Hydrogen Generator Characteristics

Hydrogen production rate	0.57 m ³ /min (20 scfm)
Hydrogen pressure	12.4 barg (180 psig)
Process used	electrolysis
Package size	1.82 m x 5.49 m x 3.05 m high (6' x 18' x 10') high
Package weight	7484 kg (16,500 pounds)
Hydrogen accumulator size	119 sm ³ (4200 scf) @ 12.4 barg (180 psig) or 8.98 m ³ (317 ft ³)
Power	180 kW
Water consumption	35 liters/hr (9.24 gph)

3.4.7 Process Piping Cost Study

The first task in the process piping cost study was to determine the required flow area. When the airflow was assumed to be about 61 m/sec (200 fps or ~10% of sonic speed) an internal protective liner was necessary to prevent erosion of the insulation.

At air temperatures in the range of 815°C to 870°C (1500 to 1600°F) metallic liners are unsuitable except for short lengths due to their high thermal expansion. Ceramic liners were found which have excellent erosion resistance but are unsuitable for other reasons. The ceramic liner materials are necessarily dense with high thermal conductivity and are fabricated by casting. The casting process requires a thickness of one inch. This dense ceramic adds 5 cm (two inches) to the piping inner diameter, provides no thermal insulation benefit, adds weight and is a very costly process.

When the inner flow area was selected to keep the gas speed to less than 30 m/sec (100 fps), conventional vacuum-formed alumina-silica insulation can be used. This does require that the piping be larger, but the complexity of fabricating insulated piping results in lower total cost.

The estimated costs of internally and externally insulated piping for the 20 MW_e plant are shown in Table 3.13.

Table 3.13 — High Temperature Piping⁽¹⁾ Costs

Application	Open Flow Area ID, inches	Pipe Size (0.375" wall)	Cost per linear foot⁽²⁾
High Pressure Header (88 psia)	22	38	\$1104
High Pressure Branch (88 psia)	14	30	\$838
Low Pressure Header (37 psia)	41	58	\$1813
Low Pressure Branch (37 psia)	28	44	\$1301

Notes:

1. 3" exterior insulation
2. Cost for piping and insulation only, no fittings, flanges or bolting materials

Several plant layouts were considered to minimize the amount of process piping. Figure 3.9, Figure 3.52, and Figure 3.54 show some preliminary layouts of a gas turbine/compressor skid with four high-pressure SOFC vessels and five low-pressure SOFC vessels. Figure 3.9 shows an arrangement that uses SOFC vessels that have nozzles on both ends. This arrangement results in the largest footage of insulated piping. It has an advantage of providing the most pneumatically balanced flow resistances. The other figures show plant arrangements that use a vessel with the air inlet and outlet connections on the same head. Figure 3.52 is a layout with a star shaped arrangement of vessels that minimizes the plan area of the plant. This minimal plan area arrangement has the disadvantages of requiring expensive air distribution vessels and

limiting access to the vessel halves that are closest to the air distributor vessels. These disadvantages are also shared by the butterfly plant arrangement (Figure 3.54 and Figure 3.55).

4. CONCLUSIONS

Study conclusions can be summarized as follows:

- A PSOFC/GT system concept of near 20 MW capacity has been devised that is conservatively capable of 67% efficiency, a value ten points greater than that achievable with the best available large-plant conventional power generation technology, and twenty points above the efficiency achieved by a conventional 20 MW-class gas turbine combined cycle power system.
- The specific power system concept developed during this study, integrating HP and LP SOFC generators with an intercooled, SOFC-reheated gas turbine, achieves an estimated power output of 19 MWe at an efficiency of 67.3% (net AC/LHV). Improvements in the performance of major system components, particularly in the SOFC PCS, for which there was no study design task, and employment of an ambient-temperature passive sorbent technology for fuel desulfurization would cause the system efficiency estimate to approach very closely the 70% efficiency target; adding a steam turbine system will result in efficiency >70%.
- The staged-cell SOFC stack design does not offer the large SOFC efficiency gain (over the standard cocurrent axial flow stack design) that was projected originally. Cell cooling in the fuel-entry cell rows reduces the average cell voltage, while there is little increase in average fuel utilization at the last cell row at meaningful current densities because of the hazard of anode oxidation.
- For the reference fuel cost of \$3.00/MMBtu, the estimated COE for the HEFPP system is 3% higher than the COE estimate for a conventional 20 MW-class gas turbine/steam turbine power system. Leveraged by its significantly higher efficiency, the HEFPP system would have a COE advantage in a higher fuel cost environment. For example, with \$6 fuel, the HEFPP COE would be 7% less than the conventional-system COE.

Recommendations:

- Desulfurization technologies not requiring a source of hydrogen, and capable of operation at ambient-temperature levels, should be developed. For the reference power system, this would increase the system efficiency by approximately 0.5 of a percentage point.
- For high efficiency SOFC/GT power systems, the SOFC power conditioning efficiency affects strongly the overall system efficiency. For example, a gain in PCS efficiency of one percentage translates to a system efficiency gain of nearly 0.6 of one percentage point. Power conditioning topologies with greater than 95% efficiency should be developed.
- Small, efficient, highly-reliable, recuperated gas turbines with turbine inlet temperatures commensurate with SOFC exhaust gas exit temperatures (870°C) are needed for deployment in hybrid cycle SOFC/gas turbine power systems. Effort

should be focused on the specification of such gas turbines (circa 4 MWe capacities for use in 20 MWe hybrid cycle power systems) and their development should be undertaken.

- A PSOFC/GT power system of 70% efficiency potential should be developed and demonstrated at the smallest capacity class practical for proof-of-concept.
- SOFC development should be pursued to improve fuel cell power density and efficiency. As with the SOFC power conditioning system, improvements in cell efficiency are effective in increasing the efficiency of high-efficiency SOFC/GT power systems.
- In the conventional SOFC generator design, reformed fuel enters the cell stack at the bottom and flows upwards in the system of communicating, parallel-flow channels defined by the cell exterior surfaces. The gas density increases as oxidation products are produced, and this could foster parallel channel instabilities, particularly at elevated operating pressures where buoyancy effects are more significant. Development work should be undertaken to confirm the operational feasibility of the conventional SOFC generator configuration at elevated pressures beyond three atmospheres.

APPENDIX

The Effect of Staging on Efficiency of Isothermal SOFC Stacks

The purpose of this section is to describe the effect of staging of the fuel stream on the efficiency of fuel cells. Staging of the fuel flow path is described as having the fuel pass a number of separate or segmented electrical circuits on its way through the fuel cell array. The current density is not uniform within any cell or stage that has a large difference in fuel concentration between the inlet and outlet locations. The terminal voltage across a stage is uniform, but the internal Nernst potential varies due to the change in reactant concentrations. With staging, one can either divide the cell structure into many regions or the fuel supply for many cells can pass sequentially through or past the cells. With staging the current density distribution becomes more uniform because the Nernst potential varies by a smaller amount within each stage. More uniform current density lowers the Joule losses and polarization losses. The voltage gain in the fuel rich stages is greater than the loss in the fuel poor stages.

It will be shown that the improvement in efficiency is small at economically viable operating current density conditions, namely at current densities near the maximum power point. The maximum power point occurs at the current density that maximizes the power output per unit area of the cell.

ISOTHERMAL ANALYSIS

The analysis is simplified by assuming that leakage of O_2 is negligible, that the cells are isothermal, and that the cathode gas is supplied at high stoich conditions. These conditions are favorable to the benefits of staging. For example, non-isothermal conditions within a stage create restraints on the flow and geometrical arrangements necessary to achieve uniform fuel consumption within the stages.

Consider that the fuel is being consumed by an electrochemical conversion as it flows along a path through a number of stages. The properties used in the analysis are applicable to solid oxide fuel cells but the generalized analysis is independent of the specific cell geometry. If the total in-stack fuel consumption along the path is F_c , the consumption in each of the n stages will be F_c/n and the consumption at the end of stage i is $z_i = (i/n) F_c$. The local fuel consumption within a stage is denoted by z . The change in consumption, dz , that occurs in elemental area, dA , is related to the local current density, j , by

$$dz = F_c (j \, dA/l_o) = (F_c/j_{av}) j \, dA/A_o = (F_c/j_{av}) j \, dx$$

or

$$dx/dz = j_{av}/(j \, F_c) \quad (A.1)$$

where dA/A has been replaced by dx .

If one also assumes that the electrical potential produced by CO is equal to that produced by H_2 , the fuel mole fraction (the sum of H_2 and CO mole fractions) is uniquely determined by the fuel consumption and the inlet composition. Thus with uniform cathode concentration or cathode composition also a function of consumption, the Nernst potential can be stated to be a function of the fuel consumption, $E_n = E_n(z)$.

The local current density at a location where the fuel consumption is z is given by

$$j = (E_n(z) - v_i)/R_c \quad (A.2)$$

Substituting this into equation (1) gives

$$dx/dz = (j_{av} \, R_c/F_c)/(E_n(z) - v_i) \quad (A.3)$$

After integration over the stage, the equation defining the cell voltage in each stage becomes

$$\int_{z_{i-1}}^{z_i} \frac{R_c \, dz}{(E_n(z) - v_i)} = \frac{F_c \Delta x}{j_{av}} \quad (A.4)$$

For n equal area stages, $\Delta x = 1/n$ for each stage. The solution for the v_i of each stage is found by iteration using numerical integration of equation (4). If the resulting value of Δx is too small/large for an assumed v_i , the value of v_i in the next iteration must be increased/decreased subject to the limit that $0 < v_i \leq E_n(z_i)$. The solutions presented here were obtained by numerical integration of equation (4). The cell resistance, R_c , was calculated as the sum of the internal resistance, the diffusion polarization resistance's for each electrode, and the activation polarization resistance. The polarization resistance's were obtained by dividing the anode or cathode polarization by the local current density. After one obtains the v_i , it is easy to plot the current density versus the non-dimension area using the values from the final iteration. The value of dimensionless area, x , corresponding to a given value of z in stage i is given by

$$x = (i-1)\Delta x + \frac{j_{av}}{F_c} \int_{z_{i-1}}^z \frac{R_c dz}{(E_n(z) - v_i)} \quad (\text{A.5})$$

The current density at this x location is given by equation (2).

RELATIONS FOR MOLE FRACTIONS, DIFFUSION POLARIZATION, AND NERNST POTENTIAL

The mole fraction of fuel in the anode gas stream at fuel consumption z is

$$y_F(z) = y_{Fo}(1 - z)$$

The mole fraction of oxygen in the cathode gas stream at fuel consumption z is

$$y_o(z) = \frac{(1 - z/s)}{(1/y_{oo} - z/s)}$$

where y_{oo} is the mole fraction of oxygen at inlet to the stack. The partial pressure of oxygen in the anode gas stream is found using an effective equilibrium constant for the methane derived fuel mixture,

$$p_{oA}(z) = \left\{ \frac{(1 - y_F(z) - y_N)K}{y_F(z)} \right\}^2$$

The Nernst potential based on the concentration in the gas streams is

$$E_n(z) = \frac{RT}{4F} \ln\left(\frac{py_o(z)}{p_{oA}(z)}\right)$$

The anode diffusion polarization is

$$\eta_A = \frac{RT}{4F} \ln\left(\frac{p_{oA}(z)}{p_{oAE}(z)}\right)$$

where

$$p_{oAE}(z) = K^2 \left(\frac{1 - y_N - y_{FAE}(z)}{y_{FAE}(z)} \right)^2$$

$$y_{FAE}(z) = y_F(z) - j(z)/S_3$$

$$p_{oA}(z) = K^2 \left(\frac{(1 - y_N - y_F(z))}{y_F(z)} \right)^2$$

The cathode diffusion polarization is

$$\eta_c = -\frac{RT}{4F} \ln(1 - j(z) / S_1 y_o(z))$$

ISOTHERMAL ANALYSIS RESULTS

The example is based on representative properties of a solid oxide fuel cell at 1000°C. The fuel selected was the DOE fuel: 96% methane, 2% nitrogen and 2% CO₂. The inlet fuel composition to the first stage is based on internal reformation with a recirculation ratio that gives an oxygen to carbon ratio, OCR, equal to 2.1. Results are obtained at two system fuel consumption levels, 85% and 95%. The corresponding in-stack fuel consumption values, F_c , are 0.6839 and 0.8946. The combined fuel mole fraction of the inlet fuel is $y_{F0} = 0.6326$ for both fuel consumption levels.

The reference current density was selected to be the value that gives the maximum power density for a single stage cell configuration. A plot of terminal voltage and power density as a function of current density is shown in Figure A.1 for $F_{cs} = 0.85$. The values of current density at the maximum power point are $j_{maxP} = 0.522$ and 0.498 A/cm², for the 0.85 and 0.95 fuel consumption values, respectively.

Figure A.2 compares the current density and voltage distributions for single stage and four stage configurations at maximum power density for 85% system fuel consumption. For the four stage configuration, the total area is divided into four equal parts. The upper two curves, the Nernst potentials, are almost indistinguishable. The middle set of curves show the uniform voltage of the single stage and the four individual voltages of the four stage configuration. The bottom set of curves compare the current density distributions. Note that the maximum to minimum change in current density is reduced for the four-stage case. Since the average stage voltage is higher than the single stage voltage at the same current density and fuel consumption, the ratio of voltage in the same as the ratio of average power densities of the staged versus single stage cells. The average power density for this current density is increased by 0.58% above the value for a single stage cell for the limiting case of 16 stages.

Figures A.3 and A.4 show the same comparison at reduced current densities, $j_{\max P}/5$ and $j_{\max P}/20$, respectively. The staged power density only increases significantly at the lower current densities. Note that current density scales were changed in these figures to separate the two sets of curves.

Figures A.5, A.6, and A.7 provide similar comparison for 95% system fuel consumption. Table A.1 compares improvements in power density at 16 stages to that at a single stage. This table illustrates that the power density can be increased by significant amounts at a specified current density only for power densities that are well below the maximum power point. Thus the benefit of increased efficiency due to staging is only available at low power densities. Since staging is only beneficial at low power per cell, many more cells are required to compensate for the low power operation. Figure A.8 shows the relative number of cells required for a fixed plant output as a function of the current density.

The analysis was also applied to a generic planar cell configuration with a significantly lower total effective resistance. The total resistance including concentration polarization losses was held constant at 0.20 ohm-cm^2 . The maximum power density for this single stage planar cell occurs at a current density of 2.10 A/cm^2 for $\text{FC} = 85\%$ and at 2.05 A/cm^2 for $\text{FC} = 95\%$.

The increase in power output for Siemens Westinghouse cylindrical cells is compared to that of planar cells in Figure A.9. The curves show the increase in power output as a result of staging. The increase is shown as a percentage of the power output of the corresponding single stage cell. The current density has been made dimensionless on the respective current density at the maximum power point for the single stage cell condition. It is seen that the use of normalized current density results in an excellent correlation of the improvement in efficiency due to staging. The slight difference in the curves for the two geometries is due to the change in diffusion polarization resistance with current density in the cylindrical cell model. Otherwise the gains appear to be a function of normalized current and utilization.

Although the curve shows significant increase in output for $\text{FC} = 0.95$, the partial pressure of O_2 at the anode/electrolyte interface exceeds the limit which is two orders of magnitude less than the equilibrium partial pressure of O_2 for the Ni/NiO/O_2 reaction. Oxidation of the anode is considered likely under these conditions. The diffusion con-

ductance of Siemens Westinghouse Power Corporation cells was used to evaluate the partial pressure of fuel at the electrolyte interface. For recirculated fuel, the fuel utilization limit is $FU = 91\%$ when the partial pressure of O_2 is maintained below the limit given above.

The effect of once through fuel flow on the increase in output due to staging was also considered. With the same fuel supply, external reformation at oxygen to carbon ratio of 2.1 gives a fuel inlet mole fraction of 0.780. The in-stack fuel utilization value equals the system fuel utilization. Results for the planar cell with constant resistance are shown in Figure A.10 for 85% and 95% system fuel utilization. As expected, the gain due to staging increases significantly if one can push the operation to higher fuel utilization for once through fuel flow (external fuel reformation). Limiting the partial pressure of O_2 to two orders of magnitude below the equilibrium value restricts fuel utilization to $FU = 89\%$.

SUMMARY OF ISOTHERMAL ANALYSIS

The results demonstrate that the effect of fuel staging on fuel cell power output is very small for current densities near the maximum power point. This is shown to be valid also for other geometries that have considerably lower resistance than the cylindrical cell. With fuel recirculation, the output at the maximum power point increases by 0.58 and 1.15% at 85 and 95% fuel utilization, respectively. The improvement due to fuel staging increases as current density is reduced or as fuel utilization is increased. Oxidation of the fuel electrode is likely with recirculated fuel for utilization above 91%.

The improvement due to staging is greater for systems with once through fuel flow as compared to those with recirculated fuel. However, oxidation of the anode is likely for once through flow when fuel utilization exceeds 89%. Although fuel staging could increase output at low current densities by more than 10% at 95% utilization, the low power density would result in a large increase in system cost.

Nomenclature for Appendix A

A_o	Total cell area, [cm ²]
E_n	Nernst potential, [V]
F_c	In-stack fuel consumption, [-]
F_{cs}	Overall system fuel consumption, [-]
I_o	Total cell current generated by n stages = $j_{av} A_o$.
i	Stage index, [-]
j	Local current density, [A/cm ²]
j_{av}	Average current density for the stage, [A/cm ²]
j_{maxP}	Current density that gives maximum average power density for a single stage Cell, [A/cm ²]
K	Effective equilibrium constant for the methane derived fuel mixture, [-]
n	Number of stages, [-]
p	Pressure of the fuel cell gases, [Atm]
Pd	Average power density, [W/cm ²]
$p_{OA}(z)$	Local partial pressure of oxygen in the anode gas stream, [Atm]
$p_{OAE}(z)$	Local partial pressure of oxygen at the anode electrolyte, [Atm]
R_c	Cell resistance index, [ohm-cm ²]
S_1	Diffusion conductance of the air electrode at one Atm, [A/cm ² -Atm]
S_3	Diffusion conductance of the fuel electrode at one Atm, [Atm/cm ² -Atm]
v_i	Terminal voltage of stage i, [V]
x	Non-dimensional cell area, = A/A_o , [-]
Δx	Non-dimensional area per stage, [-]
$y_F(z)$	Mole fraction of fuel in the anode gas at fuel consumption z, [-]
$y_{FAE}(z)$	Mole fraction of fuel at the anode electrolyte at fuel consumption z, [-]
y_{Fo}	Fuel mole fraction in anode gas at inlet to the stack, [-]
y_N	Mole fraction of nitrogen in the fuel at inlet to the stack, [-]
$y_O(z)$	Mole fraction of oxygen in the cathode gas stream at fuel consumption z, [-]
y_{OO}	Mole fraction of oxygen in the cathode gas at inlet to the stack, [-]
z	Local in-stack fuel consumption, [-]
z_i	In-stack fuel consumption at exit of stage i, [-]
η_A	Anode diffusion polarization, [V]
η_C	Cathode diffusion polarization, [V]

Figure A.1 — Power Density and Terminal Voltage for a Single Stage Cell.

Comparison of Single and 4 Staged Fuel @ max Power

Figure A.2 — Emf and Current Density Distributions in a 4 Staged Fuel Cell.

Comparison of Single and 4 Staged Fuel @ $j=j_{\max P}/5$

Figure A.3 — Emf and Current Density for Single and 4 Staged Cell.

Figure A.4 — Emf and Current Density for Single and 4 Staged Cell.

Comparison of Single and 4 Staged Fuel @ max Power

Figure A.5 — Emf and Current Density for Single and 4 Staged Cell.

Comparison of Single and 4 Staged Fuel @ $j=j_{\max P}/5$

Figure A.6 — Emf and Current Density for Single and 4 Staged Cell.

Comparison of Single and 4 Staged Fuel @ $j=j_{\max P}/20$

Figure A.7 — Emf and Current Density for Single and 4 Staged Cell.

Table A.1
Effect of Staging on Average Voltage or Power Density

j_{av} , [A/cm ²]	F_{cs} , [-]	P_d for 1 stage, [W/cm ²]	$\Delta P_d/P_d$ for 16 stages, [%]
0.5220	0.85	0.2268	0.58
0.1044	0.85	0.0785	1.53
0.0261	0.85	0.0205	4.58
0.4980	0.95	0.2099	1.15
0.0996	0.95	0.0707	5.06
0.0249	0.95	0.01786	11.87

Figure A.8 — Relative Number of Cells Required for Fixed Power Output as a Function of Current Density.

Comparison of Increase in Power Output Due to Staging.

Figure A.9 — Increase in the Power Output Due to Staging for Planar and Cylindrical Cells vs. Dimensionless Current Density.

Once Through Flow Compared with Recirculation Fuel Flow, Staged Planar Cells

Figure A.10 — Comparison of Effect of Once Through and Recirculation Fuel Flow on the Benefits of Cell Staging.

A High Efficiency PSOFC/ATS-Gas Turbine Power System

Final Report

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ABSTRACT

A study is described in which the conceptual design of a hybrid power system integrating a pressurized Siemens Westinghouse solid oxide fuel cell generator and the Mercury™ 50 gas turbine was developed. The Mercury™ 50 was designed by Solar Turbines as part of the U.S. Department of Energy Advanced Turbine Systems program. The focus of the study was to develop the hybrid power system concept that principally would exhibit an attractively-low cost of electricity (COE). The inherently-high efficiency of the hybrid cycle contributes directly to achieving this objective, and by employing the efficient, power-intensive Mercury™ 50, with its relatively-low installed cost, the higher-cost SOFC generator can be optimally sized such that the minimum-COE objective is achieved. The system cycle is described, major system components are specified, the system installed cost and COE are estimated, and the physical arrangement of the major system components is discussed. Estimates of system power output, efficiency, and emissions at the system design point are also presented. In addition, two bottoming cycle options are described, and estimates of their effects on overall-system performance, cost, and COE are provided.

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Appendix A: The Effect of Staging on Efficiency of Isothermal SOFC Stacks

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1. EXECUTIVE SUMMARY

1.1 Power System Design Description

The objective of this project was the development of a 20 MWe-class, highly-efficient (net AC/LHV > 60%), hybrid cycle (fuel cell plus gas turbine) power system concept that would take advantage of gas turbine technology developed under the U.S. Department of Energy Advanced Turbine Systems (ATS) program and achieve an attractively-low cost of electricity (COE) relative to the COE for conventional-technology gas turbine combined cycle power systems. The power system devised is based on a hybrid cycle that integrates a pressurized Siemens Westinghouse tubular solid oxide fuel cell (PSOFC) generator with a fired Solar Turbines Mercury™ 50 gas turbine. Due to its application of advanced turbine technology, and to recuperation, the Mercury™ 50 achieves a Brayton cycle efficiency of 41% to 42% (GT AC/LHV). The nominal GT AC power output is 4.5 MWe [15°C (59°F) ambient-air temperature]. The premise of the study was that applying the efficient, power-intensive Mercury™ 50, fired to its design turbine inlet temperature (TIT), in the hybrid cycle power system would minimize the required number of relatively high cost SOFC stacks needed to achieve a total system power output of roughly 20 MWe, resulting thereby in a system that would operate with a very low COE.

Operating at atmospheric pressure, the efficiency horizon for an SOFC generator is 45% to 50% (net AC/LHV), while a Mercury™ 50-class gas turbine, operating at its design point, will convert to electric power over 40% of the heat supplied to the gas turbine cycle. When a PSOFC generator and the gas turbine are integrated in a power system, a system efficiency near 60% can be achieved. This is due to the processing to power of SOFC exhaust heat by the gas turbine, the attendant reduction in the rate at which fuel is needed by the gas turbine combustor, and to the enhanced operation of the SOFC generator at elevated pressure.

The low-COE PSOFC/ATS-GT system concept developed by Siemens Westinghouse is based on the cycle depicted in Figure 1.1. The gas turbine/generator and recuperator are components in the Mercury™ 50 package, and the SOFC generator consists of an array of fuel cell stacks housed in horizontal cylindrical pressure vessels. The PSOFC generator produces DC power that is converted by the power conditioning system (PCS) to AC form suitable for export to the utility AC grid. The PCS is instrumental in

the achievement of high hybrid cycle efficiencies, and continuing work is needed to develop affordable power conditioning systems with DC-to-AC efficiencies in excess of 95%.

Figure 1.1 — Simplified PSOFC/ATS-GT hybrid power system cycle.

Figure 1.1 provides detail on the fuel desulfurization system. The power system fuel, pipeline natural gas, can contain sulfur-bearing compounds, occurring naturally, or added for leak-detection purposes, that must be reduced in concentration to preclude a sulfur reaction with fuel cell nickel components. In the power system concept, a low-maintenance fuel desulfurizer is employed in which sulfur is reacted to hydrogen sulfide in a cobalt-molybdenum catalyst bed, and the resulting H₂S is adsorbed by heated zinc oxide.

An isometric view of a 20 MWe-class PSOFC/ATS-GT power system is presented in Figure 1.2. It is composed of two power blocks, each consisting of a single Mer-

cury™ 50 gas turbine and an SOFC generator composed of four SOFC modules that are arranged in flow parallel between process air and exhaust manifolds. Each module is a horizontal cylindrical pressure vessel, flanged at the middle, that houses an assembly of 11,520 SOFCs. The overall length of a module is 11 m (36 ft), and its diameter is 3.5 m (11.5 ft). An individual fuel cell is tubular, having an active length of 1500 mm and a diameter of 22 mm. The capacity of a single power block is 12.3 MWe. Thus, two blocks are required for a 20 MWe-class power system, and additional blocks could be added to achieve higher power system capacities. Note that the power blocks are independent, and provided the common balance-of-plant equipment is available, one block could be operated while the other block is down for maintenance. For the two-block power system, the dimensions of the rectangular plot plan depicted in Figure 1.2 are 38m (125 ft) x 59m (194 ft), corresponding to a site footprint of 0.56 acre.

Figure 1.2 — 20 MWe PSOFC/ATS-GT power system site arrangement.

In the PSOFC generator design, the flows of air and reformed-fuel on the cathode and anode of each fuel cell occur cocurrently and parallel to the cell axis. This is the conventional configuration that has been used by Siemens Westinghouse in all demonstration SOFC generators designed and operated to date. It was selected for application in the PSOFC/ATS-GT system concept after an extensive analytical evaluation of the staged-cell SOFC generator concept. The staged-cell design, which retained the conventional air delivery design on the cell cathode side, but employs crossflow on the fuel (anode) side, was originally believed to offer the possibility of operating the SOFC generator at high fuel utilizations, thereby contributing to the achievement of higher generator and system efficiencies. Based upon analysis, the potential benefits of staging proved elusive for two major reasons. First, since the first cell stage is fed relatively cold fuel, the first few cell stages operate at substantially less than optimum temperature for yttria-stabilized zirconia electrolyte cells. Second, the last cell stage can not be operated at fuel utilization greater than that for the cocurrent stack because the approximately parabolic axial temperature distribution and concomitant non-uniform current density distribution place the cell hot spot at hazard for anode oxidation at roughly the same average fuel utilization achievable in a non-staged cell stack.

1.2 Power System Performance Estimates

The power system is intended for baseloaded design-point operation. At that point, the system, consisting of the two power blocks, produces 24.7 MWe net AC power with an efficiency of 59.9% (net AC/LHV). Approximately 16 MWe are derived from the SOFC generator modules, and the remainder from the two gas turbines. Thus, the ratio of SOFC-derived power to ATS GT power is approximately 2:1. The estimated rates of CO₂ and NO_x emission at the system design point are 340 kg/MWh and 0.04 kg/MWh (5 ppmv), respectively; the exhaust flow rate and temperature are 34 kg/s and 360°C.

1.3 System Cost and Economics

The system installed-cost estimate is \$1170/kWe. It includes costs related to site preparation, equipment procurement, shipping, and installation, as well as allowances for G&A, sales and marketing, and profit. Mature technologies and products were as-

sumed. The distribution of the power system installed cost is indicated in Table 1.1, where values are referenced to the total power system capacity.

Table 1.1 — PSOFC/ATS-GT System Installed Cost Summary

	Cost, \$/kWe
SOFC Generator Equipment	334
SOFC Power Conditioning Equipment	147
Gas Turbine Generator Equipment	200
Balance of Plant Equipment	178
Subtotal	859
Site Preparation	17
Project Management and Engineering	37
Overhead and Profit Allowance	241
Subtotal	295
Total Plant Cost	1154
Spare Parts, Startup, and Land Allowance	17
Total Capital Requirement	1171

COE estimates have been developed for the PSOFC/ATS-GT system, operating at its design point, and also for a competing technology, which was assumed to be 20 MWe-class gas turbine combined cycle power system. Table 1.2 summarizes input used in the COE analysis, and also the results.

Table 1.2 — Economic Analysis Summary of Results

	PSOFC/ATS-GT Power System	Conventional-Technology Power System (Gas Turbine/Steam Turbine Combined Cycle)*
No. of round-the-clock operators	1	1
No. systems in operation	5	5
Labor cost components	System operation & house-keeping maintenance	System operation & housekeep- ing maintenance
SOFC replacement	\$0.0014/system kWh	
Gas turbine/steam turbine sys- tem maintenance	\$0.004/GT kWh	\$0.007/system kWh
Power system capacity factor	0.92	0.92
Capital charge rate	15%	15%
Fuel cost	\$3/MMBtu	\$3/MMBtu
Power system capital cost	\$28.9M (\$1170/kWe)	\$14.9M (\$830/kWe)
Power Output	24.7 MWe	17.9 MWe
Emissions estimates		
CO ₂	340 kg/MWh	420 kg/MWh
NO _x	0.04 kg/MWh (5 ppmv)	0.380 kg/MWh (25 ppmv)
SO _x	0	2.5 g/MWh (4 ppmv sulfur in fuel)
Power system efficiency (net AC/LHV), %	59.9	47.9
Cost of electricity, mills/kWh	44.8	47.6

* Information Source: Gas Turbine World 1997 Handbook, turnkey power generation projects, p. 24.

The projected COE for the PSOFC/ATS-GT power system is approximately 6% less than the conventional-technology COE at the baseline \$3 fuel cost. However, as Figure 1.3 indicates, the COE advantage of the advanced power system is expected to increase with increasing fuel cost. For example, at \$6 its estimated COE is nearly 11% less than the projected conventional-technology value.

As a task in this study, a conceptual design was developed for the SOFC power conditioning system. The estimated efficiency (DC-to-grid AC) and mature-technology cost of the resulting PCS design were 95% and \$220/DC kWe, and these values are inherent in the overall performance and cost projections presented above. The PCS efficiency is instrumental in determining the overall power system efficiency. For example, a one-point increase in PCS efficiency would increase the PSOFC/ATS-GT system efficiency by approximately 6/10 of one percentage point. Additional study is needed to determine if a PCS efficiency above 95% could be achieved cost-effectively.

Figure 1.3 — Effect of fuel cost on COE estimates.

1.4 Alternative Power System Cycles

Two options were evaluated for using heat recovered from the combined exhaust of two 10 MWe-class PSOFC/ATS-GT power blocks to generate additional electric power. In one, a once-through heat recovery steam generator and steam turbine cycle were

applied. An additional 1.5 MWe were produced, increasing the total power output and efficiency to 26.2 MWe and 63.6%. In the second option, the bottoming cycle was based on an ammonia-water power system concept developed by the Energy Concepts Company of Annapolis, Maryland. In this case the system power boost was 1.9 MWe, which increased system power output and efficiency to 26.5 MWe and 64.4%. However, an economic analysis of each option indicates there may be little COE advantage relative to the baseline PSOFC/ATS-GT power system because the positive COE effects produced by the increased power output and efficiency are countered by the increased power system installed and O&M costs. Note, however, that the COE evaluations did not consider emission taxes, and adding bottoming cycles to the baseline PSOFC/ATS-GT power systems would likely gain economic attractiveness where such taxes were levied.

1.5 Conclusions

- A low-cost, highly-efficient (60%) 20 MWe-class PSOFC/ATS-GT hybrid power system concept has been devised that employs the 4.5 MWe Solar Turbines Mercury™ 50 gas turbine that was developed under the DOE ATS program. The projected hybrid cycle power system COE is 44.8 mills/kWh, which is 6% less than the estimated COE for a conventional 20 MW-class gas turbine combined cycle power system. These estimates apply to a fuel cost of \$3/MMBtu. Due to its higher operating efficiency, the PSOFC/ATS-GT power system would have increased COE advantage at higher fuel costs.
- The PSOFC/ATS-GT power system efficiency estimate is 59.9% (net AC/LHV), twelve points higher than that achieved by the conventional-technology gas turbine combined cycle power system. This estimate is based upon mature-product cell V-I characteristics that are expected to be available in the 2005 to 2010 time period, and upon current Mercury™ 50 performance.
- An SOFC PCS concept was developed that has an estimated DC-to-grid AC conversion efficiency of 95% and an installed cost of approximately \$220/DC kWe.
- By equipping the PSOFC/ATS-GT power system with a bottoming cycle (steam turbine or ammonia-water), the overall system power output and efficiency will increase. However, due to the accompanying increases in system installed and O&M costs, the gains in power output and efficiency do not necessarily translate to significant COE reductions. It is likely that the economic attractiveness of adding a bottoming cycle to the baseline PSOFC/ATS-GT power system would increase if taxes on power system emissions were levied.
- The staged-cell SOFC stack design does not offer the large SOFC efficiency gain (over the standard cocurrent axial flow stack design) that was projected originally. Cell cooling in the fuel-entry cell rows reduces the average cell voltage while there

is little increase in average fuel utilization at the last cell row at meaningful current densities because of the hazard of anode oxidation.

Recommendations:

- Development of the PSOFC/ATS-GT power system concept should continue to refine the conceptual design and performance estimates, establish the method of operating (startup, power operation, shutdown) the PSOFC/ATS-GT power system, identify and resolve SOFC and Mercury™ 50 integration issues, and develop a plan to demonstrate the power system concept.
- The PCS efficiency has a significant effect on the overall power system efficiency. For example, a gain in PCS efficiency of one percent translates to a system efficiency gain that approaches 6/10 of one percentage point. Additional conceptual work should be done to determine if PCS efficiency gains above the 95% value developed and applied in this study could be achieved cost-effectively.
- SOFC development should be pursued to improve fuel cell power density and efficiency. As with the SOFC PCS, improvements in cell efficiency would be effective in increasing the efficiency of SOFC/GT power systems.

2. INTRODUCTION

A simplified cycle diagram for a simple-cycle, atmospheric-pressure SOFC power system is provided in Figure 2.1. Fuel cell process air is supplied by the air blower, and the air is preheated as needed for SOFC thermal management using heat recovered at the recuperator from the SOFC exhaust gas. The SOFC generator operating pressure is near atmospheric, typically being in the 30 to 50 mbarg range. SOFC power systems based upon this cycle are capable of design-point electric generating efficiencies in the 45% to 50% (net AC/LHV) range. The 100 kWe SOFC combined heat and power (CHP) demonstration power system operating in the Netherlands is based on this cycle. This system, designed and fabricated by Siemens Westinghouse and EDB/ELSAM, a team of Dutch and Danish generating and distribution companies, and sponsored by EDB/ELSAM, is installed at a utility site near Arnhem. To date, the unit has logged over 15,000 operating hours, it is generating approximately 110 kWe net AC power at 46% efficiency (net AC/LHV) for the utility grid, and it also produces hot water for the local district heating system. The demonstrated system energy efficiency is nearly 75%.

Figure 2.1 — Simplified atmospheric pressure SOFC power system cycle.

Power system efficiency can be increased by integrating the SOFC generator with a gas turbine in the cycle depicted in Figure 2.2. The SOFC generator in this cycle is pressurized (PSOFC), operating on air coming directly from the compressor discharge, via the turbine exhaust recuperator. The system based upon this cycle achieves high efficiencies due to the utilization by the gas turbine generator of thermal energy in the pressurized SOFC exhaust stream, which is accompanied by a corresponding reduction in the rate at which fuel is required by the gas turbine combustor. The SOFC generator

will typically operate with an air inlet temperature of 550°C (1100°F)* and an exhaust temperature of 870°C (1600°F), while the ATS gas turbine fires at 1160°C (2125°F). Thus, with this engine implemented, fuel firing at the combustor will be reduced by approximately 50% $[(870 - 550)/(1160 - 550) * 100]$. It is noted that power system performance is also enhanced by SOFC generator operation at elevated pressure. For a given cell operating current, cell voltage, and hence cell power output and efficiency, increase with pressure.

Figure 2.2 — Simplified PSOFC/GT hybrid power system cycle.

The purpose of the present project, undertaken by Siemens Westinghouse Power Corporation, in conjunction with Caterpillar, Inc, was to develop a conceptual design for a power system that applies the PSOFC/GT cycle and utilizes Solar Turbines ATS technology to achieve a low cost of electricity relative to the COE for systems based on current conventional power generation technologies. A secondary objective was to achieve a system with high power generation efficiency. The power system capacity was to be approximately 20 MWe. The gas turbine basis for the study was the Solar Turbines Mercury™ 50 engine. The Mercury™ 50 is a recuperated, single-shaft gas turbine that was developed by Solar Turbines under the U.S. Department of Energy ATS program. Operating on natural gas fuel, the engine generates 4.5 MWe with an efficiency of approximately 41% (net AC/LHV). The turbine compressor intakes air at 16.6 kg/s (36.5 lb/s,) and operates with a pressure ratio of 9.3:1 and turbine combustor temperature of 1160°C (2125°F).

* Air temperature as delivered by the recuperator. Additional air heating occurs within the generator before cell entry.

3. RESULTS AND DISCUSSION

3.1 Power Plant Conceptual Design

3.1.1 Design Requirements and Objectives

3.1.1.1 Introduction

This section establishes requirements and specifications that were used in developing the conceptual design for a low cost, high-efficiency fossil-fueled power plant. The plant concept was to be based on a hybrid concept that integrates Siemens Westinghouse SOFC and Solar Turbines Mercury™ 50 ATS gas turbine technologies.

3.1.1.2 Design Basis

Power Plant Application	Commercial distributed-power generation
Output Power Specification	60 Hz AC, at utility-grid voltage
Utility AC Grid Connection	The power plant will be connected to the utility grid, and all net plant power will be exported to the grid.
Dispatch Mode	Base load
Power Plant Heat Recovery	Heat will be recovered for plant power generation support; no heat will recovered for site thermal application.
Power Plant Startup	The utility AC grid will be available for plant startup operations.
Installation	Outdoors.
Conceptual Design Scope	Power plant equipment between the site fuel supply point and the AC grid interface will be included in the design. The equipment considered will be essential for plant startup, operation, control, shutdown, and maintenance.

3.1.1.3 Performance Requirements

Power Plant Design-Point Capacity	20 MW net AC, +/- 2 MW
Design-Point Efficiency	High [approximately 60% (net AC/LHV)], consistent with achieving a minimum COE.
Power Plant Normal-Operation Turndown Requirement	None
Power Plant Overpower Requirement	None
Output Power Conditions	60 Hz Utility-grid quality
Acoustic Noise Control	Consistent with typical gas turbine practice.

3.1.1.4 Fuel and Oxidant Specifications at the Power Plant Design Point

Power Plant Fuel	Natural gas
Fuel composition	<u>(mol percentages)</u>
Methane	96
Nitrogen	2
Carbon dioxide	2
Sulfur bearing compounds	Mercaptans
Sulfur concentration	4 ppmv
Fuel supply pressure	1.034 barg (15 psig)
Fuel supply temperature	15C (59F)
Oxidant	Air (ISO conditions)
Air composition	<u>(mol percentages)</u>
Oxygen	20.7
Nitrogen	78.0
Carbon dioxide	0.3
Water	1.0 (60% relative humidity)
Ambient air temperature	15C (59F)
Ambient air pressure	1.014 barg (14.7 psia)

3.1.1.5 Ambient Air Ranges

Pressure	Sea level to 1500 m (5000 ft)
Temperature	-29C to 49C (-20F to 120F)
Relative Humidity	0 to 100%

3.1.1.6 Power Plant Physical Design Objectives

Major Subsystem Fabrication	Maximum practicable skid mounting at the factory.
Transportation Options	Truck, sea, air, rail
Site Installation Operations	Minimum component/skid assembly at installation site. Interconnect factory-assembled skids at the site. Interface the power plant with the site.
Plant Design Lifetime	Conventional power plant equipment - 25 years.

3.1.1.7 Power Plant Operation

Normal Power Operation Mode	Automatic, unattended, remotely monitored
Power Plant Startup	Attended
Annual Operating Time	Fifty weeks
Annual Planned Shutdown	Two weeks

3.1.1.8 System/Site Interface Requirements

Appropriate interface points will be available at the installation site for:

- Fuel supply
- Obtaining utility AC power during plant startup operations.
- Connecting the plant power output with the utility grid.

3.1.1.9 Economic Evaluation Parameters

Power System Cost Estimation Basis	Costs will be based on the projected needs of mature SOFC/gas turbine technologies and commercial power plant operation, not first-of-a-kind.
Cost Basis	Mid-1998 US dollars

3.1.1.10 Economic Evaluation Parameters (continued)

Fuel Cost	\$3.00/MMBtu (HHV)
Capital Charge Rate	15%
Availability	92%
COE Evaluation Method	Constant dollars
Power Plant Optimization Basis	Minimum COE. The target design-point COE is 10 to 20% below the COE of today's conventional plants.
Transportation Cost Basis	800 km (500 miles) - factory to installation site.
Conventional Power Plant COE Basis	Gas turbine combined cycle

3.1.2 Cycle and Power System Description

A simplified power system cycle diagram for the PSOFC/ATS-GT power system is presented in Figure 3.1. As indicated above, the direct integration of a pressurized SOFC generator and a gas turbine results in the generation of electric power at high efficiency due to the extended processing to power by the gas turbine of an additional fraction of the system fuel energy that is not converted to power by the fuel cell, to the attendant reduction in the flow of fuel to the gas turbine combustor, and to the operation of the fuel cell generator at elevated pressure. Due to its high Brayton cycle efficiency (approximately 41%), which is achieved via the high turbine operating temperature and recuperation, the Mercury™ 50 provides an excellent basis for the hybrid power system.

Figure 3.1 — Simplified PSOFC/ATS-GT hybrid power system cycle.

The system fuel is pipeline natural gas, assumed in this study to consist of 96 vol % methane, 2% nitrogen, and 2% carbon dioxide. The gas also contains sulfur-bearing compounds, occurring naturally, or added to enable leak detection. The sulfur concentration in the raw fuel, per the design requirements, is 4 ppmV, and it must be reduced to the 0.1 ppmV level prior to SOFC generator entry to preclude sulfur reaction (reversible) with SOFC nickel components. After the fuel has been hydrogenated, the desulfurizer in this power system concept processes the sulfur to hydrogen sulfide in a cobalt-molybdenum catalyst bed, and the H_2S is adsorbed on a bed of hot zinc oxide, which operates optimally at 350°C to 400°C (660°F to 750°F). This temperature level is achieved recuperatively and by electric heat addition. The power required for fuel heating is small, and the system is simple and requires low maintenance. Alternatively, the heat for this process could be derived from the turbine exhaust, at the expense of complicating the turbine exhaust design. Hydrogen for the fuel desulfurization process is generated on site.

The SOFC generator produces DC power, which is prepared for export to the utility AC grid by the power conditioning system. AC power is also produced for export by the gas turbine.

The Mercury™ 50 compressor is composed of ten axial stages. Power to drive the compressor, and the AC generator, is derived from a two-stage axial turbine. The compressor air intake rate is 16.6 kg/s (36.5 lb/s), and the design compressor pressure ratio is 9.3:1. As Figure 3.1 indicates, the rotating gas turbine components are installed on a single shaft. This issue was not investigated in detail during this project, but it is anticipated that in the developed power system, the electric load on the gas turbine generator will be modulation-capable to permit the maintenance of set point shaft speed. This feature will assure a steady flow of air to the SOFC generator, an important function of the gas turbine, and a necessity to provide for SOFC thermal management.

A pictorial view of the power system is presented in Figure 3.2. The power system is arranged in two power system blocks. Each block employs a single Mercury™ 50 gas turbine that is integrated with an SOFC generator, and each generator consists of several SOFC modules that are arranged in flow parallel between process air and exhaust manifolds. The number of SOFC generator modules (four) to be associated with each gas turbine has been determined to yield a minimum COE value. Each module is a horizontal cylindrical pressure vessel, flanged at the middle, that houses a cell stack assembly. The basic stack building block is the 576-cell substack pictured in Figure 3.3. The individual cells are tubular, with active lengths of 1500 mm and diameters of 22 mm. They are arranged in the substack in the vertical orientation, with closed ends at the bottom. Shown at the top of the substack are air distribution plena through which process air is admitted to the cell air injection tube inlets, and on the side of the substack, shown are the depleted-fuel recirculation plenum, the ejector that drives the depleted-fuel recirculation, the fuel prereformer, and the ducting for distributing the fresh-fuel/recirculated-fuel mixture to the underside of the substack. Within the substack, methane reformation occurs in the in-stack reformers located between the cell bundles, and the reformed fuel mixture is then distributed to the individual cells. At the cells, the fuel mixture and air flow cocurrently from the cell closed ends, and a fraction of the CO and H₂, typically 85% to 90%, is processed electrochemically. Twenty 576-cell substacks compose the cell stack assembly in a single generator module.

Note the two power blocks are independent. Provided the balance-of-plant equipment is available, one block could be operated while the other block is down for maintenance.

Key:	1. Mercury™ 50 Gas Turbine	10. Auxiliary Air Compressor
	2. Power Conditioning System	11. Natural Gas Compressor
	3. Air Heater	12. Desulfurizer
	4. PSOFC Generator Module	13. Electrical Cabinets
	5. Storage	14. UPS Shed
	6. 20 MW Substation	15. H ₂ Generator/Compressor and Gas Mixer
	7. N ₂ Tube Trailer	16. Water Storage Tank
	8. N ₂ Generator/Compressor	17. Startup Boiler
	9. Propane Tank	18. Control Room

Figure 3.2 — 20 MWe PSOFC/ATS-GT power system site arrangement.

Figure 3.3 — 576-cell substack.

3.1.3 Power System Performance Analysis

3.1.3.1 Analysis Basis

Basic input for the analysis such as fuel composition, fuel supply conditions, and ambient-air conditions were taken from the design requirements, Section 3.1. Additional information on key input is provided in the following:

- Cell V-I characteristic – the V-I characteristic is graphed in Figure 3.4. It is a projected characteristic for the mature SOFC product that will be available in 2005 to 2010. The V-I characteristic applies to operation at 1 atm (abs), 85% fuel utilization, and to a peak cell temperature of 1020°C (1870°F).

Figure 3.4 — Cell V-I characteristic.

- Cell voltage correction for SOFC generator operation at elevated pressure – A cell voltage adjustment for operation at pressures above 1 atm (abs) is presented in Figure 3.5. Given an operating pressure, the corresponding voltage adjustment is added to the base cell voltage from Figure 3.4. Data for Figure 3.5 were obtained from cell testing (Test No. 503) performed at Ontario Hydro Technologies (OHT) in Toronto, Ontario, Canada by OHT personnel. The test article was designed and supplied by Siemens Westinghouse. The tests covered the pressure range from 1 atm (abs) to 15 atm (abs). Over the cell current density range of interest in this conceptual design study, the adjustment for pressure is essentially independent of current density.

Figure 3.5 — Cell voltage adjustment for pressure.

- Stoichs profile – air flow to the SOFC generator is determined by the stoichs profile that is graphed in Figure 3.6. Its application results in cell operation with a peak cell temperature of 1020°C (1870°F) and a generator combustion zone exhaust temperature of approximately 870°C (1600°F). One stoich provides the normal-air flow needed to supply oxygen for the cell electrochemical process. A generator air flow based upon multiple stoichs provides the required amount of oxygen for that process, and it flattens the cell axial temperature distribution, thereby raising the cell average temperature and operating voltage.

Figure 3.6 — Stoichs profile.

- Electrochemical fuel utilization – generator fuel consumption was set at 85%, meaning 85% of the fuel admitted to the SOFC generator was consumed electrochemically on the cell active surface and by the ionic and molecular leakage of oxygen from the cathode side of the cell to the anode. Approximately 84% of the generator fuel is consumed electrochemically.
- SOFC power conditioning system efficiency – 95% (DC-to-grid AC), based on project PCS conceptual design.
- Gas turbine generator efficiency – 96%.
- Gas turbine compressor isentropic efficiency – 87.5%.
- Turbine isentropic efficiency – 87.6%.
- Turbine mechanical efficiency – 97%.
- Turbine gear box efficiency – 98%.
- Auxiliary power losses – 20 kWe [I&C, cabinet ventilation] for one 10 MWe-class power block.

3.1.3.2 Power System Performance Estimates

Power system design-point performance estimates are summarized in Table 3.1. They apply to a single PSOFC/ATS-GT power block, consisting of one Mercury™ 50 gas turbine and an SOFC generator composed of four generator modules. Two such power blocks would be arranged and operated in parallel in a 20 MWe-class power system. Performance estimates for this total system are presented in Table 3.2.

Table 3.1 — Power Block Design-Point Performance Estimates

Compressor air intake rate	16.6 kg/s (36.6 lb/s)
Compressor pressure ratio	9.5:1
Gas turbine combustor temperature	1160C (2125F)
SOFC gross AC power	7.9 MWe
Gas turbine gross AC power	4.8 MWe
Power block net AC power	12.3 MWe
Fuel flow rate to SOFC generator	0.31 kg/s (0.68 lb/s)
Fuel flow rate to gas turbine combustor	0.14 kg/s (0.31 lb/s)
Efficiency (net AC/LHV)	59.9%
Carbon dioxide emission	340 kg/MWh (750 lb/MWh)
Nitrogen oxide emission (based on 5 ppmv addition at the gas turbine combustor.	0.04 kg/MWh (0.09 lb/MWh)
Sulfur oxide emission	0
Exhaust flow rate	16.9 kg/s (37.2 lb/s)

**Table 3.2 — Power System Design-Point Performance Estimates
(Power System Definition: Two Power Blocks)**

Air intake rate	33.2 kg/s (73.2 lb/s)
Fuel flow rate	0.90 kg/s (2.0 lb/s)
Net AC power	24.7 MWe
Efficiency (net AC/LHV)	59.9%
Exhaust flow rate	33.9 kg/s (75.0 lb/s)

Corresponding to the operating point represented in Table 3.1, and for the points identified in Figure 3.7, estimates of cycle statepoint parameters are provided in Table 3.3.

Figure 3.7 — Power block statepoint diagram.

Table 3.3 — Power Block Statepoint Parameter Estimates

Statepoint ID	Mass Flow, kg/h	Temperature, C	Pressure, bara	Molecular Weight	Enthalpy Flux, MW
1	1,620	15	2.05	16.8	-0.010
2	1,620	212	12.65	16.8	0.214
3	2	15	12.65	2.0	
4	1,622	212		16.7	0.214
5	1,622	375		16.7	0.439
6	1,622	399		16.7	0.476
7	1,622	399		16.7	0.476
8	1,622	237		16.7	0.246
9	0				0
10	1,114	93		16.7	0.050
11	508	93		16.7	0.023
12	59,656	15	1.01	28.6	-0.166
13	59,656	15	1.01	28.6	-0.166
14	59,656	304	9.61	28.6	4.795
15	57,687	304		28.6	4.637
16	57,687	631	9.26	28.6	10.395
17	57,687	629		28.6	10.360
18	57,687	629		28.6	10.360
19	58,801	877		28.4	15.851
20	60,950	1160	8.89	28.2	22.467
21	60,950	656	1.06	28.2	12.090
22	60,950	360	1.01	28.2	6.215

3.1.3.3 Elevated SOFC Generator Operating Pressure

As indicated earlier, the SOFC generator design utilizes the conventional generator configuration, and in addition, the design assumes the uniform distribution of air flow between cells, and the uniform distribution of reformed-fuel flow on the cell anode side. In the conventional generator configuration, fuel flows upwards along the cell exterior surfaces (anodes) in the system of communicating parallel flow channels that is defined by the anodes. That this configuration is suitable for atmospheric-pressure generators has been confirmed via current and past field unit demonstrations, and a pressurized bundle at three atmospheres pressure has also exhibited good, stable performance. However, it is not known with certainty that this parallel flow channel configuration is suitable for higher pressure SOFC generators. Flow channel pressure drop is determined by friction and buoyancy components. At low pressures, the friction component, varying inversely with gas density, and hence with pressure, is dominant, and the

buoyancy effects of modest channel-to-channel temperature or composition variations are of little consequence. However, at elevated operating pressures, the buoyancy effect, being proportional to density differences, has more influence, while the friction pressure drop component, with its inverse pressure relationship, is diminished. At higher pressure ($P > 3$ atm) transverse channel-to-channel parameter variations could affect parallel-channel flow stability, and this is of special concern considering the density of the oxidation product stream that exits the channels at the top will be larger than the density of the reformed fuel stream that enters at the bottom. It is therefore clear that cell-stack thermal and hydraulic interactions need to be well understood, particularly relative to stack operation at elevated pressure levels. A very reasonable first step in developing this understanding would be to undertake a detailed analytical evaluation of the anode-side flow field and of the effect of operating pressure on that field.

3.1.4 Power Plant and Site Arrangement

Power system pictorial and plan views are presented in Figure 3.8 and Figure 3.9.

The arrangement includes a trailer mounted control/meeting room with sanitary facilities and four-space parking area, a storage shed, a 20 MW substation/switchyard, two fire hydrants, intake air filter house and the BOP equipment block. The BOP equipment block is shown between the control trailer and switchyard. The BOP equipment block is comprised of several equipment, component and electrical skids. Each skid is mounted to a dedicated concrete pad or foundation. The various skids are in weather-proof enclosures:

- Natural gas desulfurizer
- Pressurized nitrogen storage cylinders
- Nitrogen generator and compressor
- Hydrogen generator, compressor and gas mixing equipment

- Key:
1. Mercury™ 50 Gas Turbine	10. Auxiliary Air Compressor
2. Power Conditioning System	11. Natural Gas Compressor
3. Air Heater	12. Desulfurizer
4. PSOFC Generator Module	13. Electrical Cabinets
5. Storage	14. UPS Shed
6. 20 MW Substation	15. H ₂ Generator/Compressor and Gas Mixer
7. N ₂ Tube Trailer	16. Water Storage Tank
8. N ₂ Generator/Compressor	17. Startup Boiler
9. Propane Tank	18. Control Room

Figure 3.8 — 20 MW PSOFC/ATS-GT power system arrangement.

Figure 3.9 — 20 MWe PSOFC/ATS-GT power system site arrangement — plan view.

Not shown are the security fence and its gates around the perimeter or any storm water drainage features. These features will generally be site specific. Structural support steel, ladders, platforms, overhead high voltage power lines and their poles from the SOFC vessels and GT skid are also not shown.

For this study it is presumed that direct access to the site is via paved roadway capable of bearing the loads from the heaviest items [GT skid weight is 43,100 kg (95,000 lb), and the SOFC vessel half weight with internals is 25,400 kg (56,000 lb)]. The site must also have direct connections to natural gas, city water, sanitary sewer, telephone and electric power.

To minimize field construction, each SOFC vessel half with SOFC stacks installed is factory fabricated on a transportable trailer/skid. This foundation trailer is field mated with the trailer bearing the other half of the SOFC vessel. After proper alignment of the vessel/trailer halves, the trailers are blocked and the wheels removed. In the event that a vessel half is to be returned to the factory for maintenance or repair, the wheels can be reinstalled on the foundation trailer, the blocks removed and the trailer towed away.

Access to the switchyard is available on two sides and the BOP equipment block also has fork truck clearance between skids. The process piping is factory-fabricated internally-insulated flanged sections. Where possible, external jacketed thermal insulation is applied at the factory to minimize field construction. These sections are relatively large and heavy, varying from about 1000 kg per meter (670 pounds per linear foot) to about 1500 kg per meter (1000 pounds per foot). They are assembled on site by landing them on their pipe supports and bolting together the gasketed, flanged joints. Thermal expansion joints are provided between significant runs of piping as needed. The plant arrangement drawings do not show the detail of the flanged joints in the process piping or the pipe supports.

The arrangement uses SOFC pressure vessels with process air entry and exhaust exit at opposite ends. The pneumatic balancing for equal distribution of air to the cells is straightforward. The significant consequence of the vessel with nozzles on both ends is the need to use a relatively large quantity of large diameter, internally and externally insulated piping. This plant occupies the smallest amount of ground of the arrangements studied (see Section 3.4.5). The plot plan measures 59 m (194 ft) by 38 m (125 ft), 0.56 acres in area.

The SOFC generator modules are accessible from each end, and there is space for a vehicle access road at each end. Access for inspection, repair or maintenance is good. Routing of high voltage electric power lines from the SOFC vessels to the switchyard is efficient. The DC power generated in each vessel is routed through thermally and electrically insulated feedthroughs to a nearby package containing a DC to AC inverter with step-up transformer. The inverter/transformer package is mounted in a weather proof cabinet at the site. The inverter/transformer package terminates with three standard high voltage insulators. The three phase 13.8 kV high voltage lines rise from the inverter/transformer package and connect to their respective phase feeder lines. These feeder lines run overhead to the switchyard.

3.2 Systems and Components

A simplified process diagram that identifies systems and components is presented in Figure 3.10. In this process diagram one of two PSOFC/ATS-GT power blocks is shown, and within the power block, only one vessel of four is shown.

3.2.1 SOFC Generator

3.2.1.1 Staged-Cell Generator Concept Evaluation

Two separate analyses were made to evaluate the staged fuel cross-flow stack concept. The original analysis used the coupled thermal/electrochemical model and applied specifically to the tubular SOFC geometry. This analysis and its results are summarized below. The isothermal model was created to independently verify that the benefits of staging were much smaller than had been anticipated. The isothermal model is general and is not limited to a particular cell or stack geometry. The isothermal model is presented in Appendix A. The results of the two analyses are combined in the summary at the end of this section. The analytical models consider an SOFC bundle row consisting of four bundles, each bundle consists of a three-cell-in-parallel by eight-cells-in-series array. The bundle row is flanked by in-stack reformers.

Figure 3.10-PSOFC/ATS GT Hybrid Simplified Process Diagram

3.2.1.1.1 Coupled Analysis of the Staged Fuel Cross-flow Stack

This section gives an overview of the coupled cell/stack analysis model used to evaluate the staged fuel cross-flow stack. The model was based on cross-flow of fuel through four bundles in series. Thus, fuel passes sequentially across a total of 32 cells. The cross flow stack analysis model was developed for a stack that would replace a conventional stack that has axial fuel flow. The staged-cell stack has recirculated fuel and is coupled to the in-stack reformer board (SRB) assembly positioned between each pair of bundle rows. Thus the cell configuration of the stack remains essentially unchanged, with the exception that fuel flows in at one side, through the stack, and exits on the other side. From there the fuel passes back through the recirculation zone to the inlet side of the stack where it is mixed with fuel feed and pre-reformed. The exhaust portion of the spent fuel exits to the combustion zone as it passes through the recirculation zone. A side view of the fuel flow through the stack and SRBs is shown in Figure 3.11.

Figure 3.11 — Staged-cell generator concept.

The model is a fully coupled heat transfer, mass transfer, and electrochemical analysis. Local temperatures are used to evaluate all local properties including thermal conductivity, mass diffusion coefficients, electrical resistivity, and gas transport properties. Local temperatures and concentrations are used to calculate the Nernst potential, cell resistance, polarization losses, and the local current density. Fuel and oxidant streams are depleted based on local current density. Local heat generation includes the excess heat of reaction, Joule and polarization losses, and heating due to O₂ ionic leakage through the cell. Heat transfer coefficients are also based on properties evaluated at local temperatures. Radiation heat transfer is calculated based on the fourth power law. For all models that include an SRB, the temperature and the heat flux from the cell are matched with the temperature and the heat flux to the SRB. The reformation model of the SRB satisfies heat transfer, mass diffusion, reaction rate and chemical equilibrium conditions.

The development of the analysis model for the cross flow stack is based on a similar model that has been used for the conventional stack. To include a number of stages, the number of axial segments per stage between the closed end of the cell and the fuel recirculation zone was reduced to 20. Further, it was decided to divide the 32 cell stages into groups of cells. The cells in each group would be represented by the conditions at the center cell of the group. The group arrangement selected is shown in Figure 3.11. To provide detail at the inlet fuel side of the stack, the first two groups include only one cell each. The third group includes 2 cells, the fourth group includes 4 cells, and the fifth and the sixth groups include 8 cells each. The seventh and eighth groups include seven cells and one cell, respectively. Note that the bundle and SRB boundaries coincide with group boundaries. This was required to simplify the coupling of the SRBs and the cells.

The initial results from the staged configuration were disappointing. The problems can be described as follows.

1. The first cell row (at the fuel inlet) was too cold and final cell rows were too hot. This is a consequence of the large flow of relatively cold fuel that impacts directly onto the first stage or the first cell row. The latter cell rows become hotter as the fuel temperature increases and cell losses increase as fuel is depleted.

2. The overall temperature uniformity was poor. This is a result of the combination of the superposition of the row-to-row temperature distribution on the axial temperature distribution. The overall non-uniformity is the sum of the two.
3. The fuel concentration becomes non-uniform in the latter cell rows. The fuel is consumed at a higher rate at the hotter portions of the cell due to the lower local cell resistance. After passing a number of cell rows, the fuel becomes depleted in the central region of the cells. To preclude anode oxidation, the average fuel utilization at bundle row exit must be comparable to that for an axial-flow cell stack — a higher fuel utilization is not feasible.
4. The average cell voltage was lower than for an axial flow stack due to the lower overall average temperature.

Several modifications were made to the model stack to address the issues of temperature uniformity.

1. The inserts normally used to enhance heat transfer within the combustion zone region were removed from the air feed tubes in the last three cell bundles (24 cell rows). Removal of the inserts results in reduced pressure drop which acts to increase the cooling flow of air to the last 24 cell rows relative to the cooling flow to the first 8 cell rows. Removal of inserts also reduces the pre-heating of cooling air to the last 24 cell rows, thereby lowering their temperature relative to the inlet cell rows.
2. The fuel flow to the SRB adjacent to the first cell bundle was restricted by an orifice. This reduces the cooling of the first 8 cell rows at the inlet side of the stack. It also increases cooling of the last cell rows by shifting the reformation heat load toward the exit side of the stack.

3.2.1.1.2 Results of Coupled Analysis of Staged-Cell Stack

Results for the staged-cell stack with the modifications listed above are shown in Figure 3.12, Figure 3.13, and Figure 3.14. The cell current was 253 A/cell corresponding to a current density of 304 mA/cm². Air inlet temperature was selected to give a maximum local cell temperature of 1020°C (1870F). The temperature distributions of the cells at the center of each of the 8 cell groups are shown in Figure 3.12. Figure 3.13 displays the average cell temperature and the terminal voltage as a function of the cell row position. The fuel concentration profiles at the exit of the 8 cell groups are shown in Figure 3.14.

TCELL, Cell temperature distribution for each Group.

Figure 3.12 — Temperature profiles in staged-cell generator.

Figure 3.13 — Cell voltage and average temperature in a staged-cell generator.

YFX, Fuel Mole fraction distr., exit each Group.

Figure 3.14 — Fuel mole fraction distribution in a staged-cell generator.

The stack model modifications made significant improvements in the stack temperature uniformity, but the predicted performance of the cross-flow stack was still lower than for a conventional stack with the same maximum temperature limit. This result was obtained although several idealistic assumptions were made for the analysis which enhance performance relative to what would actually be expected. These assumptions included

1. The lateral heat losses at fuel inlet and exit were neglected. Including them would further reduce the temperature of the first cell rows.
2. It was assumed that the fuel mass flow rate was uniform per unit axial length of the cell and bypass of fuel at the inactive ends of the cells was neglected. Actual fuel flow through the central portions of the cells would be reduced due to the effect of higher temperature by the combined effects of low density and high viscosity. This is the same elevation where the current density — hence, fuel consumption, is higher than the average rate.
3. The influence of dimensional tolerances on fuel flow distributions was ignored. The actual fuel flow distribution would be sensitive to the lateral spacing between cells and to the cell-to-SRB gap.

Even with these assumptions, the fuel utilization in the staged cell design is limited due to the non-uniform consumption of fuel as it passes each cell. The fuel concentration at the exits of each stage at 92% fuel utilization is shown in Figure 3.15. In Figure 3.16, the partial pressure of O_2 at the fuel electrode is compared to the equilibrium partial pressure of O_2 for the $Ni/NiO/O_2$ reaction. The comparison is made for the exit stage cell. Oxidation of the fuel electrode is likely whenever the O_2 pressure is within two orders of magnitude of the equilibrium pressure. It is seen that this limit is exceeded for much of the cell length (data points are equally spaced).

Figure 3.15 — Fuel concentration at the exit of each group of cells.

Figure 3.16 — Nickel fuel electrode oxidation limit – cross flow stack at FU = 92%.

3.2.1.1.3 Summary of Staged-Cell Generator Concept Evaluation

The conclusion from the isothermal model analysis in the Appendix is that there is only a small improvement in stack efficiency available due to cell staging at current densities near the point of maximum power. At the maximum power point, the improvement in power output due to staging is less than 0.6% for 85% fuel utilization, and less than 1.2% for 95% fuel utilization.

In the coupled model analysis the small benefits of staging were eroded by the effects of fuel cross flow on the overall stack temperature distribution. The inlet fuel for 32 cells impinges on each of the cells in sequence. This high flow of relatively cold fuel removes several times more heat from the first cell row than is generated. Radiation from the neighboring rows provides a large amount of heat but it is insufficient to achieve satisfactory temperatures in the first rows. A way to improve the effects of cold fuel would be to heat the fuel exiting from the SRBs. We were unable to find a reasonable way to effect heat exchange between the inlet fuel and a hotter exhaust stream without incurring significant pressure drops and complicating the stack design.

The previously assumed benefit of higher fuel utilization with a staged-cell concept was not available due to uneven consumption of fuel in the direction perpendicular to the fuel flow. Fuel was locally depleted to the point of fuel electrode oxidation before the fuel utilization reached 95% even with the optimistic fuel flow distribution assumptions used in the model. Due to this effect and to the cool-cell effect in the fuel-entry rows, it was concluded that the SOFC generator module design should retain the conventional co-current fuel flow and air flow configuration.

3.2.1.2 SOFC Generator Design Description

Much of the SOFC generator design philosophy and implementation is influenced by technology derived from two sources: the conventional 100 kWe SOFC generator operating at atmospheric pressure and the pressurized SOFC generator in the 220 kWe hybrid cycle power system which is designed to operate at 3 atmospheres.

The generator submodule design philosophy is an extrapolation of the current 200 kWe SOFC generator configuration with the exception that the stacks are housed in a hori-

zontal pressure vessel. Each stack assembly is installed on a common removable support structure lined with high performance thermal insulation.

This configuration greatly simplifies assembly of internal components by utilizing common parts and a modular, reusable insulation system.

The PSOFC generator design employs the air electrode supported (AES) fuel cell derived from the porous support tube (PST) cell. To date, PST test cells have achieved over 65,000 hours of operation. The AES fuel cell operating and durability history of over 32,000 hours reflects an improved design with significant advancements in material composition and processing.

Figure 3.17 shows a simplified array of closed tubular cells composed of concentric electrodes separated by a layer of solid electrolyte. Fuel for the electrochemical reaction flows upward between the tube exteriors, while process air flows upwards in the annular space between the air feed tube and the cell inner surface. Unreacted or spent fuel is burned above the bundles of tubes to preheat the incoming air. Nickel felt contacts provide flexible electrical connections between cells.

Figure 3.17 — Schematic fuel cell array.

Each substack contains 576, 22 mm OD, 1500 mm active length tubular fuel cells, each capable of generating over 200 watts, arranged in 6 bundle rows. The cells are arranged in three parallel paths for improved reliability.

The submodule configuration is depicted in Figure 3.18. As shown, it includes ten fuel cell substacks, each fueled at the base by a fuel prereformer with integral fuel distribution manifolds and recirculation plenum, air manifold, DC power leads and stack internal insulation.

Figure 3.18 — Ten substacks isometric view.

Each substack is fed from a fuel supply system including a recirculation loop, ejectors, a prereformer and a fuel manifold with riser tubes. The recirculation plenum is used to mix the depleted fuel extracted from the stack with the fresh incoming fuel injected through an ejector nozzle. The mixture is then directed to a prereformer chamber where higher hydrocarbons are reformed to prevent carbon deposition in the stack reformers where full reformation of methane occurs. From the prereformer exit, the fuel mixture is distributed to each stack through a series of 7 bottom manifolds.

Process air is introduced into the submodule through one inlet nozzle connected to a centrally located air duct branching to individual smaller ducts feeding a large air plenum located on top of each stack. Each air plenum supplies air to 576 adjacent cells through ceramic air feed tubes. Air flows from the air plenum into the air feed tubes, which convey the oxidant to the lower, closed-end of each fuel cell.

Exhaust from the substack is collected from side manifolds and ducted through larger side ducts terminating in a common transition fitting connected to the pressure vessel exhaust nozzle.

Fuel Cell Substack. The tubular SOFC cell features a porous air electrode made of doped lanthanum manganite ceramic material. A gas-tight electrolyte layer of yttria-stabilized zirconia covers the air electrode, with the exception of an axial strip along the entire active length of the cell. This strip of bare air electrode material is covered by a thin, dense layer of doped lanthanum chromite. This layer, known as the cell interconnection, serves as the electric contact area to an adjacent cell or to a power contact. A top layer, the fuel electrode, is a nickel-zirconia cermet and covers the electrolyte surface except in the vicinity of the interconnection.

To construct an electric SOFC generator, individual cells are bundled into an array of electrically connected fuel cells, forming a monolithic structure that constitutes the basic generator stack.

The proposed fuel cell substack consists of a number of bundle arrays of 3x8 cells with an active (interconnection) length of 150 cm (59 in.). The cells are electrically interconnected in three parallel paths, each with 192 cells connected in series.

The three series-connected paths are intermittently connected by parallel shunts to improve reliability and gross electrical performance of the cell substack. An exploded view of the substack, as shown in Figure 3.19, is composed of 6 groups of cells, referred to as bundle rows, separated by electrically insulating in-stack reformers. Each bundle row includes 96 cells (three parallel paths of 32 series-connected cells).

These bundle rows are series-connected in a serpentine configuration and terminate at power take-offs that transfer current to the two opposite power leads. The metallic power leads may be air-cooled to withstand the high temperature environment surrounding the cell substack.

Each fuel cell substack is assembled externally on a common support structure prior to final installation into the pressure vessel.

Figure 3.19 — Exploded view of 576-cell SOFC substack.

Stack Liner. Each cell substack is enclosed with a liner assembly composed by vertical panels made of Haynes® 214 alloy foil, 0.25 mm (0.010 in.) thick, and each panel is joined full length by resistance or fusion welding process to form four individual subassemblies which are welded at the four corner seams.

The liner assembly incorporates a bottom metallic sealing with the lifting tray and the fuel manifold penetration sealing typical of the 220 kWe PSOFC design. This type of design minimizes fuel leakage and it provides an effective barrier between the cell substack and the outer insulation made of low purity material.

Each panel includes a layer of “puff” board tiles bonded to the surface facing the substack. “Puff” boards are manufactured from layers of Saffil blanket material saturated with an organic binder and compressed to a 3.2 mm (0.125 in.) thickness to obtain the required density. During generator start up, the organic binder burns off and the Saffil layer is free to expand and fill the volume between the cell stack outer wall and the liner, thus preventing fuel bypass which may occur as a result of foil warpage under high temperature conditions. Each panel includes a series of “puff” board tiles, 305 mm (12 in.) wide, 406 mm (16 in.) high and 3.2 mm (0.125 in.) thick, glued to the stack with an organic adhesive.

Each liner panel includes two edge flexural flanges that are formed to a special shape designed to absorb thermal differential expansion of the panel in the direction normal to the flange length. The contour of the flexural flange is designed so that it achieves

maximum displacement, minimum stress when the panels are welded side by side to form a complete subassembly; these multiple ribs produce a substantial stiffness and load capacity improvement in the direction of the ribs. The ribbed or corrugated sheet basically exhibits one way flexural behavior, since the perpendicular direction retains the properties typical of a flat plate.

Fuel Distribution System. The Fuel Distribution System includes fuel nozzle housing, a recirculation plenum system, ejector, prereformer and an array of tubular manifolds connected to vertical risers penetrating the SRB's assemblies. An isometric view of the Fuel Distribution System is shown in Figure 3.20.

Figure 3.20 — Substack fuel distribution system.

Fresh fuel is injected through an ejector nozzle that entrains depleted gas extracted from the upper zone of the fuel cell substack. This fuel mixture is directed to a pre-reforming section where partial reformation occurs within a catalytic bed.

The fuel stream then exits the pre-reformer and is manifolded through a piping network connected to horizontal manifolds installed at the base of the substack. These manifolds are coupled at the periphery of the stack by a quick connect flange assembly.

Each manifold includes a number of vertical risers that direct the fuel to the top of the in-stack reformer internal cavity. The preponderance of fuel reformation occurs in this area and a hydrogen-rich stream is fed at the base of the stack to the exterior of the tubular fuel cells. Complete reformation is finally achieved at the closed end of the fuel cell.

In-Stack Reformers. The in-stack reformer concept was developed to improve the SOFC system efficiency by reducing the airflow pumping requirements necessary to maintain cell axial temperature distributions within an acceptable range. Additionally, lower airflows reduce the required effectiveness (surface area) of the internal and external recuperators.

The SRB concept provides a large surface area for catalytic activity and heat transfer required for hydrocarbon reformation. It removes thermal energy uniformly along the entire cell length thereby resulting in sufficiently flat axial temperature distribution which helps to maintain acceptable axial temperature gradients within the substack. Moreover it has a much lower pressure drop than a conventional packed bed type external reformer.

The in-stack reformers each consist of a catalytically active inner assembly surrounded by a nickel foil liner, an outer board assembly and an embedded riser tube. Heat is radiated from the adjacent cell bundle region through the porous alumina outer board to supply the energy required for the endothermic reformation reaction.

As explained in the previous section, the in-stack reformers accomplish 85 to 100% of the reformation depending on the generator operating point.

The following is a summary of functional requirements of the in-stack fuel reformation system:

- control the cell axial temperature distribution to minimize voltage loss and maintain an acceptable axial temperature gradient.
- minimize the impact of prereformer catalyst replacement on system availability.
- provide adequate structural support for the upper air plenum and air feed tubes.
- assure electrical isolation to avoid insulation breakdown and short circuits within the stack at the maximum voltage potential in open circuit.
- maintain reformer pressure drop consistent with the capability of the recirculation ejector.
- maintain adequate fuel flow distribution uniformity within the cell stack.
- minimize stack plus reformer size and weight.
- minimize reformation cost.

Air Supply System. Process air is introduced at the top of the generator submodule into an array of air plenums. Air flows from the air plenums into the air feed tubes, which are coaxial with the fuel cells and convey the air to the bottom of the cell. The air subsequently flows upward through an annulus between the feed tube and the cell inner surface.

Spent air exiting from the open end of the cells enters the combustion zone and reacts with spent fuel to completely consume the remaining fuel. Because the air feed tubes cross the combustion zone in a manner similar to a gas-to-air heat exchanger, incoming air is heated by the exhaust gas exiting the cell stack. The exhaust is directed to an upper dome prior to being manifolded to an external common exhaust duct.

High-pressure exhaust gas is collected from all submodules and is conveyed to the gas turbine through insulated piping. The turbine drives the compressor for compressing air that is delivered to the fuel cells.

Internal Thermal Insulation Package. To minimize heat losses from the SOFC stacks and at the same time maintain the pressure vessel wall temperature within reasonable limits, it is necessary to embed all the internals in thermal insulation without impairing the capability to easily service and replace internal components including the generator stack.

Typically, stack insulation within the stack liner is constituted by high purity alumina which exhibits excellent chemical stability at high temperature in both reducing and oxidizing atmosphere.

All elements of the stack insulation structure are typically vacuum formed by utilizing alumina fibers and binders in specific ratios to yield the desired thermal and mechanical properties. These elements include fuel manifold support boards, fuel distribution boards, in-stack reformer outer boards, stack end boards, lower/upper-positioning boards, recirculation and combustor boards.

The insulation package external to the stack liner consists of a number of highly efficient thermal insulation blocks and panels interlocked to form an effective thermal barrier system between the cell stack operating at 1000°C (1800°F) and the pressure vessel at ambient air. This insulation material is fitted between each substack and around all main manifolds and piping spool pieces.

The insulation blocks and panels are constructed with silica material that has a microporous structure and is reflective to thermal radiation.

An additional layer of low density fiber material is installed between the ceramic cell stack and the composite blocks to form a compliant layer capable of accommodating cyclic thermal differential expansion while offering effective bypass gas sealing and adequate lateral load support.

PSOFC Module Description. The pressurized SOFC generator module includes twenty substacks (576 cells) electrically connected in series/parallel to the module terminals. Advantages of this type of construction include ease of assembly, flexibility in power output requirements, better serviceability and substack replacement capability.

The substacks are installed within a pair of horizontal pressure vessels containing a supporting structure designed to provide easy access to the internals and easy assembling and replacement of the submodules without interference with the installed internal components.

Pressure Vessel. The pressure vessel, as shown in Figure 3.21, is a horizontal cylindrical shell and is supported by two saddles anchored to the shipping container structural frame.

Figure 3.21 — Pressure vessel isometric views.

Pressure bearing components are required to meet construction codes such as the American Society of Mechanical Engineers (ASME) B31.1 piping code or the ASME Boiler and Pressure Vessel (BPV) Code. The allowable stresses for materials currently approved for construction under the rules of Sect. VIII, Div. 1 of the ASME BPV Code are provided in various Code Cases or in Sect. II, Part D.

The design of the components is in compliance with the ASME BPV Code, Section VIII, Division 1. The cylindrical shell material is SA-516, Grade 70 carbon steel which provides a minimum tensile strength 480 MPa (70,000 psi). The cylindrical shell thickness is determined by the tangential stress due to the design pressure. Since the maximum longitudinal stress ($PR/2t$) is only half of the maximum tangential stress, one-half the shell thickness is available for the longitudinal bending stress due to weight at the mid-span or in the plane of the saddles, assuming the vessel to behave as its own carrying beam.

As shown in Figure 3.21, two saddles including four legs and stiffener plates support the pressure vessel. The saddles are welded to the outer shell of the vessel with a 120-degree contact angle, and they could be anchored to a concrete slab in the field. The saddle reactions are highly concentrated and induce localized stresses in the shell, which are within the allowable stress conditions specified by the ASME Code.

The ASME flanged and dished head is cold formed from the same material as the pressure vessel. Each head will include one large central air inlet or exhaust outlet nozzle and smaller penetrations for the electrical DC power leads or for internal instrumentation cables.

The shell flange as shown in Figure 3.21, has an outer diameter of approximately 3.5 m (137 in.) and a length through the hub of 280 mm (11 in.) and is forged from SA-105 carbon steel material. Seventy-two 38 mm (1.5 in.) diameter radial holes are drilled through the flange for the bolts. The overall weight of each flange is approximately 2450 kg (5400 lb).

Bolting material for pressure connections must conform to the specifications listed in the ASME Code. Specifically, for this design, SA-193 Grade B7 (1 Cr-1/5 Mo) ferritic steel bolts have been selected. To minimize galling when the bolts are tightened, fasteners are made up with a thread lubricant such as Molykote paste. The proposed

bolting configuration includes 72 bolts with a 35 mm (1.4 in.) bolt diameter in a bolt circle of 3.4 m (134 in.).

Two of these pressure vessels bolted together at their flanges, as shown in Figure 3.22, form the vessel for a single PSOFC generator module.

Figure 3.22 — PSOFC generator module isometric view.

A detail drawing of the pressure vessel and associated Bill of Materials is shown in Figure 3.23.

Figure 3.23 — Pressure vessel detail drawing.

Internal Sub-module Supporting Structure. As shown in Figure 3.24, the pressure vessel incorporates internally a supporting structure that is embedded in high temperature insulation material. This supporting structure is composed by two tubular steel side members separated by a series of equally spaced high density ceramic boards over which a steel deck is finally installed.

Figure 3.24 — Stack support car.

The tubular members include a series of high temperature roller assemblies including a wheel, axle box; die steel caged roller bearing and high load rating axle. These wheel assemblies are typically used in furnace and kiln cars for a variety of heat treatment processes.

This rolling structure provides a convenient open deck for stack assembly, support for piping, ducting and electrical instrumentation without interfering with the assembly/disassembly of the stacks. Once the internals are fully assembled on the support structure, the entire assembly is rolled into the pressure vessel over two rails welded to the shell wall. Once in position, the wheels are locked and secured.

This method of supporting and transporting the stack submodules is particularly advantageous for in-situ stack replacements or quick field installation in the event that the pressure vessels are procured locally or are already installed at the power plant.

A complete assembly of the module internals, including the SOFC stacks, ducting and supporting structure is shown in Figure 3.25.

EXPLD STATE: STACK_OUT

Figure 3.25 — Isometric view of module with internal components.

Internal Air/Exhaust Ducting. When the substacks are inserted and positioned within the pressure vessel, they are connected to a common central air feed manifold by utilizing individual spool pieces connected to the inlet nozzle of each substack. The central manifold uniformly distributes the incoming pressurized process air to the individual substacks. The air manifold is positioned longitudinally within the pressure vessel, is supported by the ceramic frames and is embedded in the same insulation material surrounding the supporting structure. The final connection of the air manifold is made in correspondence to the side flanged cover nozzle through an expansion joint/adaptor.

The exhaust gas flow from each substack is directed to an exhaust collecting plenum overhanging each substack and is subsequently manifolded into a central duct exiting through the exhaust outlet nozzle.

The air ducting operates at a maximum temperature of 760°C (1400°F) and the exhaust ducting temperature does not exceed 870°C (1600°F). Because the internal differential pressure between process air and exhaust is less than 60 mbar (0.9 psi) as a result of only frictional losses and generator pressure drop, ducting rather than piping is utilized internally. Externally, all piping is fabricated to withstand a 10 barg (150 psig) design pressure.

Fuel Inlet Piping. The fuel inlet piping will operate at a pressure of 3.5 to 5 bar (50 to 70 psi) differential inside the pressure vessel and up to 12 to 14 bar (180 to 200 psi) externally. The final pressure will be determined on the basis of the selected generator operating range, fuel composition and gas turbine selection.

Electrical Interconnections. Internal electrical interconnections between stacks are accomplished through utilization of a main DC bus bar and flexible cable connectors clamped to each stack power lead.

High reliability electrical feed-throughs are utilized on the pressure boundary of the module in order to guarantee sound electrical connection between the internal stacks and the external power conditioning system.

Generator Instrumentation. The SOFC generator module is equipped with a variety of instrumentation providing for automatic control with manual capability for plant operation, monitoring and diagnostics.

The generator stack instrumentation consists primarily of DC voltage taps and cell stack thermocouples. Within the cell stack, there are a number of voltage taps monitoring the progressive buildup of accumulated cell voltages with the first tap near ground potential and the last at the maximum DC voltage. The stack external terminal voltage is also monitored as well as the main generator module terminals.

Several thermocouples are embedded within each cell stack at different elevations to monitor and control the temperature of the generator stack.

A number of pressure taps may be included into each stack fuel supply system to monitor differential pressures around the fuel ejector systems and to provide gas sampling as required.

3.2.2 Gas Turbine System — Caterpillar Solar Turbines Mercury™ 50

3.2.2.1 Integration of Mercury™ 50 to PSOFC/ATS-GT System

INTRODUCTION

Selection and integration of a well matched gas turbine engine to the pressurized solid oxide fuel cell subsystem is a key factor in determining the lowest COE integrated system. The Caterpillar Solar Turbines Mercury™ 50 recuperated gas turbine engine was chosen as the gas turbine to integrate in this lowest cost of electricity PSOFC/ATS-GT concept system. See Figure 3.26, Figure 3.27, and Figure 3.28.

Figure 3.26 — Mercury™ 50 recuperated gas turbine, pictorial view.

Figure 3.27 — Mercury™ 50 recuperated gas turbine, plan and elevation views.

Figure 3.28 — Mercury™ 50 turnkey system test at solar turbines, San Diego, California.

The Mercury™ 50 gas turbine was chosen due to its high thermal efficiency, power rating, modular design, reliability, and low cost of maintenance. The design point specifications and pertinent fuel cell integration points of the Mercury™ 50 as a stand alone gas turbine engine at 80 and 59°F ambient, sea level conditions are as follows:

Parameter		
Ambient Air Temperature, °C (°F)	27 (80)	15 (59)
Turbine Electric Power, MWe	4.21	4.57
Overall Efficiency- Generator Busbar, %	40.0	41.5
Natural Gas Fuel Flow, kg/h (lb/h)	791 (1742)	828 (1823)
Overall Airflow, kg/s (lb/s)	16.0 (35.3)	16.6 (36.5)
Compressor Pressure Ratio	9.25	9.33
Recuperator Airflow, kg/s (lb/s)	15.5 (34.2)	16.0 (35.3)
Recuperator Air Outlet Temperature, °C (°F)	631 (1168)	621 (1149)
Recuperator Air Outlet Pressure, bara (psia)	8.8 (127.5)	9.1 (131.8)

The performance quoted is based on the following component parameters:

Parameter	
Generator Efficiency	97.0%
Gearbox Efficiency	98.7%
Fuel LHV	20,610 Btu/lb
Inlet Air Relative Humidity	60%
Compressor Inlet Delta P	0
Exhaust Delta P	3.8%
Compressor Pressure Ratio	9.33
Combustor Outlet Temp	1208°C (2206°F)
Combustor TRIT	1163°C (2125°F)
Compressor Flow	16.7 kg/s (36.5 lb/s)
Recuperator Gas Flow	16.8 kg/s (36.9 lb/s)
Fuel Added	0.23 kg/s (0.5 lb/s)
Air Lost	0.09 kg/s (0.2 lb/s)
Combustor DP/P	3.0%
Compressor Adiabatic Eff	87.5%
Expander Adiabatic Eff	87.6%
Recuperator Effectiveness	92%
Recuperator Air Out Temp	621°C (1149°F)

The list price of the Mercury™ 50 gas turbine engine in 1999 US dollars is \$1,950,000 for the base engine and \$2,450,000 for a complete turnkey system. The total maintenance cost of the Mercury™ 50 system is \$0.00377/kW-hr. The price of the Mercury™ 50 includes \$250,000 required to integrate the engine to the pressurized, solid oxide fuel cells and modification to the combustor subsystem, especially modifications required in high temperature materials and insulation required of the higher 870°C (1600°F) combustor air inlet temperature.

Modifications to Mercury™ 50. The combined Mercury™ 50 / PSOFC system was optimized by Siemens Westinghouse with input from Caterpillar Research regarding the Mercury™ 50 performance. The combined PSOFC/ATS-GT system optimized for the lowest cost of electricity resulted in the Mercury™ 50 turbine operating very near its 59°F/sea level design point. The lowest COE point resulted in the Mercury™ 50 running at a pressure ratio of 9.5:1 versus a design point of 9.3:1 which resulted in a power output of 4.8 MWe versus a design point of 4.6 MWe. A minor adjustment to the compressor variable geometry opening stators 5.7 degrees was required to rematch the Mercury™ 50 to the integrated PSOFC/ATS-GT optimum point. Combustor outlet

TRIT was maintained at 1163°C (2125°F) with overall thermal efficiency of the engine dropping only 0.5 point.

The combustor will be receiving airflow from the PSOFC generator modules, containing approximately 16% oxygen, at 870°C (1600°F) versus an engine design point airflow temperature from the recuperator of 621°C (1149°F). The eight injectors of the combustion system shown in Figure 3.29 and Figure 3.30 will need to be resized to maintain the proper optimum combustion characteristics; however, the combustor is required to raise the engine flow 318°C (605°F) versus a normal temperature rise of 586°C (1056°F) for the base Mercury™ 50 engine design point on a 15°C (59°F) day. Modifications to the combustor main case would need to take place to be able to withstand the 870°C (1600°F) PSOFC exhaust. Incorporating internal insulation, similar to that used in the PSOFC exhaust piping, would be the prime path; however, redesign of the combustor case using high temperature materials should be investigated in any future design.

Figure 3.29 — Mercury™ 50 combustor.

Figure 3.30 — Mercury™ 50 injectors.

Physically, with the combustion section at one end of the engine, the modular and innovative design of the Mercury™ 50 facilitates the easy integration of the engine with the PSOFC system. As shown in Figure 3.26, the Mercury™ 50 normally has a single connection between the recuperator air outlet and the combustor at the 12 o'clock position. The recuperator air outlet will be extended axially outward for connection with the air heater inlet and the turbine combustor section will be rotated 90 degrees for easy connection with the PSOFC exhaust as depicted in Figure 3.31. As conceived, these combustor modifications are expected to increase the price of the Mercury™ 50 engine by US \$250,000.

Figure 3.31 — Modifications required for PSOFC integration.

3.2.2.2 Mercury™ 50 Recuperated Turbine Engine Background

In recognition of the fact that gas turbines are playing an increasingly important role in power generation, the U.S. Department of Energy (DOE) is supporting the development of advanced, land-based gas turbine systems that meet stringent emissions requirements while significantly increasing fuel efficiency. Notable exceptions notwithstanding, government support has been largely absent from industrial turbine development activities to date. The consequences of this absence have essentially been twofold: 1) product development activities that are almost exclusively market-driven; and 2) limited incremental improvements to the underlying technology. The financial underwriting that comes from government support of development activities frees companies to pursue step improvements in the underlying technology without unduly risking profitability in the process.

The ATS concept was jointly created by the DOE and the industrial turbine industry to develop highly efficient, environmentally superior gas turbine systems for both utility and industrial power generation applications. By co-developing a set of symbiotic goals between the government and industry at the onset of the program, the ATS program has secured strong support from both the public and private sectors. As a result, the

program has come to represent one of the most successful government/industry partnerships in recent years, one in which both the public and private sectors will realize significant returns on their collective investment. This section of the report defines the ATS program and product goals, and explains Solar's involvement as well as the salient details of the Mercury™ 50 gas turbine and package design.

PROGRAM GOALS

The stated objectives of the industrial ATS program are a joint representation of the societal goals of government combined with the economic goals of the commercial marketplace (U.S. DOE, 1994), involving efficiency, environment, fuel flexibility, cost of power, and reliability and maintainability. These goals are based on quantitative improvements relative to the markets and systems available in 1991.

Efficiency. A 15 percentage point improvement in the efficiency of the gas turbine system compared to the best technology in its class that is available in the market (1991). The efficiency is based on the lower heating value (LHV) of natural gas.

Environment. Environmental superiority under full and part-load operating conditions without the use of post-combustion controls. Oxides of nitrogen (NOx), carbon monoxide (CO) and unburned hydrocarbon (UHC) emissions will be lowered to achieve acceptance in severe non-attainment areas.

Fuel Flexibility. Natural gas fired ATS systems are to be adaptable to future firing with biomass and coal-derived fuels.

Cost of Power. Busbar energy costs at 10% less than current state-of-the-art turbine systems (1991) meeting the same environmental requirements.

Reliability and Maintainability. Equivalent to current state-of-the-art turbine systems (1991).

The DOE's offices of Fossil Energy, Energy Efficiency and Renewable Energy are jointly responsible for the ATS program. Management of the program is coordinated with the Environmental Protection Agency, the Electric Power Research Institute, the Gas Research Institute, gas turbine manufacturers, the DOE and others. The ATS program was developed in response to the Energy Policy Act of 1992 and outlines an eight year, four-phase timetable. Supporting research, development and evaluation of key enabling

technologies is being carried out via a university consortium, consisting of more than 40 colleges and universities across the United States, coordinated and administered by the South Carolina Energy Research and Development Center at Clemson University under the ATS umbrella.

Since the industrial and utility markets are distinctly different in terms of users and marketplace requirements, the ATS program has been segregated into two separate elements to independently address these needs. Industrial gas turbine systems are differentiated from their larger utility counterparts on the basis of mass flow: industrial turbines are classified as those with a mass flow greater than 10 kg/s (22 lb/s), but not to exceed 55 kg/s (120 lb/s). Though funded separately, the Ceramic Stationary Gas Turbine (CSGT) program, for which Solar is the sole contractor, is also considered a part of the overall ATS program.

SOLAR'S ROLE IN ATS

On September 14, 1995, Solar Turbines was awarded a six-year contract to develop two sizes of ATS products. Solar's efforts can be viewed as three distinct parts based upon a logical partitioning of the development tasks. Phase 3A covers the design of the ATS and the component level testing necessary to validate the higher risk elements of the design prior to proceeding to the next phase of the program. Phase 3B addresses the integrated subsystem testing to further demonstrate suitability of the design(s) at the system level, as well as the assembly and test of the demonstrator packages themselves. Finally, Phase 4 will encompass the field installation and test of the ATS in existing production facilities as a demonstration of the commercial viability of the final design.

Solar's approach to product improvements and introduction has traditionally followed marketplace demand. The purchasers of Solar's products, over 80% of whom are repeat customers, claim that the commercial success and viability of next-generation products, such as ATS, in the industrial marketplace will depend upon several factors:

- High reliability and durability
- Customer perception of low risk
- Broad power range with excellent turndown
- World-class product support

- Low first cost
- Low life-cycle cost
- Ability to permit the installation in the face of stringent regulatory environments.

PRODUCT GOALS

In developing the Mercury™ 50 gas turbine concept, Solar considered these market-place requirements in conjunction with the DOE/industry ATS program objectives. The most significant demand peak was noted at approximately 5 MWe in output, driven primarily by the industrial power generation and emerging distributed generation market segments for whom efficiency and first cost are key requirements. For applications such as this, a single-shaft configuration appeared to be the best approach.

In response, Solar developed the Mercury™ 50 gas turbine, which was announced to the public in 1998 at the ASME Turbomachinery Exhibition in Orlando, Florida. It is an optimized recuperated gas turbine, nominally sized at 4.2 MWe on a 27°C (80°F) day (4.9 MWe at ISO conditions). This environmentally superior system incorporates a highly flexible combustion system that can be configured for either ultra lean-premixed or catalytic combustion, with an ultimate goal of 5 ppmv NO_x on natural gas without the use of add-on, post-combustion pollution control devices. The Mercury™ 50 gas turbine is a single-shaft machine designed around an optimized recuperated cycle that is targeted to meet the rapidly expanding demand for highly efficient, environmentally superior turbine-based power systems in the industrial power generation markets.

MERCURY™ 50 GAS TURBINE DESIGN

Increases in firing temperatures and compression ratios, combined with the use of advanced cooling, sealing and materials technologies have been the traditional approaches to improving turbine performance. While such technology improvements can offer significant benefits, industrial gas turbine users have traditionally avoided new technologies out of concern for the potential impacts on overall reliability. Instead, they prefer a sensible balance between the demonstrated benefits of key technologies and the high reliability requirements demanded by their processes.

This led Solar to evaluate the use of compound cycles based upon demonstrated technologies as the approach most likely to yield a step improvement in efficiency without unduly risking reliability in the process. Initially, a wide variety of candidate cycles was assembled for consideration. The benefits and drawbacks of each concept were qualitatively assessed against key customer buying criteria, including performance, cost, risk, growth capability, reliability and customer acceptance. From this process, two cycles were carried forward as potential solutions: intercooled and recuperated (ICR) and recuperated.

Recuperation of the classic Brayton cycle gas turbine is a well-known method of improving cycle efficiency that involves the addition of a heat exchanger to recover some portion of the exhaust heat that would otherwise be lost to the cycle. The differential between the exhaust gas temperature and the compressor discharge temperature determines the extent of the efficiency improvement that can be realized. Intercooling the Brayton cycle involves the addition of an interstage heat exchanger, to reject a portion of the heat generated during the compression process, which significantly reduces the work input required by the compressor. The combination of intercooling and recuperation into an ICR cycle produces an increase in both specific power and thermal efficiency. This enables the recovery of a greater portion of the thermal energy in the exhaust gas at the expense of an increase in the complexity of the machine. Figure 3.32 graphically contrasts the flow paths associated with simple cycle and recuperated cycles.

Figure 3.32 — Comparison of simple and recuperated cycles.

A more detailed process known as quality functional deployment (QFD) was then used as a formal means of evaluating the technical characteristics of both competing cycles versus the perceived customer requirements. In the final analysis, the recuperated cycle was selected as the most favorable due to the less complicated cycle and lower firing temperature, which improve system availability, emissions and the user's perception of risk. In the future, other efficiency-enhancing cycle modifications, such as inter-cooling, will become more attractive in the marketplace as the benefits, reliability and costs of the more complex systems are incrementally proven relative to the cost of fuel.

Solar's technical approach to ATS emphasizes the use of system-level design solutions that take advantage of a wide variety of demonstrated technological advances, each providing sufficient margin to assure the superior durability and availability required by industrial gas turbine users. Combinations of innovative primary and backup design solutions have been carefully blended to offer maximum cycle efficiency and emissions reductions with minimal risk, as adequate design margin is maintained within each selected technology.

Primary Surface Recuperator. Recuperated cycles have been applied to gas turbines in the past with varied degrees of success. In general, bulky tube-shell or plate-fin heat exchangers were added to existing turbines via somewhat elaborate and often cumbersome piping and support arrangements, with little or no attempt made to optimize the

cycle. The result was a lukewarm performance improvement accompanied by poor thermal transient response, thermal cracking and other mechanical performance problems within the recuperator that often failed to meet customer expectations. Yet, when a prime mover, such as Solar's Mercury™ 50 gas turbine, is designed to incorporate the recuperator from the start, significant gains can be realized without causing the mechanical performance issues that have occurred in the past.

At the heart of the Mercury™ 50 gas turbine lies Solar's proven primary surface recuperator (PSR). As outlined in Figure 3.33, the construction is rugged and the modular nature of the design gives it superior flexibility to handle thermal stresses. Air cells are constructed from 0.1-mm (0.004-in.) thick sheets of Type 347SS folded into a corrugated pattern. This folded shape maximizes the primary surface area that is in direct contact with exhaust gas on one side and compressor discharge air on the other. Pairs of these sheets are welded together around the perimeter to form air cells. Pairs of these sheets are welded together around the perimeter to form air cells.

Figure 3.33 — Construction of the primary surface recuperator (PSR).

There are no internal welds or joints within the air cells, which are the basic building block of the PSR. Groups of these cells are sandwiched together via an arrangement of clamping bars and are welded to intake and discharge headers to form the recuperator

assembly. Developed by engineers at Caterpillar Inc. in the early 1970s for vehicular applications, the PSR has undergone numerous refinements and has been successfully applied to a variety of gas turbines over the years. To date, Solar's PSRs have accumulated in excess of 2.3 million operating hours with few of the incipient problems typical of alternative recuperator technologies.

In contrast, the nature of the PSR presents several major advantages over plate-fin and tube-shell heat exchangers. The PSR is inherently resistant to low cycle fatigue failure. Clamping the cells rather than rigidly welding them to one another allows the assembly to flex freely to relieve stresses, rather than concentrating stresses at the weld locations. Similarly, high cycle fatigue has not been a problem for the PSR due to the clamped design: the stacking of cells presents multiple friction interfaces for energy absorption. This characteristic also provides excellent exhaust sound suppression, eliminating the need for an additional silencing device and its associated pressure drop.

As illustrated in Figure 3.34, PSRs are significantly smaller and lighter than competing technologies, have superior performance, improved reliability and can easily accommodate the thermal transients associated with start-ups, shutdowns and full-load transients during turbine operation. The PSR provides the high effectiveness (>90%), moderate pressure drop and long life demanded by industrial turbine applications.

Figure 3.34 — Comparison of recuperator sizes.

Advanced Component Efficiency (ACE) Compressor. The Mercury™ 50 gas turbine core design incorporates Solar's latest generation of compressors, the ACE compressor. Working in concert with Dr. John Adamczyk from NASA's Lewis Research Center in Dayton, Ohio, this line of highly efficient, rugged compressors has brought the latest

aerodynamic design codes and modeling techniques to the design of industrial turbine compressors. These techniques were first applied during the redesign of the Mars™ 100 compressor in 1993.

The ACE compressor design is characterized by the use of three-dimensional wide chord airfoils that are lightly loaded, resulting in a 40% reduction in the number of airfoils for a given pressure rise. These design characteristics result in a lower cost compressor that offers efficiencies equivalent to today's best aeroderivatives across a wider range of operating conditions. Validation of this design was completed in July 1997 at the U.S. Air Force's Compressor Research Facility at Wright-Patterson Air Force Base in Dayton, Ohio. During the course of this testing, compressor performance was fully mapped and compared versus the design goals, as well as that of Solar's current compressors. The results validated the ACE design codes and yielded an overall efficiency improvement that is more than two points better than the compressors in use on Solar's current line of gas turbines.

The single-shaft Mercury™ 50 gas turbine design uses a 10-stage ACE compressor (Figure 3.35) for a 9.1:1 pressure ratio. A variable inlet guide vane (IGV) is followed by two stages of variable guide vanes (VGV) that are interconnected and controlled as a unit for optimum compressor control across the operating load range.

Figure 3.35 — Advanced component efficiency compressor.

Turbine Design. The design of the Mercury™ 50 turbine section (Figure 3.36) incorporates several advanced features. A two-stage design was chosen for its inherent cost advantages and is characterized by a turbine rotor inlet temperature (TRIT) of 1163°C (2125°F) and a highly loaded, fully cooled first-stage turbine. The second stage incorporates cooled vanes and uncooled, shrouded blades. Although a three-stage design

would appear to offer superior efficiency, the additional air required to cool the added stage effectively negates any performance advantage.

Figure 3.36 — Mercury™ 50 turbine section.

Turbine Cooling. Besides its impact on cycle efficiency, minimizing the use of cooling air has become increasingly important. For example, ultra lean-premixed combustion diverts a larger proportion of compressor discharge air to reduce primary zone temperatures, which makes the remaining cooling air a scarce commodity. Cooling the Mercury™ 50 turbine is an even greater challenge due to the fact that 593°C (1100°F) air from the recuperator is the primary source of cooling. Meeting operating life requirements in this environment has driven advances in both turbine materials and cooling methods. Two advanced technologies have been applied: 1) the refined versions of the film and impingement cooling techniques used on Solar's Taurus™ 70 gas turbine, and 2) a unique leading edge cooling scheme referred to as "vortex cooling."

Vortex cooling development has been under way for several years at Solar and involves the use of swirled cooling flow to the leading edge cooling circuit. The resulting vortices serve to direct the more dense, cooling air to the heat transfer surface and increase turbulence, both of which greatly improve leading edge cooling for a given

amount of cooling air. Vortex cooling offers significant growth potential in terms of cooling effectiveness without incurring the performance penalties associated with showerhead cooling, which would otherwise be required in this high temperature application. This technique and the corresponding results were recently validated on a set of Mars first-stage turbine blades. This basic design, along with further refinements developed during the course of the Mars gas turbine test series, has been incorporated in the Mercury™ 50 first-stage turbine blades.

Materials. The materials used in the manufacture of the turbine section are a key element in the durability of the Mercury™ 50 gas turbine and represent a mix of current materials and next-generation alloys that are relatively new to industrial turbines (Figure 3.37).

Figure 3.37 — Mercury™ 50 turbine materials.

The film-cooled first-stage nozzles are constructed from directionally solidified MAR-M-247, while the uncooled second-stage vanes are constructed from an equiaxed version of the same material. In the case of the unshrouded first-stage blades, a third-generation single-crystal alloy, CMSX-10, has been selected for use. These blades will be mounted to a Waspaloy disk, using a version of Solar's traditional fir-tree design for

blade retention, modified to allow for improved disk-post cooling. The second-stage blades are of a shrouded design and are manufactured from a second-generation single-crystal alloy, CMSX-4.

The shrouded design was chosen due to its superior performance in terms of reduced tip leakage and the improved aerodynamics that derive from the increased aspect ratio of the blades. The uncooled second-stage blades have driven a unique set of requirements for the second-stage disk, for which traditional materials have insufficient mechanical properties. A powdered metal forging of Udimet 720 has been selected for use. The fine grain structure of the rim makes it suitable for the higher rim operating temperatures of the Mercury™ 50 gas turbine while still maintaining sufficient low cycle fatigue strength at the hub.

Rotor Design. A three bearing, single-shaft rotor design has been selected for benefits in dynamic performance and stability during transient operating conditions, such as hot restarts and emergency shutdowns from full load. The bearing arrangement (Figure 3.38) incorporates viscous damped rolling element bearings, in lieu of hydrodynamic bearings, in order to reduce the sizes and, therefore, the cost of the oil sump, pumps, coolers and associated piping. The system employs two roller bearings in the compressor section and a hydraulic damping system designed by Solar for the Mercury™ 50 gas turbine application. The damping system is fed from the bearing oil supply and mitigates dynamic transients during operation. The thrust bearing, a viscous damped ball bearing, is located downstream of the second-stage turbine rotor. The turbine rotors are overhung to minimize the distance between bearings and to allow the thrust bearing to be located in a cooler environment.

Figure 3.38 — Mercury™ 50 rotor and bearing arrangement.

Combustor. There are almost as many approaches to dry emissions control as there are manufacturers, though two technologies have attracted the bulk of the attention: lean-premixed combustion and catalytic, or flameless combustion. The mechanical configurations associated with each are distinctly different and lend themselves to different design approaches within the gas turbine. As will be described later, the design of the Mercury™ 50 combustor will readily accommodate either technology so that it can be tailored to the needs of the particular installation.

Ultra Lean-Premixed Combustion. The ultra lean-premixed design (Figure 3.39 and Figure 3.40) of the Mercury™ 50 gas turbine is a continuation of Solar's successful SoLoNOx™ technology. The design incorporates eight ultra lean-premixed injectors, a backside cooled annular combustor liner and closed-loop CO feedback control to achieve single-digit NOx emissions across the operating range.

Figure 3.39 — Mercury™ 50 combustor.

Figure 3.40 — Mercury™ 50 injectors.

Refinement of the lean-premixed concept developed for the SoLoNOx product line to support single-digit NOx emissions requires the use of additional dilution air through the injector body to reduce primary zone temperatures to 1454°C (2650°F). The use of variable geometry effectively modulates airflow through the injectors as a function of load to maintain consistent primary zone temperatures; however, the pressure drop across the combustor increases significantly at lower loads as excess dilution air is bypassed around the injector. Bleeding air from the combustor case at lower power settings can mitigate this, but at the expense of part-load efficiency. The compensating nature of the variable geometry design is such that the pressure drop across the combustor remains relatively constant, allowing lean operation across a wide load range without the need for compressor bleed at intermediate loads. The injectors are designed to operate in concert with the air diverter valve to bypass excess dilution air to maintain a reasonably constant fuel/air ratio and stable primary zone temperatures necessary for sustained low emissions operation.

Conservation of cooling air and the effects of combustor liner cooling on the formation of CO during the combustion process have brought about a change in the way the Mercury™ 50 combustor liner is cooled. As primary zone temperatures are reduced to support single-digit NOx combustion, CO levels tend to rise, particularly in the vicinity of the relatively cool liner walls, where the CO-to-CO₂ reaction is quenched. Minimizing CO and NOx drives a balance between higher temperatures for reduced CO and lower temperatures for reduced NOx. The alternative liner cooling scheme, referred to as augmented backside cooling (ABC), maintains higher wall temperatures as compared to louver or effusion cooling techniques to further retard the formation of CO during the combustion process. This, in turn, allows the reduced primary zone temperatures necessary for single-digit NOx without sacrificing an increase in CO emissions.

The combustor liner is protected by a ceramic thermal barrier coating (TBC) that is plasma sprayed on the metallic substrate. The addition of dimples and/or trip strips to the back side of the liner serves to break up the boundary layer and promote improved heat transfer. Liners manufactured from continuous fiber-reinforced ceramic composites (CFCC), a design being developed for the CSGT program, are also being considered for use in the Mercury™ 50 gas turbine application as this technology matures. Improvements in mixing capability and minimization of combustor oscillations are key de-

sign advances that have been incorporated into the Mercury™ 50 dual fuel injectors. The use of airfoil main air swirlers with integral gas injection at the upstream edges improves premixing to promote lean combustion while preventing flashback. In the liquid mode, the use of air atomization during lightoff and air cooling of the tip have been added as refinements to the basic design. The incorporation of closed-loop emissions controls will enable the Mercury™ 50 gas turbine to maintain single-digit emissions across a wide range of loads and operating conditions. A CO sensor has been added to the turbine exhaust and is used to drive a proportional control loop that continually adjusts compressor inlet guide vanes, the air diverter valve and the fuel control to accurately control the level of CO and, hence, NOx formation during operation.

Catalytic Combustion System. The design of the catalytic combustion system follows more than four years of development work during which Solar refined and rig tested the basic design of the catalyst beds to prove their suitability for use in a gas turbine application. In a test rig, air and fuel are thoroughly mixed and allowed to partially react in the catalyst bed, where temperatures are kept sufficiently low to avoid damage to the substrate and supporting structure. Burnout is completed downstream of the catalyst bed, where temperatures remain sufficiently low to avoid the formation of NOx.

Initiation of the catalytic reaction normally requires the use of a preburner, which is essentially a small version of a SoLoNOx injector upstream of the catalyst bed. The preburner is used to raise the temperature of the fuel/air mixture to 510 to 538°C (950 to 1000°F), in order to initiate the catalytic reduction of the fuel and air, and is the major source of what little NOx is produced by catalytic combustion systems. One of the many advantages offered by the recuperated cycle of the Mercury™ 50 gas turbine is that it has eliminated the need for preburners and their associated NOx emissions. The combination of variable compressor and combustor geometry act together to allow control of mass flow through the machine as a function of load. Recovery of exhaust heat via the recuperator then allows combustor inlet air to be maintained at approximately 593°C (1100°F) at virtually all operating conditions.

The catalytic combustion system of the Mercury™ 50 gas turbine uses five catalyst cans, approximately 228.6 mm (9 in.) in diameter, and the same compensating geometry as the ultra lean-premixed system to maintain a consistent fuel/air ratio across the 50-to-100% operating range. A cooled can-to-annular transition piece is used to channel

the combustor discharge to the first-stage turbine nozzles. Provisions have been made to accommodate rapid changeout of the catalyst beds as required with a minimum of downtime.

PACKAGE DESIGN

The traditional method of improving gas turbine performance has generally been limited to design changes within the gas turbine itself. These typically have been incremental changes to existing technologies that allow marginal improvements to firing temperatures, pressure ratios or cooling efficiencies. Unfortunately, this approach has failed to capitalize on improvements that can be realized in other areas of the gas turbine system. Specifically, the layout of the turbine in concert with the package has a significant impact on the cost, efficiency, reliability and maintainability of the gas turbine system.

In the case of a typical package, ancillary systems consume nearly 8% of the useful power of the machine and provide an excellent opportunity for improvement (Figure 3.41). As a consequence, Solar's approach to the design of its ATS engine has been heavily focused at the system level. Examples of this have been the use of variable frequency ac drive motors for lube oil pumps and cooling fans, where the power output can be tailored to meet the actual load requirements. Extensive use of advanced analytical design tools, such as computational fluid dynamics (CFD) combined with advanced aero codes and the use of vortex generators in the exhaust casing to improve pressure recovery, have contributed toward significant reductions in exhaust and inlet losses.

Advanced design and modeling techniques have also been applied to the gearbox design. The use of double helical gearing in the epicyclic gearbox, an approach typically reserved for much larger applications, has eliminated the need for thrust bearings and improved component efficiency in the process. Finally, in concert with generator suppliers, Solar is applying technologies typically reserved for use in much larger generators to achieve similar improvements in component efficiencies. The use of profiled conductors for the stator windings and thin-layer laminates made from advanced materials are two key contributors to this achievement.

Figure 3.41 — Parasitic losses in a gas turbine package.

General Layout. One of the most significant Mercury™ 50 gas turbine innovations is the layout of the core prime mover, around which the package has been synergistically designed. The results offer significant advantages in terms of cost, performance and maintainability.

Combustor Flexibility. The need to accommodate multiple combustion systems has been a major factor in the design of the Mercury™ 50 gas turbine. Locating the combustion system between the turbine and compressor, as dictated by past convention, limits the ability to make changes without significantly impacting other subsystems. The centerline design fixes the length of the combustor, and the compressor and turbine cases define the mating interfaces. This has led to the adoption of a unique layout that will accommodate either the ultra lean-premixed or catalytic combustion systems and represents a new innovation in the design of industrial gas turbines. The combustor case and diverter valve assembly are common to both systems, while the end cover, liners, injectors and manifolding are unique to each combustion system.

Simplified Flow Path. The complex flow path of the recuperated cycle also provided design challenges that were effectively addressed in the layout. The recuperator is supplied with air from the compressor discharge plenum on one end, while heated air is collected at the other end and supplied to the combustor. Turbine exhaust is collected

at the bottom center of the recuperator and discharged from the top. On a conventional turbine, this flow path requires a somewhat complex series of piping, elbows, flanges, bellows and pipe hangers. By altering the arrangement of the components, as shown in Figure 3.42, a graceful flow path can be accommodated that naturally follows the flow of the recuperator. Compressor discharge is at the end of the machine in the same plane as the inlet header to the recuperator. The inlet to the combustor is located at the end of the machine in the same plane as the recuperator discharge header. Finally, the turbine exhausts at the center of the gas turbine and discharges up through the recuperator.

Figure 3.42 — Mercury™ 50 core gas turbine and recuperator flow path.

A uniquely designed component referred to as the center frame is used to interconnect the turbine and compressor sections in lieu of the hot strut designs typically used by gas turbines. Since the load carrying members of the center frame are located in a cool environment external to the turbine flow path, axial movement of the turbine case due to thermal growth is minimized, permitting tight tip clearances.

Electric Power Generator. Net shaft power is delivered at the turbine cold-end through a helical-gear gearbox to an air-cooled AC generator. For 60 Hz output, which was the basis for this project, the 14,200 rpm turbine shaft speed is reduced by the gearing to 1,800 rpm, the generator design speed. Mercury 50 AC generators are available that can deliver design output voltages in the 3,300 V – 13,800 V range, the high value being assumed in this study. The study also assumed that Mercury 50 power output would be delivered directly to the utility AC grid; no consideration was given to an overall power management system that would integrate and deliver to the grid the combined power outputs of the PSOFC generator modules and the Mercury 50 gas turbine.

Modular Assembly and Maintenance. Another factor in the design of the package was the desire for modularity in which each of the major subsystems, including combustor, turbine, compressor, recuperator, gearbox and generator can be changed independently in the field in a single work shift. The ability to replace modules, instead of an entire turbine, can also limit the amount of onsite troubleshooting required because the modules can be quickly replaced and repaired off line. The modular nature of the design, as depicted in Figure 3.43, is central to the Mercury™ 50 gas turbine 98.5% availability goal targeted at the beginning of the program.

Figure 3.43 — Modular arrangement of Mercury™ 50 advanced turbine system.

Changeout of modules and major components is accommodated as follows:

- **Liner/Injector Changeout.** Injectors and ignitors can be unbolted and withdrawn from the end cover for individual servicing as required. In addition, the combustor end cover can be unbolted, removed and withdrawn from the case to provide direct access to the combustor liners, injectors, ignitors and manifolding. These components can be serviced individually or replaced as an entire assembly as required.
- **First-Stage Turbine Nozzle.** The entire combustor assembly can be unbolted and withdrawn to provide clear access to the first stage turbine nozzles for replacement without disturbing the remainder of the turbine.
- **Turbine Assembly and Thrust Bearing.** The turbine shaft coupling is unbolted from the compressor via an access port in the center frame. Then, following removal of the combustor assembly as previously described, the turbine assembly is unbolted from the center frame and withdrawn and replaced as a unit.
- **Compressor Assembly.** The turbine shaft coupling is disconnected as previously described and the compressor is unbolted from the center frame. The compressor can then be withdrawn and replaced as a unit.
- **Gearbox and Generator.** Unbolting the couplings for these units allows them to be individually removed from the package and replaced without disturbing the other components.

Multiple Rating Capability. Further operating flexibility has been added through the use of multiple ratings that allow power output and time between overhaul (TBO) to be traded off and tailored to the specific application. Turbine rotor inlet temperature (TRIT) is inversely varied as a function of annual operating hours, thus allowing one size of turbine to serve in a variety of applications for continuous, intermediate and peaking duties. This multiple rating approach also allows the customer to maintain a reasonably consistent TBO schedule on a calendar basis from one application to another.

SUMMARY

The demand for low cost, reliable power has increased dramatically in recent years. With this increased demand, the industry has seen a push toward power generation equipment in smaller sizes in a wide variety of applications. This size range has long been dominated by reciprocating engines, where their low first cost and high efficiency, compared to small industrial turbines, have given them an advantage in applications where footprint, emissions and maintenance are not key issues. Emerging societal trends, however, are pushing toward progressively lower emissions, fuel flexibility and unobtrusive installations in non-industrial locations.

Large domestic reserves of clean-burning natural gas have made this the fuel of choice in many applications. Also, the need to cleanly burn a wide variety of "fuels of opportunity," such as digester gas, landfill gases, and refinery by-products, has placed ever-increasing importance on the design of combustion systems. Together, these marketplace trends are tipping the scales in favor of industrial turbines for many applications. This has led Solar to design the Mercury™ 50 gas turbine from the centerline out to synergistically incorporate efficiency, environmental performance, fuel flexibility, cost of power, reliability and maintainability.

Matching a power generation product that effectively meets the varied goals of a diverse customer base with the emergence of markets such as distributed generation has created excellent growth opportunities for manufacturers of industrial power generation equipment.

The Mercury™ 50 gas turbine has been designed to offer superior performance and operating flexibility at a price that is competitive with alternative power generating technologies. Scaled to different sizes as is currently envisioned, the Mercury™ 50 gas turbine family is capable of satisfying the needs of a wide array of users in a number of different applications. It represents a balanced approach to the trade-off between the benefits of new technology, low cost and high reliability that are targeted directly at the needs of the industrial power generation market.

3.2.3 SOFC Power Conditioning System

The power conditioning system design concept and the material presented in this report section were developed by Mesta Electronics, Inc., North Huntingdon, PA.

3.2.3.1 Introduction

This section describes the design approach, design justification and preliminary packaging for a DC/AC Power Conditioning System (PCS) to be used in conjunction with the PSOFC generator modules. The PCS must be capable of converting the available DC power from the modules to AC power, which must then be phase locked and voltage equalized to the power grid.

To accommodate varied distribution voltages and power requirements, the approach consists of a ($p \times s$) modularized-matrix of inverters. The power conversion equipment will consist of several smaller, high frequency, pulse width modulated (PWM) controlled inverters where the p elements represent the parallel connections and the s elements represent the series connections. The number of parallel elements is dependent upon the power requirements of the system and the current limitations of the switching devices. The number of series elements is dependent upon the open circuit voltage rating of the PSOFC and the collector-emitter direct voltage limitation of the switching device.

The modular-matrix power converter design will be complimented by a highly sophisticated microprocessor based control center design. Utilizing the latest state of the art, high-speed microprocessor with DSP (digital signal processor) complement, the control will integrate the proper A/D interfaces and multiplex the inverter controls. The control center will be programmed to receive information about voltage, current and other pertinent data from each module in the matrix, make the necessary calculations and/or decisions and output the appropriate command to each inverter in the matrix.

The control center will be responsible for the active equalization of all inverter voltages and currents. To ensure proper sharing based on this information, the control will synchronize and phase lock the output to the distribution line frequency and, through a PWM gating waveform, produce a high quality sinusoidal voltage output.

3.2.3.2 Design Concept

General. As illustrated in Figure 3.44, the power conditioning system servicing a single PSOFC generator module consists of several parallel columns of full bridge inverters. Each column, termed herein a Super Module, has one or more Basic Inverter Modules in series, and each inverter drives the primary winding of a transformer with a high frequency PWM waveform. The corresponding secondary of each winding is connected in series and/or parallel to an adjoining module to obtain the desired power level and voltage output. Each inverter/ transformer circuit is a separate DC-to-AC power conditioner. The actual number of DC-to-AC power conditioners will be determined as the voltage and power requirements dictate.

Figure 3.44 — PCS design concept.

This matrix type arrangement of many DC-to-AC power conditioners enables very large power conditioning systems to be constructed from relatively simple smaller conditioners. The modular approach and microprocessor control will permit the construction of a power conditioning and control system for just about any selected input or output voltage. Modular design also provides an inherent system redundancy resulting in increased reliability. Losing a single DC-to-AC power conditioning unit will not result in a complete shutdown of the system but instead will result in slightly reduced performance. Furthermore, using relatively small DC-to-AC power conditioners will permit the use of higher switching frequencies, thus resulting in smaller sized magnetics and filtering networks, which in turn will result in lower system size, weight and cost.

Design Considerations. The two main considerations in determining the proper design approach are the PSOFC ratings and the commercial availability of the switching devices used in the inverter sections. During the design and device selection process, the first order concerns are the open circuit voltage rating and the available full load current. While commercially available isolated gate bi-polar transistors (IGBTs) have voltage ratings up to 3300 V, the devices chosen for this analysis are restricted to 1200V. Presently, IGBTs rated for higher than 1200 Volts exhibit much greater switching losses which inhibits their use at higher switching frequencies and their thermal management becomes more difficult.

The open circuit voltage rating is determined from the V-I characteristic curve for the PSOFC. The end points of the curve define the maximum voltage rating and the maximum DC current available. The maximum RMS ripple current generated by the PCS is to be 2% of the DC current setpoint, for setpoints 370 ADC and above. From the curve, a single PSOFC generator module has the following ratings.

PSOFC Module Terminal Ratings

Output Power	2100 kW
Open Circuit Voltage	2000 VDC (± 1000 VDC)
Nominal Operating Voltage	1400 VDC Terminal –terminal
Full Load Current	1480 ADC
Total Number of PSOFC Modules	4 (per 10 MWe-class power system) 8 (per 20 MWe-class power system)
Module Ripple Current	$\leq 0.5\%$ full-scale for currents $> 25\%$ of module rating

Lower power PSOFC designs are also considered which have a slightly lower open circuit voltage of 800 VDC. This lower open circuit voltage is used to define the optimum choice of a single inverter. To accommodate the 2000 VDC O.C. voltage of the 2100 kW PSOFC Module, three inverters would be arranged in a series string. This would give the inverter a series string with 3600 V withstand capability. Since all the external circuit parameters are not yet defined, this voltage rating would give the inverter string a comfortable safety margin. Table 3.4 summarizes the design approach.

Table 3.4 — Super Module Development

Definition	Open Circuit Voltage	# Inverters in Series	Individual Inverter kW	Total Available kW	Devices
Inverter	800 VDC	1	128.2 kW	128.2 kW	400A/1200V
Super Module	2000 VDC	3	128.2 kW	384.6 kW	400A/1200V

Basic Inverter Module Construction. The most basic component of the power conversion system is the Basic Inverter Module. The Module, configured as indicated in Figure 3.45, consists of the DC link section, full wave power inverters comprised of three inverter poles, the output transformers and control section. The rating for the Basic Module will be 128.2 kW at 480 VAC or 600VAC at the output terminals through the step-up transformer.

The DC link section will be the interface section between the energy converter and the PSOFC source. It will consist of the input DC breaker, branch circuit dc fuses, pre-charge network and ancillary control circuitry, and it will be the point of coupling to the PSOFC module.

The Basic Inverter Module will be configured as a full wave bridge, which will convert the DC bus voltage to a PWM controlled, 3Ø, output voltage. The PSOFC will be buffered from the DC bus by electrolytic capacitors, which will provide transient load ride-through capability to the inverter.

Figure 3.45 — Basic inverter module.

The PCS will be designed utilizing the Basic Inverter Module as a fundamental building block that can be configured to meet any voltage and power requirements. When three Basic Inverter Modules are connected in a series bridge arrangement, the higher input voltage requirements are satisfied. This grouping is referred to as a Super Module, Figure 3.46.

Figure 3.46 — Super module.

The Super Module is connected to a PSOFC module with an open circuit voltage rating of 2000 VDC at a maximum current of 1480 ADC. Each inverter in the Basic Inverter Module drives the primary winding of a step up, isolation transformer by way of a PWM waveform. The corresponding secondary of each winding is cascaded in series or parallel to obtain the desired output voltage. Each inverter in the Super Module is a current source/voltage source inverter ensuring proper load sharing between inverter sections in the total matrix. In turn, the Super Modules are connected in series/ parallel cascades to match the load voltage and power requirements as shown in Figure 3.47.

Figure 3.47 — One-line diagram for power conditioning system associated with four PSOFC modules.

Power Conditioning System for 7.9 MW AC. As shown in Figure 3.47, to facilitate a medium or high-voltage line voltage, multiple Super Modules are cascaded by connecting the secondary windings of the inverter transformers in a series connection. The three-phase connection is arranged such that the delivered line to line voltage can be as high as 13.8 kV. The deliverable power to the grid from four PSOFC modules is characterized as follows:

Output Power	7.9 MW
Line to Line Voltage(V_{L-L})	13800 VAC
Line Current (I_L)	321AAC

Power Conditioning System for 15.8 MW. To consolidate power from eight PSOFC generator modules, and deliver that power to the AC grid, two systems of the type described above are paralleled as indicated in Figure 3.48. There is no change system voltage level, but the power delivered increases by a factor of two due to the increased current capability.

Figure 3.48 — One-line diagram for power conditioning system associated with eight PSOFC modules.

Representation Only
Not Meant as Schematically Correct

High Voltage Transition Section. The connection to the medium or high voltage power grid should be made through a High Voltage Transition Section. This cabinetry will house the connection panels for the PCS output voltages. At this juncture, secondary windings will be connected in parallel or series per system voltage requirements. In addition, high voltage, fused potential transformers (PTs) will be provided to step the high voltage down to control circuit level voltages. Current transformers (CTs) will be provided to monitor line currents.

High voltage 15 kV-class disconnect switches will be used to provide interlocked isolation to the Super Modules for routine maintenance. These switches will be fused to protect the system from load fault currents. The switches should have a visible indication of the switch position status. Operator safety interlocks such as lockouts and key interlock provisions should be included. Control circuitry will be used to interlock doors, offer over-current protection and provide personnel safety lock-outs. These switches are commercially available.

3.2.3.3 Technical Discussion

Technical Issues & Problems. A number of design approaches were considered. The first approach considered was to provide multiple directly paralleled switching devices to provide the necessary primary current to an isolation transformer, which would step the voltage up to the distribution voltage level. This approach was dismissed since paralleling multiple devices would require extensive snubber circuits to counter switching voltage transients. Another issue is thermal management. Water-cooling would be required to provide enough cooling medium to keep the physical size to a minimum. This approach would not be flexible in its adaptability to other specifications. The decision was made to investigate a modular approach. Iterations were run to optimize the modules for maximum efficiency and cost.

Device Selection. The evaluation is based on commercially available IGBTs and the heatsink of choice, which is forced air-cooled. Dual IGBT modules were chosen as the device used in the design. These modules would be used as the switching device for $\frac{1}{2}$ of the inverter pole. The dual module will minimize switching transients. The maximum nameplate current rating for these devices is 300 or 400 amperes. Present voltage ratings are 600, 1200 and 1700 Volts, collector to emitter (V_{ces}). To maintain safe

design levels under normal switching events, maximum peak voltage transients must be kept at $< 90\%$ of the V_{ces} rating.

The first calculations were run using 1700 V IGBT modules. Two devices in series would be required to withstand the open circuit voltage requirement. These devices were not chosen because the switching losses at higher frequencies were too great. To lower the switching frequency would increase the cost of the filtering circuits, magnet components and DC bus capacitors.

600 V IGBT modules were evaluated. Four devices in series would give a 2400 V withstand. This would push the design to above the 90% limit but may still be acceptable. However, the cost differential between the 600 and 1200 V modules does not offset the increased number of modules required for the same current density.

The 300 A rated devices were evaluated. As shown in Table 3.5, which summarizes the results of the device evaluation, the thermal management looks good but it will take 1/5 as many devices to achieve the same current output.

The device number in Table 3.5 refers to a manufacturer or specific device. The different collector currents (I_c) reflect the current rating of the device. The Primary Line Current value is the required inverter current supplied to the isolation transformer. The Number of Parallel Inverters refers to the minimum number of inverters determined by the current density limitation of the device. Device 4 would be chosen based on the losses and current rating.

Table 3.5 — IGBT Module Evaluation

IGBT Module Evaluation						
	1200V				600V	
Device	1	2	3	4	5	6
Namplate Current Rating (Ic)	300	300	400	400	300	400
Primary Line Current	232.7	232.7	232.7	232.7	232.7	227
Number of Parallel Inverters	5	5	4	4	5	4
Vce sat(volts)	2.7	2.1	2.4	2.15	1.6	1.5
Turn-On Energy (mWs)	21	19	24	23	7	10
Turn-Off Energy (mWs)	21	23	29	28	7	19
Total Switching Energy (mWs)	42	42	53	51	14	29
Total Switching Loss @fc(W)	420	420	530	510	140	290
Conduction Loss/IGBT (W)	209	162.3	223.5	200	124	139.5
Diode Conduction Loss (W)	120	120	153	158	87	107
Diode RR Loss (W)	112	86	115	110	33	25
Total IGBT Loss (W)	629	582.3	753.5	710	264	429.5
Total Diode Loss (W)	232	206	268	268	120	132
Total Loss per Phase (W)	861	788.3	1021.5	978	384	561.5
Total H/S Losses -3 Ph.(W)	2583	2364.9	3064.5	2934	1152	1684.5
Delta Tj-c/IGBT (degrees C)	23.5	21.8	18.8	17.75	11.88	19.3
Delta Tj-c/Diode (degrees C)	20.88	18.54	16.75	16.75	15	9.9
Delta Tc-s/module (degrees C)	16.35	14.97	19.41	18.6	7.29	10.66
Tjunction @Ta=40 degrees C(C)	135.73	130.31	129.96	128.1	109.17	114.86

The proposed modular/matrix approach is flexible in its application to various distribution voltages and power demands. By using proven designs in a matrix arrangement, reliability and cost effectiveness are maximized for the power density desired.

Inverter Rating. The inverter must be sized according to the power available and the output requirements. Referring to Figure 3.49, which shows a single phase representation of a 7.9 MW section of the total power system, and assuming the output

Figure 3.49 — Power conditioning concept for four PSOFC modules.

windings are connected in a wye, then the line to neutral voltage (V_{L-N}) is

$$V_{L-N} = \frac{13,800V}{\sqrt{3}} = 7967 \text{ Volts.}$$

Each of the DC/AC converter sections must support $\frac{1}{4}$ of this voltage or 1991V at 232.7AAC. Since three inverter sections are connected in a series configuration each section will have available $\frac{1}{3}$ of the working voltage of the PSOFC or 467 VDC. Interpretation of the PSOFC characteristic voltage/load curve indicates that adjustments in power component sizing must be made to accommodate a lower DC supply voltage and stray voltage drops must be taken into account. Assuming a worst-case insertion loss of 17 V, and utilizing a PWM voltage waveform, causes a reduction in the effective RMS voltage delivered to the transformer primary. The resultant voltage is 318 V RMS. The deliverable power from the inverter is 128.2 kW.

The following illustration demonstrates the design flow and justification.

$P_{\text{available}} = 2100 \text{ kW}$	Available power from PSOFC
$V_{\text{bus-min}} = 1400 \text{ VDC}$	Minimum DC operating voltage from PSOFC.
$V_{\text{bus}} = V_{\text{bus-min}} / \# \text{ of series inverters}$	Available bus voltage per series inverter.
V_{drop}	Circuit voltage drops ($\approx 17 \text{ V}$)
$V_{\text{net}} = V_{\text{bus}} - V_{\text{drop}}$	Minimum peak switching voltage after circuit voltage drops.
$V_{\text{RMS}} = V_{\text{net}} \times 0.707$	Effective RMS voltage delivered using PWM control.
$I_{\text{inv}} = 232.7 \text{ A}_{\text{RMS}}$	Maximum RMS current delivered to the transformer primary.
InvkW	Maximum power delivered by inverter.
Power	Inverter power times the number of inverters (15) per Super Module.

$$V_{\text{bus}} = \frac{1400}{3} = 467$$

$$V_{\text{net}} = 467 - 17 = 450$$

$$V_{\text{RMS}} = 0.707 \times 450 = 318 V_{\text{RMS}}$$

$$\text{InvkW} = V_{\text{RMS}} \times I_{\text{inv}} \times \sqrt{3} = 128.2 \text{ kW}$$

$$\text{Power} = 15 \times 128.2 \text{ kW} = 1923 \text{ kW}$$

Snubbers. One of the most common problems encountered in high frequency, high-current-density inverter designs is the voltage control during turn-off and turn-on of the switching devices. Trapped energy in the circuit stray inductance manifests itself as damaging overshoot voltages, which can easily result in failed devices. One of the design decisions on how to minimize and possibly eliminate these voltage overshoots is to closely couple the DC bus to the inverter pole. To accomplish this, dual inverter modules are used. In addition, local DC bus capacitors will be used to supply load matched storage elements to each inverter section. Controlling the gate firing pulses may also aid in controlling the overvoltages by softer turn-on or turn-off gating characteristics. Should snubbers be necessary to control overshoots in the voltage, the snubbers will be an energy recovery type, which will put the energy back into the system actively rather than just dissipate the energy as losses.

3.2.3.4 Performance Characteristics

The efficiency of the power conversion equipment is dependent upon two main components, the power switching devices and the magnets (both isolation transformers and filter inductors). As has been previously discussed, the IGBTs are chosen to optimize minimal losses. The transformers and inductors will be designed with special attention given to not only the fundamental frequency losses but the high frequency losses as well. Winding materials and techniques will be employed to minimize the DC and high frequency I^2R losses. Core loss due to eddy currents (fringing) and hysteresis will be minimized by; 1) using proper core materials and 2) designing flux densities consistent with the efficiency goal.

Loss Estimates

Transformer Loss. Regardless of the transformer configuration (one large transformer driven by multiple inverters or each inverter having a dedicated transformer), the losses would remain a percentage of the overall kVA of the magnetics. The transformers will be designed with maximum efficiency and losses restricted to no more than 2%.

IGBT Loss. IGBT loss is a function of the collector current, switching frequency and turn-on/ turn-off energy. Approximately 2800 W are dissipated per inverter as per Table 2. This loss would result in a total of 2.3% IGBT losses for the system.

Additional Losses. Ancillary losses are estimated at 0.5%. These losses include but are not restricted to losses in the DC bus capacitors, control circuitry and wiring.

In conclusion, the total loss estimate is $\approx 5\%$, corresponding to a PCS efficiency of approximately 95%. It is considered that this value appropriately represents the efficiency with which the PCS converts DC power at the PSOFC module terminals to grid-quality AC power at the grid interface.

Power Factor

The power factor will be 0.90 or better from 10 to 100% power.

3.2.3.5 Cost Estimate

System Costs. The cost estimates for the overall system design are based on the inverter capacity rather than the actual delivered power. Since the inverter is designed to open circuit voltage specifications and delivered current, the inverter load line is defined as 2.96 MVA (2000 VDC, 1480 ADC). Utilizing limited quantity price breaks for 120 inverters, the cost is estimated at \$0.168/W or KVA based on PCS capacity. Relating this price to the delivered power is a direct ratio of the design capacity to delivered power. This ratio would put the cost of delivered power at approximately \$0.24/W delivered power to the grid.

Increased sales volume would have an impact in the material cost of about a maximum of 20-30% savings. Following typical business profit/cost ratios, this reduction in material cost would impact sales pricing by approximately 10% since the material cost accounts for about 30-32% of the sales price. Additional savings may be realized by reduced labor costs as efficiency increases with familiarity based on volume. This reduction would impact labor content, which would generally account for 10-15% of the sales price. A reasonable reduction in labor costs could account for 10% reduction in man-hours. A 20% reduction, representing volume effects on materials and labor, was applied in developing an appropriate commercial inverter cost estimate. The resulting inverter cost estimate is \$397,800 per PSOFc generator module.

Transformer Costs. The magnetic costs are dependent on the design approach taken. The least expensive approach would use a high kVA, utility type transformer configuration. Typical distribution configurations would use 1 ϕ transformers connected in a three phase arrangement. These transformers would be "off the shelf" items and would reflect a commodity pricing structure. Utility transformers are low cost items with prices in the \$5-6/kVA range. This magnet approach is not practical for an inverter matrix approach because the primary currents would be in the kiloampere range for these power levels. This would require direct paralleling of the IGBTs, and since IGBT ratings are limited, the control configuration becomes extremely difficult to manage. The multiple transformer approach as discussed makes the control more manageable and economic, but the transformer costs increase to about \$22/kVA.

An avenue that could be investigated would involve a consolidation of the magnetics by having one transformer core with multiple isolated primary windings. The cost would

be impacted by a reduction in overall weight (less core material than the multiple, single transformers), but the labor costs increase because the process is no longer a continuous winding design (stop and start). Each PSOFC module group of four modules could have one transformer with multiple windings rated for 2.1 MVA. This would drive the transformer costs down. The actual savings would have to be determined at time of order placement.

Designs could be based on copper or aluminum windings. The cooling specification could be for convection, forced air-cooling or oil filled. This price will vary as the market varies with copper and aluminum raw material market value. Quantity buys of a similar design could have a significant impact on this figure. The modular approach proposed would facilitate this quantity pricing structure for the transformers.

At \$22/kVA, the estimated transformer cost per PSOFC module is \$46,200.

High Voltage Section. The 15 kV-class disconnect switches needed to isolate the PCS from the utility grid are commercially available from a variety of manufacturers. List pricing is around \$5000. Options will increase this price. The options available include, but are not limited to, motor operated mechanisms, fuse failure indication and customer interface contacts.

The high voltage section should be outsourced and the pricing would reflect prevailing prices.

On the basis that each 10 MWe-class power system would require two disconnect switches, and allowing for options, the estimated switch cost is \$5000 per PSOFC module. The total estimated PCS and disconnect-switch cost is therefore \$449,000 per PSOFC module, which corresponds to approximately \$215/DC kW.

3.2.3.6 Physical Dimensions

The inverters will be housed in modified NEMA 12 enclosures as indicated in Figure 3.50. The enclosure dimensions for one Super Module would be approximately 2.3m (h), 1m (w), and 0.9m (d) [7.5 ft (h), 3.3 ft (w), and 3 ft (d)].

Figure 3.50 — Inverter cabinet concept.

Four cabinets would be required by each PSOFC generator module. Three would house four inverters each, as shown in Figure 3.50, while the fourth cabinet would contain three inverters (inset). The PCS cabinetry arrangement concept for four PSOFC modules (one PSOFC/ATS-GT power block) is depicted in Figure 3.51.

These drawings are for concept depiction only, and they indicate the cabinet volume to house the inverter components. During the detail design, every effort should be made to optimize the physical arrangement of the inverters so that the connection points to the PSOFC provide for the shortest bus distances possible. This would minimize DC losses and lessen material and installation costs.

Figure 3.51 — PCS cabinet concept for four PSOFC generator modules.

3.2.4 Instrumentation and Controls System

The PSOFC modules and auxiliary equipment will be equipped with instrumentation and controls that provide for automatic operational control, and with manual capabilities, for plant operation. The instrumentation and controls (I&C) system will also provide for remote monitoring and operations, data collection, and diagnostics.

3.2.4.1 Hardware

PLC. The heart of the instrumentation and controls system is the programmable logic controller (PLC). The PLC provides for safe, reliable, and steady power system operation over its operating range. The controllers will scan inputs, execute preprogrammed logic, and set the state of outputs based on the control strategy.

Input/Output Modules. The I/O modules will link the instrumentation signals to the PLC. Thermocouples, voltage sensors, and current sensors will provide raw analog data through the I/O modules. The PLC converts the signals to digitized engineering units for display, and it provides for an appropriate output response to various system actuators.

Communication Modules. The communication modules will link the PLC to one or more PSOFC Control/Data computers. In addition, the communication modules will provide the control signals to the PCS for power control.

Electrical Control Hardware. The electrical control hardware is composed of those components that gather the input and output signals and respond to the needs of the main PLC process controller. Power supplies, signal conditioners, relays, current and voltage sensors, etc., comprise the electrical control hardware. These components form the control circuits that implement the commands from the PLC. The electrical control hardware also provides for continuous control of auxiliary devices such as blowers, valves, and the PSOFC generator/gas turbine electrical loading. The components will be packaged for efficient space utilization and convenient maintenance access.

Operator Interface. An operator interface computer will be supplied as part of the I&C system, to provide a display of the critical PSOFC generator parameters, and to allow for operator input of control parameter setpoints. The operator interface will consist of a computer, graphics display monitor, large-capacity hard drive, and printer. Provisions

for alarm monitoring/logging, operator input for state transition, manipulation of remote alarm action and control parameter setpoints will be incorporated in the operator interface.

3.2.4.2 Software

The I&C system is a PLC based system that stores and uses information about the status of the PSOFC generator modules, the gas turbine systems, and the support systems. Data are read continuously and updated by the control system. Some data are also directed to, or taken from, the operator interface. The system software specifies the response to data, and controls the state of the generator. The system detects alarm conditions, and controls both modulating and discrete devices.

The power system will have a table of pre-defined operating states. For example, Pre-Op, Purge, Heat, Load, Run, Cool, Stop, and SStop are typical PSOFC system states. The control system software contains the logic that determines the operating state of the PSOFC generators and all valid transition states. The control system will allow the operator to transition from the present state to a valid transition state only.

Discrete control logic will be used to change the states of control solenoid valves, and to provide valve status information to the operator.

Analog control logic will be set to detect out-of-limit alarm signals. Response to an alarm signal will depend on the severity of the alarm. All alarms warn the operator. Others may result in a shutdown of the system.

Operator interface software package will be provided for supervisory control and data acquisition.

3.2.5 Electrical Distribution System (EDS)

The PSOFC electrical distribution system (EDS) refers to the distribution of AC power to the instrumentation and control systems, and to any support system loads. A one line diagram of the EDS conceptual design for one PSOFC/ATS-GT power block is shown in Figure 3.52. It does not include the PSOFC generator DC output power, the AC power

out of the inverter, or the gas turbine AC output power. The requirements of the EDS system are to provide 480 VAC, three phase power, 208 VAC three phase power, and 120 VAC single phase power to the PSOFC generator auxiliary system components. The power source of the EDS main circuits and branch circuits, is a step down transformer that is tied to the primary high voltage AC bus. The transformer lowers the bus voltage from the grid 13.8 kV to 480 VAC, three phase, for distribution and use within the I&C systems of the generating plant. A further reduction from 480 VAC is required to feed the 208 VAC, three phase loads, and the 120 VAC, single phase loads.

The 480 VAC bus may be required to supply three phase power for the heavier PSOFC loads, including pumps, compressors, and other internal service loads and house loads. The 480 VAC system feeds will have manual disconnects for maintenance, and circuit breakers for electrical protection in the event of a system fault.

A 208 VAC and 120 VAC distribution panel will be provided for the distribution of power to heaters, steamers, boilers, dc power supplies, solenoid valves, and all other low voltage components required to monitor and control the PSOFC generator system operation. The panel will be supplied with, and located within, the PSOFC system to distribute power to the loads. In addition, 208 VAC, three phase, and 120 VAC, single phase back-up power will be made available for critical PSOFC system loads in the event of a grid outage. Should an outage occur, the non-essential loads will be disconnected from the 480 VAC bus, and only those required loads will be fed from the back-up supply. The back-up supply will continue to power the essential components until grid power is restored, or an orderly shut-down of the PSOFC system occurs.

3.2.6 Fuel Processing System

The supply of natural gas is assumed to be within the conventions of the U.S. distribution system. In the U.S., the distribution mains range in size from 0.508 m. (20 in.) to 1.07 m. (42 in.), and contain gas at 13.8 barg (200 psig) to 103.4 barg (1500 psig). Compressor stations are normally located 80 km (50 miles) to 100 km (60 miles) apart and occasionally as far as 300 km (180 miles). Gas flows in the mains at 24 km/hr (15 mph or 22 fps). The natural gas is reduced in pressure at the local natural gas supply utilities. These utilities add an odorant and distribute the gas through lines at about 4.1 barg (60 psig). The pressure at residential customers varies from 17 mbarg to 4 barg

(0.25 psig to 60 psig). The most commonly used gas piping is rated to operate at 6.9 barg (100 psig). For this study, it was assumed that the supply of natural gas at the plant site is at 1.03 barg (15 psig) and contains a sulfur-based odorant. The gas line size needed for a 20 MW plant is 30.5 cm (12 in.) at 1.03 barg (15 psig) supply pressure.

On the plant site, the natural gas is filtered in an automatic duplex inlet filter (see Figure 3.10). One side of the filter is on standby while the other side performs the filtering. When the differential pressure across the filter reaches the set pressure corresponding to a dirty filter element which needs to be changed, the automatic duplex filter uses the gas supply pressure for the motive force to actuate the three way valves and divert the flow through the standby side. A dirty filter indication is sent to the control system and filter element changing is scheduled. Filter element changing is performed with the unit on line and fuel flow is not interrupted.

The natural gas enters the inlet of one 100% capacity natural gas compressor. The compressor discharges to an accumulator which provides a pulsation dampening function as well as a reservoir of natural gas at compressor discharge pressure to ride through supply interruptions when the compressor stops and goes off line when the accumulator high pressure set point is attained. The compressor operates on accumulator pressure control, starting on low pressure and stopping on high pressure. The compressor design flow rate is set slightly higher than the maximum plant fuel consumption rate. During normal plant operation, the compressor will be idle about one fourth of the time.

From the natural gas accumulator, the fuel passes through a gas recuperator. Hot natural gas from the outlet of the desulfurizer system exchanges heat with the cooler natural gas leaving the natural gas accumulator. This preheated natural gas then leaves the gas recuperator at a temperature of 375°C (700°F).

The natural gas desulfurizer includes one hydrodesulfurization catalyst vessel in series with two series-mounted, 100% capacity, adsorbent (zinc oxide) vessels. At the entrance to the catalyst vessel, a very small quantity of hydrogen, about 1% (by volume) of the fuel flow, is mixed with the natural gas. The mixture is electrically heated to raise its temperature to near 400°C (750°F). Each adsorbent vessel includes an electrical gas heater under automatic temperature control to ensure that the zinc oxide bed and natural gas are at suitable temperature for effective sulfur adsorption.

The catalyst is cobalt-molybdenum, which promotes the conversion of complex sulfur compounds to H₂S. The gas then flows to the adsorbent vessels. Zinc oxide is ineffective for H₂S adsorption at ambient temperatures, but it is very effective at temperatures above 277°C (530°F). Preheating the natural gas/hydrogen mixture in the gas recuperator, followed by supplemental electric heating to the optimum operating temperature of 400°C (750°F), permits the use of minimum sized adsorbent beds.

Sulfur removal effectiveness is checked monthly with the drawing of a sample from the line between the adsorbent vessels. When the sample shows detectable sulfur, the valve alignment is set to enable the on-line replacement of the spent upstream adsorbent and its return to service as the downstream adsorbent.

3.2.7 Gas Supply Systems

The power system includes both nitrogen and hydrogen generating equipment. The nitrogen is stored on-site in pressurized cylinders until needed. The hydrogen is produced as demanded and not stored, thereby reducing fire protection concerns.

3.2.7.1 Nitrogen Supply System

The inlet to the natural gas compressor accumulator has a connection to a nitrogen purge system (see Figure 3.10). The nitrogen supply system is normally isolated from the natural gas supply and is activated under two conditions:

- Emergency or safety stop (SSTOP)
- Maintenance of natural gas system components

When activated, the nitrogen purge system connection isolation valve and gas vent valve both automatically open. Nitrogen gas from a pressurized tank flows through the accumulator, gas recuperator, desulfurizer vessels and outlet duplex filter to the vent. This dilutes and forces the natural gas out of the system until a nonflammable concentration of gas remains.

SOFC Cover Gas. The SOFC stacks require an inert or reducing environment on the cell fuel side when the stack temperature approaches and exceeds 600°C (1100°F).

During normal operation, this environment is provided by the fuel. When fuel is not present, during startup and shutdown operations, nitrogen cover gas provides the required inert atmosphere. A commercially available nitrogen generating system produces this cover gas. This system uses a selective membrane to separate nitrogen from the other constituents of air.

A pre-packaged, skid mounted nitrogen generation system produces nitrogen gas whenever demanded by the low-pressure control in the nitrogen storage accumulator. The system stops when high accumulator pressure is achieved. The nitrogen generating equipment may occasionally operate to refill the accumulator if there are leaks. The nitrogen system is sized to produce in seven days a sufficient quantity of nitrogen to support one plant shut down. The storage accumulator has provisions to be filled with nitrogen from an external source, if necessary.

Nitrogen generation uses commercially available equipment that consumes electricity and compressed air to generate nitrogen gas. The nitrogen is separated from the compressed air through a selective membrane. The nitrogen is 98% pure. Nitrogen flows under its own pressure from the accumulator without the need for additional pumping. As the nitrogen is generated, hydrogen (about 4% by volume) is added to scavenge impurity oxygen.

3.2.7.2 Hydrogen Supply System

A pre-packaged, skid-mounted hydrogen generation system produces hydrogen continuously during plant start and normal operation. During normal operation, the hydrogen is mixed with the natural gas to form a gas/hydrogen mixture with approximately 1% (vol) hydrogen. This mixture is heated and delivered to the cobalt-molybdenum catalyst in the fuel desulfurization system.

The hydrogen generation process uses commercially available equipment that consumes electricity and electrolyzes water to generate the hydrogen gas at the maximum rate required during power system operation. This gas is produced as needed, and is compressed and stored as it flows from the electrolysis chamber.

3.2.8 Balance of Plant (BOP) Equipment

3.2.8.1 Startup Boiler

A natural gas heated boiler is supplied to provide a source of steam for plant start up. This commercially available, prepackaged, skid mounted boiler is a complete self-contained system with all auxiliary components and instrumentation. Integration with the SOFC power system requires connections with natural gas, demineralized water, and electric power supplies. The steam outlet is piped to the SOFC generator modules, and the boiler blowdown line is piped to a drain.

Boiler operation is manually controlled for the short time that it is needed. The boiler is used only for plant startup, and it is drained and laid up dry for prolonged periods of inactivity during normal power system operations.

3.2.8.2 Auxiliary Air Compressor

The auxiliary air compressor system includes an internal combustion engine driven compressor, inlet air filter, compressed air filter/dryer and air accumulator plus the necessary engine starter, instrumentation and controls. The dual-fuel engine can use either natural gas or propane gas. The primary engine fuel is the normal natural gas supply. In the event of loss of natural gas supply, the engine automatically switches over to propane, obtaining that fuel from a pressurized, on-site, liquid-propane storage vessel.

The compressor operates to fill an air accumulator. The compressor engine starts and stops based upon pressure control signals from the accumulator. The auxiliary air compressor is sized to provide the airflow to cool the cell stacks during a shutdown operation. Stack cooldown to less than 260°C (500°F) occurs in about three hours.

The compressed air accumulator provides a pulsation dampening function at the compressor outlet and a steady flow of auxiliary air when engine operation pauses momentarily during automatic fuel supply changeover.

3.2.8.3 Water Treatment and Storage

Potable water is taken from the site municipal supply, filtered, deionized, and stored. The storage volume is sized to provide the water needed by the hydrogen generating

system and startup boiler system during a normal plant start. The storage tank level is automatically maintained by a level-sensing supply-water control valve.

The water system includes a pump to transfer deionized water from the storage tank to the hydrogen generating system and startup boiler.

3.2.8.4 Process Air Piping

The design of the process air and gas piping to be installed between the gas turbine and the PSOFC generator modules must satisfy several criteria:

- The piping wall temperature should be less than or equal to 371°C (700°F). This criterion allows the use of standard piping material instead of higher cost alloys rated for high temperatures. All standard piping steels suffer no loss of strength until the working temperature exceeds 400°C (750°F).
- The external insulation jacket surface temperature must be at or less than the OSHA 29CFR1910.107(l)(7) limit of 65.6°C (150°F). Conventional jacketed, mineral wool is suitable for this application.
- The speed of the air inside the inner insulation is less than 30 m/s (100 ft/s). This air speed does not require an internal protective liner for the insulation and permits the use of less expensive, commonly available insulating materials without concern for erosion of the surface. Vacuum formed alumina-silica fiber insulation was found to have good thermal resistance and appropriate abrasion resistance.

Applying these criteria, the nominal diameter of the piping selected for the SOFC air inlet and exhaust outlet headers is 97 cm (38 in.), and the diameter of the branch pipes that connect the headers and the SOFC modules is 71 cm (28 in.). For both pipes, the internal-insulation thickness is 19.4 cm (7.63 in.), and a thickness of 7.6 cm (3 in.) was selected for the external insulation.

3.2.9 Plant Operation

Plant operations are summarized in the following sections, including plant startup, operation, normal shutdown, and emergency shutdown.

3.2.9.1 Preparations for Normal System Start

The following services are available:

- Natural gas compressor is operational and gas supply pressure is in range
- Auxiliary air compressor is operational and compressed air accumulator pressure is in range
- Demineralized water storage tank is filled to required level with deionized water
- Nitrogen system accumulator pressure is in range
- Hydrogen system is operational

System Configuration:

- Initial system status — unpressurized at ambient temperature
- Fuel piping — fuel supply control valves for all SOFC modules are closed. Fuel supply control valves for both air heaters are closed. Combustor fuel supply is off. Desulfurizer vent to atmosphere is closed.
- Vessel fuel vent system piping — vessel fuel vent system isolation valves from each SOFC module is closed.
- Auxiliary air compressor piping — Isolation valves to SOFC module air supply headers are closed.
- Cover gas piping — Isolation valves to SOFC module fuel supply headers are closed.

Operator Actions Sequence. Start turbine air compressor motor. When airflow through turbine is established, fire the combustor to bring the turbine to normal speed. Allow the turbine/generator to free wheel as the flow of air passes through the SOFC vessels to the turbine inlet.

Fire the air heater on the SOFC vessel inlet header. Control firing to limit the air heater exit air temperature to 500°C above the lowest temperature SOFC vessel stack.

When the highest temperature SOFC stack reaches 500°C, open the cover gas system supply isolation valves to admit inert gas to the header that feeds the SOFC modules.

Begin adding steam to the SOFC generator modules, via the fuel supply system, when combustion zone temperatures reach 600°C.

When the SOFC vessels reach 675°C stack temperature, begin to draw current. Start low fuel flow. Slowly increase the power being drawn from the stack. Fuel flow increase will automatically follow current increase.

Isolate cover gas system from the SOFC vessels headers when the current from each stack reaches 300 A. Fuel reformation and recirculation is sufficient to maintain a self-sustaining reducing atmosphere in the fuel region.

Begin to load the turbine generator. Increase the power drawn as the temperature increases in order to maintain airflow (shaft speed) at design value(s).

Turn off air heaters when the current from each stack is approximately 460 A. At this time the recuperated air temperature is self-sustaining.

3.2.9.2 Normal Operation

The plant run state is entered when the stack temperature achieves 850°C and the current draw is 300 A from each stack. During normal operation, the operator has control of:

- Fuel utilization
- Current
- Stack temperature

Setting the control system set points for any two of these parameters determines the third.

The plant is expected to operate unattended for 50 weeks with periodic remote monitoring and control adjustments made via telephone modem access for a remote location. At the end of the scheduled normal time, operators return to the site to supervise a normal shutdown.

While most maintenance will be performed during the annual scheduled plant outage, some activities may need to be performed during normal operation. The replacement of natural gas desulfurizer adsorbent will normally be performed during the annual maintenance period. On-line absorbent changeout is done as discussed earlier. The duplex filters on the inlet and outlet of the desulfurizers will normally last for years be-

fore developing enough filtrate cake to produce the pressure drop to trigger automatic change over. The duplex filter changes position to pass flow through the standby filter when across the filter media increases to the dirty filter set point. Spent filter elements will be changed as the need arises.

3.2.9.3 Normal Shutdown

Start Cover Gas System. When cover gas system is fully operational, the turbine AC generator is unloaded.

Shutting down the SOFC stacks is accomplished with simultaneous control actions to:

- Stop natural gas flow to the stacks
- Start cover gas system flow to the stacks
- Unload current to ~35 A per stack and transfer power to stack energy dissipaters

The turbine will continue to rotate and produce compressed air flow, running on the stored heat from the stacks. After system cooling has proceeded and the turbine/compressor approaches stall, the generator will be used as a motor to drive the compressor.

Stop flow of cover gas system when stack temperatures are less than 500°C.

Continue the supply of auxiliary air until stack temperatures are less than 100°C.

Automatic Stop Conditions. The plant will automatically enter STOP when any of the following conditions are sensed:

- Low generator voltage
- Low generator current
- High stack temperature for 10 minutes
- Low UPS battery voltage

The sequence of automatic shutdown actions is the same as if the shutdown were initiated by manual control action.

3.2.9.4 Normal Maintenance

At the end of a normal run time, the plant is shut down for normal maintenance. The maintenance period is two weeks. During this period the activities include:

- inspection of turbine and generator rotating parts
- lubricating oil changes
- air filter element replacement
- instrumentation and control system calibrations
- repair or replacement of failed or improperly operating components
- fire protection system inspection and testing

3.2.9.5 Emergency Operations

To accommodate the occurrence of abnormal situations, the plant control system is equipped to automatically take control actions. The priorities for automatic emergency control actions are 1) personnel protection and 2) property protection. The extent of emergency conditions that are accommodated are shown in Table 3.6.

It is assumed that no two failures occur simultaneously.

Table 3.6 — Plant Emergency Situations

Plant Mode	Control Sequence			
	Loss of Grid	SOFC Failure	Turbine or Generator Failure	Cover Gas System Failure
Startup	Start auxiliary air compressor Enter aux. air stack cooling sequence Maintain cover gas flow until stack temperatures are less than 500°C	Enter normal stop sequence	Start auxiliary air compressor Enter aux. air stack cooling sequence Maintain cover gas flow until stack temperatures are less than 500°C	Depressurize Vent Fuel Side Start primary air cooling sequence Reestablish purge with nitrogen
Normal Operation	Divert turbine/generator power to resistance banks, SOFC to open circuit.	Enter normal stop sequence	Start auxiliary air compressor Enter aux. air stack cooling sequence Start and maintain cover gas flow until stack temperatures are less than 500°C	N/A
Normal Shutdown	Start auxiliary air compressor Enter aux. air stack cooling sequence Maintain cover gas flow until stack temperatures are less than 500°C	Continue normal shutdown sequence	Start auxiliary air compressor Enter aux. air stack cooling sequence Maintain cover gas flow until stack temperatures are less than 500°C	Depressurize Vent Fuel Side Start auxiliary air compressor Purge with nitrogen

3.3 Power System Installed Cost and Cost of Electricity Estimates

Table 3.7 summarizes the development of the power system installed cost, and in Table 3.8 the system COE estimate is developed. Mature-product conditions (circa 2008) are assumed. Input for the COE analysis is identified in Table 3.8.

Table 3.9 provides the COE estimate for the conventional power generation technology, a 20 MWe-class gas turbine/steam turbine combined cycle. At the reference fuel cost, \$3.00/MMBtu, the PSOFC/ATS-GT power system COE estimate is lower than the conventional-technology COE estimate by approximately 6%. Figure 3.53 shows the effect of fuel cost on the COE estimates for the PSOFC/ATS-GT and conventional-technology power systems. Due to its higher operating efficiency, the PSOFC/ATS-GT system will exhibit increasing COE advantage as the fuel cost rises above the \$3 value.

Figure 3.53 — Effect of fuel cost on COE estimates.

Table 3.7 — Power System Installed-Cost Estimate*

Installed Equipment Costs						
	Equipment	Freight	Installation	Totals		
SOFC Generator	8,183,259	28,000	42,101	8,253,360		
Gas Turbine System	4,900,000	7,000	33,535	4,940,535		
SOFC Power Conditioning System	3,592,000	10,500	21,666	3,624,166		
Instrumentation, Controls, and Electrical Cabinets	873,542	7,000	199,520	1,080,062		
Switchyard and Electrical Distribution	1,006,600	7,000	244,524	1,258,124		
Fuel Supply System	213,279	1,750	10,000	225,029		
Hydrogen Supply System	114,595	1,750	10,000	126,345		
Purge Gas Supply System	116,262	1,750	10,000	128,012		
Auxiliary Air Supply System	166,477	1,750	6,510	174,737		
Startup Boiler System	85,551	1,750	11,316	98,617		
Piping and Insulation	1,045,062	14,000	197,690	1,256,752		
Site Buildings				36,159		
Totals	20,296,627	82,250	786,862	21,201,898		
Project Cost Summary						
Installed Equipment						21,201,898
Project Management, Engineering, Construction Indirects, and Permitting						922,469
Site Preparation						427,238
Grading, utilities installation					146,072	
Foundations installation					231,435	
Structural steel installation					49,731	
G&A, R&D, Sales & Marketing, Profit Allowance						5,948,708
Total Plant Cost						28,500,313
Spare Parts Allowance						257,631
Startup						150,000
Land						20,000
Total Capital Requirement						28,927,943

*Site preparation and installation cost estimate provided by ICF Kaiser, Pittsburgh, PA under subcontract.

Table 3.8 — Power System Cost of Electricity Estimate

COE Calculation Basis		Cost Summary	
No. round-the-clock power system operators	1		
No. plants on system	5	Fixed O&M, \$/kW-year	
Operator labor cost, \$/man-hour	75	Plant operation & control	5.3
Housekeeping maintenance, man-hours/week/system	20	Housekeeping maintenance	1.1
Housekeeping labor cost, \$/man-hour	25	Administration (30% of operation & maintenance labor)	1.9
System rating, MW net ac	24.7	Total Fixed O&M, \$/kW-year	8.3
Gas turbine rating, MW ac	9.6	Variable O&M, mills/kWh	
Gas turbine maintenance cost, \$/GT kWh	0.00377	SOFC replacement	1.4
SOFC replacement cost, \$/SOFC generator module	468,920	Gas turbine maintenance	1.5
SOFC replacement interval, years	10	Desulfurizer adsorbent/catalyst replacement	0.1
Power plant capacity factor	0.92	Total Variable O&M, mills/kWh	3.0
Desulfurizer adsorbent & catalyst, \$/year	11,385		
Interest rate (SOFC replacement cost calculation), %	6.0	Total O&M COE, mills/kWh	4.0
Power system efficiency (net AC/LHV), %	59.87	Capital COE, mills/kWh	21.8
Capital charge rate, %/year	15	Fuel COE, mills/kWh	19.0
Power system capital cost, \$/kW	1,172	Total COE, mills/kWh	44.8
Fuel Cost, \$/MMBtu	3.0		
Fuel higher heating value, Btu/lb	21873		
Fuel lower heating value, Btu/lb	19711		

Table 3.9 — Conventional-Technology Power System Cost of Electricity Estimate

Information Source: Gas Turbine World 1997 Handbook, p. 24			
System Classification: Gas Turbine/Steam Turbine Combined Cycle			
Gas Turbine: General Electric LM 1600			
Steam Turbine Cycle: Two-pressure, 4.8 MWe			
Net System Power Output: 17.9 MWe			
System Efficiency: 47.9% (net AC/LHV)			
Turnkey Price: \$14.9M (\$830/kWe)			
Basis		COE Summary	
No. round-the-clock power system operators	1	Fixed O&M, \$/kW-year	
No. plants on system	5	Plant operation & control	7.3
Operator labor cost, \$/man-hour	75	Housekeeping maintenance	1.5
Housekeeping maintenance, man-hours/week/system	20	Administration (30% of operation & maintenance labor)	2.6
Housekeeping labor cost, \$/man-hour	25	Total	11.4
Power system maintenance, \$/kWh	0.007		
Power system capacity factor	0.92	Variable O&M, mills/kWh	7.0
Power system capacity, net AC kWe	17900		
Power system efficiency (net AC/LHV), %	47.9	Total O&M COE, mills/kWh	8.4
Capital charge rate, %/year	15		
Power system capital cost, \$/kW	830	Capital COE, mills/kWh	15.4
Fuel cost, \$/MMBtu (HHV)	3.00		
Fuel higher heating value, Btu/lb	21873	Fuel COE, mills/kWh	23.7
Fuel lower heating value, Btu/lb	19711	Total	47.6

3.4 Conceptual Design Trade-Off Studies

3.4.1 SOFC Generator Sizing

Introduction

An analysis was performed to determine the number of cell substacks that should be included in the SOFC generator, which in turn is to be integrated with the Solar Turbines Mercury™ 50 gas turbine in the PSOFC/ATS-GT hybrid power system. The analysis assumed that the engine would be operated at its design point, as established by the DOE Advanced Turbine Systems (ATS) program under which the Mercury™ 50 gas turbine was developed. The analysis task, then, was to define the SOFC generator that would best utilize the Mercury™ 50 in the integration of the PSOFC and ATS-GT technologies.

The criterion used in specifying the SOFC generator was minimum COE. This particular conceptual design project was proposed on the premise that applying the power-intensive gas turbine in a PSOFC/GT hybrid cycle power system would minimize the number of relatively high-cost SOFC stacks required by the power system. Due largely to the need to fire the gas turbine, the system will not achieve the highest electric efficiency, but an attractively-low COE estimate could still result.

Analysis Basis

- Cycle configuration – Figure 3.54.
- Gas turbine – The gas turbine employed by the power system, as already indicated, is the Solar Turbines Mercury™ 50 engine. The turbine was assumed to be operated at its design point, (combustor firing temperature = 2125°F, compressor air intake rate = 36.5 lb/s, pressure ratio ~ 9.5:1). It utilizes advanced gas turbine technology that was developed under the DOE ATS program.
- SOFC generator configuration – the generator is composed of pressurized generator modules. Each module is a horizontal pressure vessel housing twenty 576-cell substacks. (For reference, the EDB/ELSAM 100 kW SOFC generator includes the equivalent of two such substacks.) As Figure 3.54 indicates, the SOFC exhaust is the oxidant stream for the gas turbine combustor. In selecting the optimum number of cell substacks to be included in the power system, the number was varied

systematically, and in each case, the cell current density was selected to establish an SOFC generator combustion zone exhaust temperature of 870°C.

- The conventional SOFC stack configuration was assumed.
- Air heater operation – The air heater shown in Figure 3.54 is assumed to operate only during system startup. It is not fired during normal steady-state power operations.

Figure 3.54 — Simplified PSOFC/ATS-GT hybrid power system cycle.

- Cell V-I characteristic – the mature-product characteristic.
- SOFC stoichs model – it was required that the SOFC stoichs specification conform as closely as possible to the stoichs profile determined by thermal-hydraulic analysis for 1020°C peak-cell and 870°C combustion-zone exhaust temperatures.
- SOFC fuel consumption – 85%
- SOFC cell stack fuel bypass leakage – 1%

- Power lead cooling air flow – 0
- Power lead and stack internals power loss – assumed 1% of cell power.
- SOFC power conditioning system overall efficiency – 94%, a representative value.
(This study was performed before completion of the power conditioning conceptual design, wherein a PCS overall efficiency of approximately 95% was projected (Section 3.2.3.4). The optimum number of SOFC substacks was not affected.)

Discussion

Results of the SOFC generator sizing study are graphed in Figure 3.55. The graph shows system electric efficiency and COE plotted against the system net AC power output. These results apply to a hybrid cycle power system consisting of an SOFC generator, composed of the indicated number of cell substacks, integrated with a single Mercury™ 50 gas turbine. The estimated system efficiency increases with the number of cell stacks because the cell current density is able to move progressively to lower values at which the SOFC operating efficiency is higher. Initially, the system COE decreases in response to the addition of cell stacks, a direct result of the improving efficiency. However, the SOFC cost effect eventually becomes dominant, and the cross-over between efficiency dominance and installed-cost dominance establishes the optimum number of SOFC substacks. The estimated minimum COE occurs with 72 substacks. However, 80 substacks were selected for power system design because it causes a very small COE increase relative to the minimum, but an efficiency gain of nearly 6/10 of one percentage point is achieved. It is expected that the 80 substacks would be packaged in four horizontal cylindrical pressure vessels, 20 substacks per vessel.

A power block consisting of an SOFC generator composed of 80 cell substacks, and integrated with the Solar Turbines Mercury™ 50 gas turbine, will generate 12.3 MWe of net AC power at an efficiency of 59.5% (net AC/LHV). Two blocks are needed for a 20 MWe-class power system.

Figure 3.55 — Effect of PSOFC generator sizing on system performance and COE.

While an SOFC generator composed of 80 cell substacks will result in the minimum cost of electricity, it is recognized that the spread between the minimum and maximum COE values in Figure 3.55 is not large, only 3%. The advantage in designing with 80 cell substacks is that in addition to achieving the minimum COE value, a more attractive power system efficiency is also achieved. In addition, the cells are loaded less heavily at the minimum COE design point. At the 48 substack point, the cell current density is 400 mA/cm², while at the 80 substack point, it is less than 300 mA/cm².

Conclusion

The optimum (minimum COE) PSOFC/ATS-GT power system that employs a single, as-designed Solar Turbines Mercury™ 50 gas turbine will include an SOFC generator that consists of the equivalent of eighty 576-cell substacks. This system will generate 12.3 MWe (net AC) at an efficiency of 59.5% (net AC/LHV). Multiples of this system, or power block, could be arranged in parallel to achieve a desired total system power output.

3.4.2 Steam Turbine Bottoming Cycle Option

The reference PSOFC/ATS-GT power system discussed above is based on a hybrid-cycle power block in which a PSOFC generator is integrated with a Solar Turbines Mer-

cury™ 50 gas turbine. The exhaust gas at the recuperator exit, see Figure 3.7, is still fairly hot, approximately 360°C, and it could be applied to the generation of steam, which could be used in turn to drive a steam turbine/generator. Thus, as an alternative power system, one based on a PSOFC/ATS-GT/Steam Turbine (ST) cycle has been analyzed. Results are reviewed in this section. Note that in this work, only the effects of adding the steam turbine cycle to the baseline PSOFC/ATS-GT power system have been considered; the optimization of the PSOFC/ATS-GT part of the cycle was not revisited.

Gas turbine steam injection is another potential use for the steam produced. This option has not been considered since the study approach was to apply the Mercury™ 50 largely as it was developed by Solar Turbines during the DOE ATS program, as opposed to attempting any re-optimization of the engine (pressure ratio and TIT) for steam injection application.

Discussion

The cycle for the PSOFC/ATS-GT/ST power system is shown in Figure 3.56. The basic PSOFC/ATS-GT cycle is unchanged from its configuration in the reference power system discussed earlier in the report, but a heat recovery system generator (HRSG) and steam turbine are added. The HRSG is a once-through unit, consisting of economizer, evaporator, and superheater sections. The evaporator generates steam at 31.0 bara (450 psia) for supply to the turbine with 28°C (50°F) of superheat. The deaerator operates on steam extracted from the steam turbine at 1.5 bara (22 psia), 0.14 bara (2 psi) above the deaerator pressure. The steam turbine was assumed to be a condensing unit, and the condenser operating pressure was set at 0.04 bara (0.6 psia).

A summary of steam system design-point parameter values is provided in Table 3.10, and the estimated power system performance is summarized in Table 3.11. Note that the system considered was assumed to be composed of a single HRSG and steam turbine that operate on the combined exhaust from two PSOFC/ATS-GT power blocks. As defined above, each power block is composed of four PSOFC generator modules and a Mercury™ 50 gas turbine. Relative to the reference hybrid power system, the addition of the steam turbine cycle increases the system power output by approximately 1.5 MWe, and the system efficiency by nearly 4 percentage points.

Table 3.11
PSOFC/ATS-GT/ST Power System Design-Point Performance Estimates

Parameter	
Compressor air intake rate	33.1 kg/s (72.9 lb/s)
Compressor pressure ratio	9.5
Gas turbine combustor temperature	1160C (2125 F)
SOFC gross AC power	15.9 MWe
Gas turbine gross AC power	9.6 MWe
Steam turbine gross AC power	1.6 MWe
Steam turbine cycle efficiency (net steam cycle AC/cycle heat input)	29.7%
Auxiliary power requirement	0.8 MWe
System net AC power	26.2 MWe
Efficiency (net AC/LHV)	63.6 %
Carbon dioxide emission	317 kg/MWh
Nitrogen oxide emission (based on 5 ppmv addition at the gas turbine combustor)	.038 kg/MWh
Sulfur dioxide emission	0

3.4.2.1 Power System Cost and Cost-of-Electricity Evaluation

The power system installed-cost estimate, including the HRSG and steam turbine, is presented in Table 3.12. The equipment and installation costs for the steam turbine cycle equipment were obtained through contact with potential suppliers. Table 3.13 provides the power system COE development. The estimated COE for the system equipped with the steam turbine cycle is 49 mills/kWh, over 4 mills higher than the COE estimate for the baseline PSOFC/ATS-GT power system (see Table 3.8). An important factor in the COE estimate is the steam-system operator basis (number of operators and pay rate). As Table 3.13 indicates, this particular calculation assumed that one round-the-clock on-site steam system operator would be required, in addition to the power system operator who monitors/controls several power systems from a remote location.

COE vs fuel cost estimates are graphed in Figure 3.57 for the PSOFC/ATS-GT/ST power system, and the estimates for the conventional-technology and PSOFC/ATS-GT power systems are repeated from Figure 3.53. The curves indicate that the PSOFC/ATS-GT/ST power system, with a steam system operator, would have no COE advantage relative to the baseline PSOFC/ATS-GT power system, and advantage relative to the conventional-technology system only at moderate to high fuel costs. However, if the system design, by exploiting factors such as the boiler type, size, volume, etc, could do without the steam operator, it would reduce the PSOFC/ATS-GT/ST COE estimates by

approximately 4 mills over the entire fuel cost range. This would make it clearly competitive with the conventional-technology system, regardless of fuel cost, but there would still be no clear COE advantage relative to the baseline PSOFC/ATS-GT system. This result indicates that the positive COE effect of the additional power output and higher efficiency produced by the addition of the steam turbine cycle is largely canceled by higher installation and O&M costs.

Figure 3.57 — COE estimates for the PSOFC/ATS-GT/ST power system.

Table 3.12
Power System Installed-Cost Estimate
(PSOFC/ATS-GT/ST Power System)

Installed Equipment Costs	Equipment	Freight	Installation	Totals		
SOFC Generator	8,183,259	28,000	42,101	8,253,360		
Gas Turbine System	4,900,000	7,000	33,535	4,940,535		
SOFC Power Conditioning System	3,592,000	10,500	21,666	3,624,166		
Instrumentation, Controls, and Electrical Cabinets	873,542	7,000	199,520	1,080,062		
HRSG, Feed & Condensate System	276,900	7,000	138,500	422,400		
Steam Turbine	1,700,000	3,500	138,500	1,842,000		
Switchyard and Electrical Distribution	1,006,600	7,000	244,524	1,258,124		
Fuel Supply System	212,952	1,750	10,000	224,702		
Hydrogen Supply System	114,420	1,750	10,000	126,170		
Purge Gas Supply System	116,262	1,750	10,000	128,012		
Auxiliary Air Supply System	166,477	1,750	6,510	174,737		
Startup Boiler System	85,551	1,750	11,316	98,617		
Piping and Insulation	1,045,062	14,000	197,690	1,256,752		
Site Buildings				36,159		
Totals	22,273,024	92,750	1,063,862	23,465,795		
Project Cost Summary						
Installed Equipment						23,465,795
Project Management, Engineering, Construction Indirects, and Permitting						922,469
Site Preparation						427,238
Grading, utilities installation					146,072	
Foundations installation					231,435	
Structural steel installation					49,731	
G&A, R&D, Sales & Marketing, Profit Allowance						6,541,627
Total Plant Cost						31,357,130
Spare Parts Allowance						455,270
Startup						150,000
Land						20,000
Total Capital Requirement						31,982,400

Table 3.13
Power System Cost of Electricity Estimate
(PSOFC/ATS-GT/ST Power System)

COE Calculation Basis		COE Summary	
No. round-the-clock power system operators	1	Fixed O&M, \$/kW-year	
No. plants on system	5		
No. round-the-clock steam system operators per power system	1		
Operator labor cost, \$/man-hour	75	Plant operation & control	30.1
Housekeeping maintenance, man-hours/week/system	20	Housekeeping maintenance	1.0
Housekeeping labor cost, \$/man-hour	25	Administration (30% of operation & maintenance labor)	9.3
System rating, MW net ac	26.2	Total Fixed O&M, \$/kW-year	40.5
Gas turbine rating, MW ac	9.6		
Steam turbine rating, MW ac	1.6	Variable O&M, mills/kWh	
HRSG, steam turbine maintenance cost, \$/kWh	0.0106		
Gas turbine maintenance cost, \$/kWh	0.00377	SOFC replacement	1.3
SOFC replacement cost, \$/SOFC generator module	468,920	HRSG, steam turbine system maintenance	0.6
SOFC replacement interval, years	10	Gas turbine maintenance	1.4
Power plant capacity factor	0.92	Desulfurizer adsorbent/catalyst replacement	0.1
Desulfurizer adsorbent & catalyst, \$/year	11,367	Total Variable O&M, mills/kWh	3.3
Interest rate (SOFC replacement cost calculation), %	6.0		
Power system efficiency (net AC/LHV), %	63,634	Total O&M COE, mills/kWh	8.4
		Capital COE, mills/kWh	22.7
Capital charge rate, %/year	15	Fuel COE, mills/kWh	17.9
Power system capital cost, \$/kW	1,222	Total COE, mills/kWh	49.0
Fuel Cost, \$/MMBtu	3.0		
Fuel higher heating value, Btu/lb	21873		
Fuel lower heating value, Btu/lb	19711		

3.4.3 Advanced Ammonia-Water Power Cycle for the PSOFC/GT Hybrid System

The ammonia-water power system concept and the material presented in report Sections 3.4.3.1 to 3.4.3.4 were developed by Energy Concepts Company, Annapolis, MD.

3.4.3.1 Introduction

This section presents the results of an analysis of an ammonia-water absorption power cycle, which is adapted to be powered by the 360°C exhaust gas from the 20 MWe-class PSOFC/ATS-GT power plant. The ammonia-water power cycle is found to provide an additional 1.9 MWe from the waste heat - more than is possible with other waste-heat powered cycles.

This conceptual analysis consists of a description of the cycle; design heat and mass balance; indication of system size and footprint; and estimation of the installed cost and the operation and maintenance cost. This study is limited to consideration of this cycle exclusively in the stand-alone bottoming cycle role. Note that in this work, only the effects of adding the ammonia-water power cycle to the baseline PSOFC/ATS-GT power system have been considered; the optimization of the PSOFC/ATS-GT part of the cycle was not re-visited.

3.4.3.2 Background

Ammonia-water power cycles have been known for at least fifty years. Vapor phase ammonia is generated from the water absorbent at high pressure, expanded to produce work, and then absorbed at low pressure. The three recognized attributes of the ammonia-water power cycles are: they excel with low-temperature heat ($<350^{\circ}\text{C}$); the temperature glide of the vapor-generation process allows a close match to that of the heat source; and the high pressures allow the use of compact and inexpensive equipment. However, heretofore, various deficiencies have prevented their wide application. Prior-art absorption power cycles have been limited in the degree to which they can match heat-source temperature glide, thus limiting their useful work production and ability to provide high energy-conversion efficiency. The simple, traditional cycle provides only about a 20% advantage over the Rankine cycle. More efficient cycles, with a high level of recuperation of internal heat, can provide up to 70% improvement over the Rankine cycle, but they are more complex, and many of them involve a problematic total evaporation of the fluid mixture at high temperature. Energy Concepts Co. has been pursuing the development of highly integrated ammonia-water absorption refrigeration cycles [1,2] and power cycles [3] that avoid the technical risk and much of the complexity of the prior cycles. With the newly developed improved power cycle and ongoing component development, large amounts of waste heat in the temperature range of 100°C to 350°C can be economically converted to useful power, rather than discharged into the environment.

Historically low energy costs have inhibited technology that provides energy-efficiency improvements. There was not enough economic incentive to develop and adopt such technologies. However, the economic and environmental factors are changing; energy costs are increasing and there is rising concern over the CO_2 emissions associated with

inefficient power cycles. The advanced ammonia-water power cycle provides the one missing ingredient for the realization of hybrid fuel-cell power cycles with high energy-conversion efficiency.

3.4.3.3 Ammonia-Water Power Cycle

The ammonia-water power cycle is applied as the bottoming cycle on the 20 MWe-class PSOFC/ATS-GT hybrid system. The exhaust gas exiting the system recuperator is used as the heat source for the ammonia-water power cycle. Temperature and mass flow rates of the exhaust gas are 360°C and 121,900 kg/hr, the combined exhaust from two PSOFC/ATS-GT power blocks. The composition of the exhaust stream is: N₂ 74.6%, H₂O 9.6%, CO₂ 4.7%, and O₂ 11.2%.

Relative to the ammonia-water power cycle, the basic assumptions for the conceptual design are:

1. Turbine and generator efficiencies are 85% and 95%, respectively.
2. Cooling media is water; either from cooling tower or natural sources.
3. The material of construction for the heat/mass transfer units and piping is carbon steel, and an appropriate corrosion inhibitor is used.
4. Pressure losses in transfer lines and components are not considered for the cycle analysis at this stage.

Power Cycle Description. The power cycle, shown in Figure 3.58, consists of a heat-recovery unit, thermal generator, internal recuperator, ammonia absorber, and turbine plus electric generator. The system requires three solution pumps, one transfer-fluid pump, and a cooling-water pump. Only one solution pump is shown in Figure 3.58, the other two are integrated in the recuperator. For better control and integration of the power cycle, a closed flow loop with water as transfer fluid is used for heat recovery. This approach maintains low pressure drop on the exhaust gas side and thereby minimizes the backpressure for the gas turbine. About 90% of the heat from the transfer fluid is used for generating vapor in the thermal generator. The counter-current heat and mass transfer processes in the thermal generator facilitate an efficient heat extraction from the transfer fluid. The remaining heat is used in the recuperator, which is an

Figure 3.58 — Flow schematic of the ammonia-water power cycle.

integrated heat and mass transfer unit for recuperating internal and external heats. Ammonia vapor in the generator is internally superheated. The vapor then drives the turbine and the resulting low-pressure ammonia vapor flows to the absorber, where it is absorbed in the strong (high absorbing capacity) solution from the generator. The resulting weak (low absorbing capacity) solution is pumped to the recuperator, where it is divided into three streams for balancing the heat and mass transfer from different sections of the recuperator. As shown in Figure 3.58, one stream from the recuperator returns to the absorber, and also the recuperator and absorber are thermally coupled. Heat integration in the recuperator is a key design feature of the power cycle, because the overall conversion efficiency of the cycle depends on the thermal performance of the recuperator.

Heat flows in the major components are summarized in Table 3.14. The total thermal energy recovered from the exhaust gas stream is 10,305 kWth, which is 77.3% of the residual LHV of the exhaust gas. Of this, about 20% is converted to electric power. A simple Rankine cycle would convert only 12% of the heat recovered from the exhaust-gas stream. After accounting for pumping losses, the net energy produced by the ammonia-water cycle is 1,907 kWe. The recuperator recovers a significant part of the heat of absorption. The remaining heat of absorption is rejected to cooling water at inlet temperature of 21.1°C; 5,129 kWth from the absorber and 3,095 kWth from the recuperator.

Table 3.14 — Energy Flows in the Ammonia-Water Power Cycle

Components	Heat Rate kW	Power Generation kW
Heat recovery unit	10,305	
Generator – heat from transfer fluid	9,224	
Recuperator		
Heat from transfer fluid	1,081	
Heat rejected to water	3,095	
Absorber – heat rejected to water	5,129	
Turbine power		2,081
Electric generator power		1977
Pumping power		70
Net power		1,907

Cycle Parameters. A cycle-analysis model was used to determine the optimized design parameters of the cycle. The model calculates the thermodynamic state points of the cycle, the overall heat and mass balance, and heat and mass transfer rates for each component. Table 3.15 presents the thermodynamic properties and mass flow rates at the state points identified in Figure 3.58. The exhaust gas is cooled from 360°C to 93.8°C, by heating the transfer fluid from 76.7°C to 243.3°C. The transfer fluid is maintained at a pressure of 40 bar to prevent boiling. Ammonia vapor is generated at a pressure of 32 bar and superheated to 234.2°C, with ammonia concentration of 92% by weight. The outlet conditions from the turbine are: pressure, 2.05 bar; temperature, 50.7°C; and vapor quality, 0.94. The integrated system of recuperator and absorber provides this relatively low ammonia-absorption pressure, which is responsible for the high energy-conversion efficiency. Ammonia vapor from the turbine and the strong solution from the thermal generator at 103.9°C flow to the absorber. The weak solution at 29.4 °C is pumped to the recuperator. Two solution streams from the recuperator at 93°C and concentrations of 26% and 65% return to the thermal generator to complete the cycle.

Table 3.15 — State Points for the Ammonia-Water Power Cycle

State Point	Temperature °C	Pressure Bar	Flow Rate kg/s	Enthalpy kJ/kg	Concentration weight fraction
1	29.4	2.05	13.3	25	0.29
2	93.5		6.3	312	0.26
3	93.9		4.5	453	0.65
4	234.2	32.0	4.9	2183	0.92
5	50.7	2.05	4.9	1683	0.92
6	103.9	2.05	5.9	430	0.01
7	76.0	2.05	2.5	232	0.26
8	243.3	40.0	14.9		
9	94.2		14.9		
10	76.7		14.9		
11	360.0	~1	33.8		
12	93.8	~1	33.8		
13	21.1		40.6		
14	39.4		40.6		
15	21.1		64.7		
16	40.2		64.7		

System Dimensions and Interfacing. Components of the ammonia-water power cycle are assembled into three major subsections and installed as skid-mounted units. Their approximate dimensions are as follows. The absorber and associated components are assembled on a horizontal skid with a footprint of 3 x 10 m, and height of 3.4 m. The generator and recuperator are installed on a vertical skid with a footprint of 3 x 3 m, and height of 8 m. The ammonia turbine/generator set and the heat-recovery unit are installed in a skid with the same dimensions as the absorber skid, and can be mounted either alongside or on the top of the absorber skid, depending on plot space availability. When adequate supply of cooling water is not available at the site, a fourth skid is required for the cooling system. It is expected that with appropriate equipment rearrangement to facilitate the consolidation of exhaust gas flows from the two PSOFC/ATS-GT power blocks (see Figure 3.8), the footprint projected by the 20 MWe-class PSOFC/ATS-GT, including the ammonia-water power cycle equipment, would still be approximately 0.6 acre.

The major interfacing requirements include electric power line, cooling water, exhaust-gas supply line, instrument air, and sensor cables. An onboard-control and data-acquisition system monitors performance and can provide some local control capability. Main control and monitoring of the ammonia-water power cycle is performed from the central control room provided for the PSOFC/ATS-GT hybrid system.

3.4.3.4 Economics

Installed Cost. Table 3.16 presents the estimated installed cost for the ammonia-water power cycle. Two cost estimates are presented – one for near-term fabrication of a single unit, and the other for serial fabrication of 10 to 100 units per year in the year 2010. In the latter case, many economic benefits are achieved: volume procurement; improving component technology as field experience is gained; standard engineering-design package availability; and serial production and assembly infrastructure – jigs, fixturing, etc. Overall, the cost declines from \$3,880/kWe for the one-of-a-kind project to \$1,550/kWe for the serially produced product.

Table 3.16 — Installed Cost for the Ammonia-Water Power Cycle

	Present Cost (Single Unit) \$	Mature Technology Cost (2010) \$	Cost Reduction Factors*
Components			
Thermal generator	225,000	86,540	2.6
Recuperator	125,000	48,080	
Absorber	250,000	96,150	2.6
Ammonia turbine	800,000	307,690	2.6
Heat-recovery unit	300,000	187,500	1.6
Pumps plus controllers	100,000	62,500	1.6
Control and sensors	50,000	31,250	1.6
Subtotal	1,850,000	819,710	
Assembly and Installation	5,550,000	2,134,610	2.6
Installed Cost	7,400,000	2,954,320	
Cost/kWe	3,880	1,550	

* Factor 2.6 for technology development and volume production
Factor 1.6 for volume production

The costs for components unique to the power cycle (thermal generator, recuperator, and absorber) are based upon Energy Concepts experience in fabricating and procuring similar components, albeit sizes at only about 15% of the scale required here. Standard scale factors are applied to extrapolate to larger sizes. The remaining items are off-the-shelf components, and cost estimates are derived from vendor quotes for similar items.

Operation and Maintenance (O&M) Cost. Since the system uses waste heat, there is no fuel component of the operating cost. The primary cost is the pump power, which has already been factored in to calculation of the net power production. The system is designed to operate unattended, with displays and key controls in the central control room. Hence, no additional operating staff is required for the ammonia-water power cycle, beyond the operator already present in the control room. One person will be required for startups, scheduled shutdowns, and routine maintenance. The routine maintenance labor is estimated at 8 man-hours per week. The overall maintenance requirement is dominated by the ammonia turbine, followed by cooling tower and pumps. The annual total maintenance cost is estimated at \$75,000, plus an allowance for periodic major overhaul of the ammonia turbine.

3.4.3.5 PSOFC/ATS-GT/Ammonia-Water Power System Performance Estimates

Performance estimates for the PSOFC/ATS-GT/Ammonia-Water power system are summarized in Table 3.17.

Table 3.17
PSOFC/ATS-GT/Ammonia-Water Power System
Design-Point Performance Estimates

Parameter	
Compressor pressure ratio	9.5
Compressor air intake rate	33.1 kg/s (72.9 lb/s)
Gas turbine combustor temperature	1160C (2125F)
SOFC gross AC power	15.9 MWe
Gas turbine gross AC power	9.6 MWe
Ammonia-water power system gross AC power	1.9 MWe
Auxiliary power requirement	0.8 MWe
System net AC power	26.5 MWe
Efficiency (net AC/LHV)	64.4%
Carbon dioxide emission	313 kg/MWh
Nitrogen oxide emission (based on 5 ppmv addition at the gas turbine combustor)	0.038 kg/MWh
Sulfur dioxide emission	0

3.4.3.6 Power System Cost and Cost-of-Electricity Evaluation

An installed-cost estimate for the PSOFC/ATS-GT/Ammonia-Water power system is presented in Table 3.18. It was developed integrating estimates generated by Siemens Westinghouse and Caterpillar/Solar Turbines for the PSOFC/ATS-GT power system with the ammonia-water system costs developed by Energy Concepts Company and discussed above. The development of a COE estimate for the system is presented in Table 3.19. The ammonia-water system O&M cost was calculated using the \$75,000 annual maintenance estimated by Energy Concepts plus an assumed annual sinking-fund allowance that would accumulate \$500,000 in five years for a major overhaul of ammonia-water power system equipment. Viewing the systems as being similar in O&M complexity, it is noted that the resulting O&M cost, \$0.0106/kWh, was also applied in evaluating the COE for the PSOFC/ATS-GT/ST power system (see Table 3.13).

COE estimates for the PSOFC/ATS-GT/Ammonia-Water power system, in addition to those for the conventional-technology and the baseline PSOFC/ATS-GT systems are graphed in Figure 3.59. As in the analysis of the PSOFC/ATS-GT/ST power system, the PSOFC/ATS-GT/Ammonia-Water and baseline PSOFC/ATS-GT power systems are found to have similar COE performance projections, with the effect of additional power generation and efficiency produced by installing the ammonia-water power system being largely offset by higher installation and O&M costs.

Table 3.18
Power System Installed-Cost Estimate
(PSOFC/ATS-GT/Ammonia-Water Power System)

Installed Equipment Costs						
	Equipment	Freight	Installation	Totals		
SOFC Generator	8,183,259	28,000	42,101	8,253,360		
Gas Turbine System	4,900,000	7,000	33,535	4,940,535		
SOFC Power Conditioning System	3,592,000	10,500	21,666	3,624,166		
Instrumentation, Controls, and Electrical Cabinets	873,542	7,000	199,520	1,080,062		
Ammonia-Water Power System	2,956,000	10,500	277,000	3,243,500		
Switchyard and Electrical Distribution	1,006,600	7,000	244,524	1,258,124		
Fuel Supply System	212,954	1,750	10,000	224,704		
Hydrogen Supply System	114,421	1,750	10,000	126,171		
Purge Gas Supply System	116,262	1,750	10,000	128,012		
Auxiliary Air Supply System	166,477	1,750	6,510	174,737		
Startup Boiler System	85,551	1,750	11,316	98,617		
Piping and Insulation	1,045,062	14,000	197,690	1,256,752		
Site Buildings				36,159		
Totals	23,252,127	92,750	1,063,862	24,444,898		
Project Cost Summary						
Installed Equipment						24,444,898
Project Management, Engineering, Construction Indirects, and Permitting						922,469
Site Preparation						427,238
Grading, utilities installation					146,072	
Foundations installation					231,435	
Structural steel installation					49,731	
G&A, R&D, Sales & Marketing, Profit Allowance						6,835,358
Total Plant Cost						32,629,963
Spare Parts Allowance						553,181
Startup						150,000
Land						20,000
Total Capital Requirement						33,353,143

Table 3.19
Power System Cost-of-Electricity Estimate
(PSOFC/ATS-GT/Ammonia-Water Power System)

COE Calculation Basis		Cost Summary	
No. round-the-clock power system operators	1		
No. plants on system	5	Fixed O&M, \$/kW-year	
Operator labor cost, \$/man-hour	75	Plant operation & control	5.0
Housekeeping maintenance, man-hours/week/system	20	Housekeeping maintenance	1.0
Housekeeping labor cost, \$/man-hour	25	Administration (30% of operation & maintenance labor)	1.8
System rating, MW net ac	26.5	Total Fixed O&M, \$/kW-year	7.7
Gas turbine rating, MW ac	9.6		
Ammonia turbine rating, MW ac	1.9	Variable O&M, mills/kWh	
Ammonia-water power system maintenance cost, \$/kWh	0.0106	SOFC replacement	1.3
Gas turbine maintenance cost, \$/GT kWh	0.00377	Ammonia-water power system maintenance	0.8
SOFC replacement cost, \$/SOFC generator module	468,920	Gas turbine maintenance	1.4
SOFC replacement internal, years	10	Desulfurizer adsorbent/catalyst replacement	0.1
Power plant capacity factor	0.92	Total Variable O&M, mills/kWh	3.5
Desulfurizer adsorbent & catalyst, \$/year	11,367		
Interest rate (SOFC replacement cost calculation), %	6.0	Total O&M COE, mills/kWh	4.5
Power system efficiency (net AC/LHV), %	64.402	Capital COE, mills/kWh	23.4
Capital charge rate, %/year	15	Fuel COE, mills/kWh	17.6
Power system capital cost, \$/kW	1,259	Total COE, mills/kWh	45.6
Fuel Cost, \$/MMBtu	3.0		
Fuel higher heating value, Btu/lb	21873		
Fuel lower heating value, Btu/lb	19711

Figure 3.59 — COE estimates for the PSOFC/ATS-GT/ammonia-water power system.

3.4.4 Effect of Power System Type on Emissions

Estimates of CO₂, NO_x, and SO_x emissions for the PSOFC/ATS-GT power systems considered in this study are compared on a relative basis in Figure 3.60 with estimates for the conventional-technology system. For CO₂, the PSOFC/ATS-GT systems show advantage due to their higher efficiencies, and it is noted that their COE advantage would increase further in localities in which a carbon tax is levied. Further, it is recalled that earlier discussion indicated that the PSOFC/ATS-GT systems involving the steam turbine and ammonia-water power cycles could have little COE advantage relative to the baseline PSOFC/ATS-GT power system. However, this assessment was based on no consideration of a carbon tax, and considering the tax would increase the economic attractiveness of the bottoming cycles.

The NO_x estimates assumed an exhaust NO_x concentration of 25 ppmv (15% O₂) for the conventional-technology power system, and 5 ppmv (15% O₂) for the PSOFC/ATS-GT power systems. The latter value corresponds to the Mercury 50 development goal for operation on natural gas without the use of post-combustion pollution control devices.

Similar to the effect of a carbon tax, a tax on NO_x emission would improve the COE performance of the PSOFC/ATS-GT power systems relative to the conventional-technology COE, and it would also improve the COE performance of PSOFC/ATS-GT systems with bottoming cycles relative to the performance of the baseline PSOFC/ATS-GT system.

For the SO_x comparison, the reference value for the conventional-technology power system was determined assuming a fuel sulfur content of 4 ppmv, per the project design requirements, and the installation of no sulfur-emission control hardware. For the PSOFC/ATS-GT estimates, it was assumed that all system fuel was processed by the fuel desulfurization system, and that the fuel sulfur content, per SOFC requirement, was reduced by the system from 4 ppmv to 0.1 ppmv.

Figure 3.60 — Power system relative emissions estimates.

3.4.5 Power System Arrangement Studies

The power plant arrangement studies investigated the differences resulting from use of SOFC modules that had process air inlet and exhaust nozzles on opposite ends versus the nozzles on the same end. Since the process air piping will be large, heavy, and expensive, arrangements were examined to minimize the quantity of process air piping.

The arrangement that minimizes the process air piping for SOFC modules with nozzles on opposite ends is the reference arrangement that is discussed in Section 3.1.4. Although the amount of piping is greater, it has been selected for ease of installation and continuing inspection and maintenance activities. It occupies a smaller amount of real estate than the plant arrangements that support SOFC modules with inlet and exhaust nozzles on the same end.

Three alternative plant arrangements that use SOFC module pressure vessels with process air connections on the same end were also considered. The alternative 1 system is depicted in Figure 3.61 and Figure 3.62, and alternative 2 is shown in Figure 3.63 and Figure 3.64, and alternative 3 is shown in Figure 3.65 and Figure 3.66.

Figure 3.61 — Alternative 1 system arrangement — isometric view.

Figure 3.62 — Alternative 1 system arrangement — plan view.

Figure 3.63 — Alternative 2 system arrangement — isometric view.

Figure 3.64 — Alternative 2 system arrangement — plan view.

Figure 3.65 — Alternative 3 system arrangement — isometric view.

Figure 3.66 — Alternative 3 system arrangement — plan view.

3.4.5.1 Alternative 1 System Arrangement

This arrangement is shown on a site 63 m (208 ft) by 56 m (183 ft), or 0.87 acres. The pressure vessels are set like the spokes of a wheel on radii from a central air distribution plenum, and a vessel design is used in which the process air inlet and outlet nozzles are on the same end of the vessel. The vessel and SOFC stack design are relatively more difficult due to the need to ensure uniform air distribution within the stack. Its advantage is the minimization of the quantity of large bore, internally and externally insulated piping. The cost savings for elimination of the process air piping is partially offset by the cost of the air distribution plenum. This arrangement was investigated for its potential to yield lowest cost. It was hoped that the reduction in land area and process piping length would override the disadvantages. This was not the result. The advantage that was hoped to be gained by the alternative 1 configuration in its use of less land area did not materialize.

The primary disadvantages of the arrangement are inefficient routing of high voltage lines, difficult initial installation, and poor access to the SOFC vessels for maintenance.

3.4.5.2 Alternative 2 System Arrangement

This system would be installed on a site 69 m (226 ft) by 36 m (117 ft), or 0.61 acres. This site is only slightly larger than the preferred arrangement. The vessels are set in rows receiving and returning process air to air distribution headers at one end of the vessels. The vessel and SOFC stack designs are relatively more difficult due to the need to ensure uniform air distribution within the stack.

The alternative 2 site has vehicle access roadways on both ends of the vessels. Routing of high voltage electric power lines from the SOFC vessels to the switchyard is efficient.

The primary disadvantages of the arrangement are difficult initial installation and poor access to the SOFC vessel halves that have the process air nozzles.

3.4.5.3 Alternative 3 System Arrangement

This arrangement is a permutation of the alternative 1 arrangement with process air nozzles on one end of each vessel. The vessels are set like spokes of a wheel connected to a central air distribution plenum vessel. This arrangement occupies less land area than alternative one because the two power blocks are stacked one on top of the other. The plant site is smaller, only about 0.7 acre. Its dimensions are 51 m (166 ft) by 56 m (183 ft).

The relative advantage to alternative 1 is a smaller amount of real estate. The disadvantages to this arrangement are many. Initial installation will be difficult and routing of electrical power lines will be inefficient. Routine inspection and maintenance will be extremely difficult.

3.4.6 Desulfurization System Cost Study

This study compared an ambient temperature activated carbon desulfurization system with a heated catalytic hydrolization and adsorption system. The results of the study shows the heated system to be advantageous as discussed below.

The demonstration SOFC plants have used activated carbon beds contained within carbon steel pressure vessels to desulfurize natural gas. The significant advantage of the activated carbon is that the process works at room temperature. The main disadvantage is that the process is not a strong chemical bonding on the carbon and that the amount of sulfur adsorbed per unit of carbon is relatively low. As a result, the activated carbon quantities are large or the adsorbent must be replaced too frequently.

A study was performed to assess the cost of using alternate sulfur adsorbents. The use of zinc oxide was found to be very attractive. The volumetric consumption of zinc oxide is about two orders of magnitude less than the consumption of activated carbon for the same duty. In order for zinc oxide to be effective, however, the natural gas temperature must be above 277°C (530°F). The optimum desulfurization temperature is at about 400°C (750°F). Operation at elevated pressures also produces a substantial improvement in sulfur removal.

If using activated carbon as an adsorbent, a 20 MW_e plant is projected to consume 173 m³ (6100 ft³) of activated carbon per year. The material replacement cost is \$578,000

at \$2.50 per pound. Additionally, the disposal cost is \$145,000 at \$175 per 55-gallon drum. Thus, the annual cost for activated carbon is \$723,000.

This is compared to an estimated annual consumption of only 0.85 m³ (30 ft³) of zinc oxide, which operates at 12.4 barg (180 psig) and 400°C (750°F). The replacement cost for this material is \$8100 at \$270 per cubic foot. The disposal cost is estimated as \$715 per year for four 55-gallon drums. The desulfurization process does not rapidly consume the cobalt molybdenum catalyst, and practical equipment can be designed to require replacement every ten years. Assuming ten-year changeout, the annual replacement cost is \$920 for two 0.57 m³ (20 ft³) Co-Mo catalyst beds. The spent cobalt-molybdenum can be sold to a reclaiming company for about \$0.50 per pound. This gain is offset by the shipping cost with the result being no net gain or loss to dispose of the Co-Mo. The annual cost for desulfurization adsorbent and catalyst cost is \$9735. Combined with the cost of consuming about 180 kW for electrolysis of water to produce hydrogen, the total annual desulfurization cost is about \$90,000, considerably less than the annual activated-carbon cost. In addition, the first cost of the large vessels for activated carbon would be higher than the combined cost for the much smaller vessels for heated zinc oxide, the H₂ generator, the natural gas recuperator and electric gas heater, even with the added complexity of supporting separate adsorbent and catalyst beds. Thus, the study concluded that the preferred fuel desulfurization technique would employ the cobalt molybdenum catalyst/zinc-oxide-adsorbent technique.

3.4.7 Cover Gas System Cost Study

This study compared three alternatives for ways to provide a protective nitrogen cover gas for the SOFC stacks. The alternatives were using purchased bottled nitrogen gas; purchased liquefied nitrogen, which would be vaporized and used as needed; or nitrogen gas generated and stored on site. The cost-effective alternative is the generated-nitrogen approach.

The demonstration SOFC plants have relied upon supplies of NHMIX (nitrogen/hydrogen mixture) cover gas stored in pressurized bottles. Limited testing of SOFCs has shown that under some conditions it is not necessary to provide a reducing atmosphere on the fuel side of the SOFCs. It is expected that at the time that a 20

MWe plant is supplied, the process may require a cover gas consisting only of nitrogen. This study proceeded on this assumption.

In the current commodity gas market, 28.3 m³ (1000 ft³) of bottled nitrogen costs \$12.03.

The required quantity of nitrogen which must be stored at the plant are based on the following assumptions:

- The initial stack temperature is 566°C (1050°F). The stack-cooling rate after shutdown is -3°C/minute. Cover gas is needed until the stack temperature is reduced to 260°C (500°F). Thus, the minimum supply of cover gas must be sufficient for 183 minutes.
- The amount of cover gas needed for starting up the plant is the same as that needed for cooling the plant after shut down.
- Seven percent margin is added to the calculated gas quantities for uncertainty and conservatism.
- 242.5 cc/minute of nitrogen is needed for each SOFC. To support the 92,160 SOFCs in the plant, this results in 22 m³/min (790 scfm) of nitrogen.

Each plant start up or shut down consumes 4,380 m³ (155,000 scf) of nitrogen. The gas consumption cost is \$1,859 for nitrogen for a single shut down. For one plant start and stop per year, the direct annual gas usage cost is \$3,718. In addition to the cost of the gas, the demurrage for a large nitrogen tube trailer is \$950/month. The delivery fee for a trailer is built into the demurrage. Total costs for use of purchased bottled nitrogen gas is \$15,108 per year.

The costs for using purchased bottled gas are considered constant because the compressed gas industry is mature and gas is supplied as a commodity. Over a twenty year plant life, the total cover gas cost, based upon using bottled gases, is \$302,000.

The amount of nitrogen merits consideration for buying and storing in the liquid form. The dewar for storing 4,376 m³ (11,234 pounds or 154,525 scf) of liquid nitrogen must be at least 70.8 m³ (250 ft³). Liquid nitrogen costs \$3.40 per 28.3 m³ (1000 scf). The delivery to fill the dewar will include an excess of 5% (225 lbm) to allow for vaporization. This cost is \$552 for 4,594 m³ (162,251 scf) of liquid nitrogen plus a \$125 delivery fee. Every month the dewar must be topped off with an additional 5% to maintain the

minimum required inventory. This is an additional \$28 for the nitrogen plus a \$125 delivery fee. The monthly cost of dewar and instrumentation rental is \$750. The total cost for two fills of purchased liquid nitrogen, monthly replenishment and equipment rental, is \$11,388 per year. Over a twenty year plant life, the cover gas costs based upon using bottled hydrogen and liquid nitrogen sum to \$228,000.

As a basis of comparison, the installation of nitrogen generating equipment was assessed and resulted in the preferred (lowest cost) option. A prepackaged nitrogen generator system with accumulator, compressor, vent, pressure relief and pressure control is available for about \$20,500. The high-pressure gas storage bottles cost about \$60,000, for a total equipment cost of \$80,500. The unit has been sized to fill the high-pressure storage accumulator in seven days. While operating, the unit will consume electrical power and compressed air. Once the accumulator is filled, the unit will normally be idle. It will start upon low-pressure demand signal from the accumulator and operate until the accumulator is refilled. The functional specifications for the unit are shown in Table 3.20.

Table 3.20 — Nitrogen Generator/Compressor Characteristics

Nitrogen production rate	28.3 m ³ /hr (1000 scf/hr)
Nitrogen pressure before compression	6.9 barg (100 psig)
Nitrogen pressure after compression	165 barg (2400 psig)
Nitrogen purity	>98%
Process used	Selective Membrane
Package size	1.52 m x 1.83 m x 1.93 m (60" x 72" x 76")
Nitrogen separator storage tank	1.9 m ³ (500 gallons), ASME tank
Nitrogen compressor accumulator	Twelve – 2.19 m ³ (77.5 ft ³) ASME / CGA bottles
Electrical power requirements	110 vac / 60 Hz / 1 phase for separator 440 vac / 60 Hz / 3 phase, 5 kW for compressor
Compressed air supply required	1.9 m ³ /min (66 scfm) @ 175 psig

Table 3.21 summarizes the costs for the various options investigated for supply of a nitrogen system.

Table 3.21 — Costs for Options Investigated for Supply of Nitrogen System

	Bottled Nitrogen	Liquid Nitrogen	Nitrogen Generating System
	\$3,718 Annual Gas Cost (Delivery Included)	\$2,388 Annual Gas Cost & Delivery Fees	\$20,500 Initial N ₂ Equipment Cost (one time cost)
	\$11,400 Annual Bottle Trailer Demurrage	\$9000 Storage Dewar & Instrumentation Annual Rental	\$60,000 Bottle Trailer (one time cost)
Annual Cost	\$15,108	\$11,388	\$4025 pro-rated
20 Year Cost	\$302,000.	\$228,000	\$80,500

3.4.8 Hydrogen Gas Generation Cost Study

Hydrogen is needed by the fuel desulfurization process and to scavenge oxygen impurity from the nitrogen generated on-site. There are several suppliers of commercially available hydrogen generating equipment both in the USA and Canada. A prepackaged hydrogen generator/compressor system with accumulator and pressure control is currently available for \$245,000, but can be purchased in quantity (>100 unit per year) for \$61,250 to \$122,500 (Supplier's estimates). The functional specifications for the unit are shown in Table 3.22.

Table 3.22 — Hydrogen Generator Characteristics

Hydrogen production rate	0.57 m ³ /min (20 scfm)
Hydrogen pressure	12.4 barg (180 psig)
Process used	Electrolysis
Package size	1.82 m x 5.49 m x 3.05 m high (6 ft x 18 ft x 10 ft) high
Package weight	7484 kg (16,500 pounds)
Hydrogen accumulator size	119 sm ³ (4200 scf) @ 12.4 barg (180 psig) or 8.98 m ³ (317 ft ³)
Power	180 kW
Water consumption	35 liters/hr (9.24 gph)

3.4.9 Process Piping Cost Study

The first task in the process piping cost study was to determine the required flow area. When the airflow was assumed to be about 61 m/sec (200 fps or ~10% of sonic speed) an internal protective liner was necessary to prevent erosion of the insulation. At air temperatures in the range of 815°C to 870°C (1500 to 1600°F) metallic liners are unsuitable for except for short lengths due to their high thermal expansion. Ceramic

liners were found which have excellent erosion resistance but are unsuitable for other reasons. The ceramic liner materials are necessarily dense with high thermal conductivity and are fabricated by casting. The casting process requires a thickness of one inch. This dense ceramic adds 5 cm (two inches) to the piping inner diameter, provides no thermal insulation benefit, adds weight and is a very costly process.

When the inner flow area was selected to keep the gas speed to less than 30 m/sec (100 fps), conventional vacuum-formed alumina-silica insulation can be used. This does require that the piping be larger, but the complexity of fabricating insulated piping results in lower total cost.

The estimated costs of internally and externally insulated piping for the 20 MW_e plant are shown in Table 3.23.

Table 3.23 — High Temperature Piping⁽¹⁾ Costs

Application	Open Flow Area ID, inches	Pipe Size (0.375" wall)	Cost per linear foot ⁽²⁾
Header	22	38	\$1104
Branch	12	28	\$785

Notes:

1. 3" exterior insulation
2. Cost for piping and insulation only, no fittings, flanges or bolting materials

3.5 References

1. Erickson, D.C., Papar, R., Anand, G., and Tang, J., Refinery Waste Heat Powered Absorption Refrigeration – Cycle Specification and Design, Proceedings of ASME, AES-Vol. 38, pp. 391-402, 1998.
2. Erickson, D.C., and Anand, G., Absorption-Augmented Engine Refrigeration, ASHRAE Transactions, Vol. 105, 1999.
3. Erickson, D.C., and Anand, G., Absorption Power Cycle with Two Pumped Absorbers, Patent Pending.

4. CONCLUSIONS

- A low-cost, highly-efficient (60%) 20 MWe-class PSOFC/ATS-GT hybrid power system concept has been devised that employs the 4.5 MWe Solar Turbines Mercury™ 50 gas turbine that was developed under the DOE ATS program. The projected hybrid cycle power system COE is 44.8 mills/kWh, which is 6% less than the estimated COE for a conventional 20 MW-class gas turbine combined cycle power system. These estimates apply to a fuel cost of \$3/MMBtu. Due to its higher operating efficiency, the PSOFC/ATS-GT power system would have increased COE advantage at higher fuel costs.
- The PSOFC/ATS-GT power system efficiency estimate is 59.9% (net AC/LHV), twelve points higher than that achieved by the conventional-technology gas turbine combined cycle power system. This estimate is based upon mature-product cell V-I characteristics that are expected to be available in the 2005 to 2010 time period, and upon current Mercury™ 50 performance.
- An SOFC PCS concept was developed that has an estimated DC-to-grid AC conversion efficiency of 95% and an installed cost of approximately \$220/DC kWe.
- By equipping the PSOFC/ATS-GT power system with a bottoming cycle (steam turbine or ammonia-water), the overall system power output and efficiency will increase. However, due to the accompanying increases in system installed and O&M costs, the gains in power output and efficiency do not necessarily translate to significant COE reductions. It is likely that the economic attractiveness of adding a bottoming cycle to the baseline PSOFC/ATS-GT power system would increase if taxes on power system emissions were levied.
- The staged-cell SOFC stack design does not offer the large SOFC efficiency gain (over the standard cocurrent axial flow stack design) that was projected originally. Cell cooling in the fuel-entry cell rows reduces the average cell voltage while there is little increase in average fuel utilization at the last cell row at meaningful current densities because of the hazard of anode oxidation.

Recommendations:

- Development of the PSOFC/ATS-GT power system concept should continue to refine the conceptual design and performance estimates, establish the method of operating (startup, power operation, shutdown) the PSOFC/ATS-GT power system, identify and resolve SOFC and Mercury™ 50 integration issues, and develop a plan to demonstrate the power system concept.
- The PCS efficiency has a significant effect on the overall power system efficiency. For example, a gain in PCS efficiency of one percent translates to a system efficiency gain that approaches 6/10 of one percentage point. Additional conceptual work should be done to determine if PCS efficiency gains above the 95% value developed and applied in this study could be achieved cost-effectively.
- SOFC development should be pursued to improve fuel cell power density and efficiency. As with the SOFC PCS, improvements in cell efficiency would be effective in increasing the efficiency of SOFC/GT power systems.

APPENDIX A

The Effect of Staging on Efficiency of Isothermal SOFC Stacks

The purpose of this section is to describe the effect of staging the fuel stream on the efficiency of fuel cells. Staging of the fuel flow path is described as having the fuel pass a number of separate or segmented electrical circuits on its way through the fuel cell array. The current density is not uniform within any cell or stage that has a large difference in fuel concentration between the inlet and outlet locations. The terminal voltage across a stage is uniform, but the internal Nernst potential varies due to the change in reactant concentrations. With staging, one can either divide the cell structure into many regions or the fuel supply for many cells can pass sequentially through or past the cells. With staging the current density distribution becomes more uniform because the Nernst potential varies by a smaller amount within each stage. More uniform current density lowers the Joule losses and polarization losses. The voltage gain in the fuel rich stages is greater than the loss in the fuel poor stages.

It will be shown that the improvement in efficiency is small at economically viable operating current density conditions, namely at current densities near the maximum power point. The maximum power point occurs at the current density that maximizes the power output per unit area of the cell.

ISOTHERMAL ANALYSIS

The analysis is simplified by assuming that leakage of O_2 is negligible, that the cells are isothermal, and that the cathode gas is supplied at high stoich conditions. These conditions are favorable to the benefits of staging. For example, non-isothermal conditions within a stage create restraints on the flow and geometrical arrangements necessary to achieve uniform fuel consumption within the stages.

Consider that the fuel is being consumed by an electrochemical conversion as it flows along a path through a number of stages. The properties used in the analysis are applicable to solid oxide fuel cells but the generalized analysis is independent of the specific cell geometry. If the total in-stack fuel consumption along the path is F_c , the consumption in each of the n stages will be F_c/n and the consumption at the end of stage i is $z_i = (i/n) F_c$. The local fuel consumption within a stage is denoted by z . The change in consumption, dz , that occurs in elemental area, dA , is related to the local current density, j , by

$$dz = F_c (j \, dA/l_0) = (F_c/j_{av}) j \, dA/A_0 = (F_c/j_{av}) j \, dx$$

or

$$dx/dz = j_{av}/(j \, F_c) \quad (\text{A.1})$$

where dA/A has been replaced by dx .

If one also assumes that the electrical potential produced by CO is equal to that produced by H_2 , the fuel mole fraction (the sum of H_2 and CO mole fractions) is uniquely determined by the fuel consumption and the inlet composition. Thus with uniform cathode concentration or cathode composition also a function of consumption, the Nernst potential can be stated to be a function of the fuel consumption, $E_n = E_n(z)$.

The local current density at a location where the fuel consumption is z is given by

$$j = (E_n(z) - v_i)/R_c \quad (\text{A.2})$$

Substituting this into equation (1) gives

$$dx/dz = (j_{av} \, R_c/F_c)/(E_n(z) - v_i) \quad (\text{A.3})$$

After integration over the stage, the equation defining the cell voltage in each stage becomes

$$\int_{z_{i-1}}^{z_i} \frac{R_c \, dz}{(E_n(z) - v_i)} = \frac{F_c \Delta x}{j_{av}} \quad (\text{A.4})$$

For n equal area stages, $\Delta x = 1/n$ for each stage. The solution for the v_i of each stage is found by iteration using numerical integration of equation (4). If the resulting value of Δx is too small/large for an assumed v_i , the value of v_i in the next iteration must be increased/decreased subject to the limit that $0 < v_i \leq E_n(z_i)$. The cell resistance, R_c , was calculated as the sum of the internal resistance, the diffusion polarization resistance for each electrode, and the activation polarization resistance. The polarization resistances were obtained by dividing the anode or cathode polarization by the local current density. After one obtains the v_i , it is easy to plot the current density versus the non-dimension area using the values from the final iteration. The value of dimensionless area, x , corresponding to a given value of z in stage i is given by

$$x = (i-1)\Delta x + \frac{j_{av}}{F_c} \int_{z_{i-1}}^z \frac{R_c dz}{(E_n(z) - v_i)} \quad (\text{A.5})$$

The current density at this x location is given by equation (2).

RELATIONS FOR MOLE FRACTIONS, DIFFUSION POLARIZATION, AND NERNST POTENTIAL

The mole fraction of fuel in the anode gas stream at fuel consumption z is

$$y_F(z) = y_{Fo}(1 - z)$$

The mole fraction of oxygen in the cathode gas stream at fuel consumption z is

$$y_o(z) = \frac{(1 - z/s)}{(1/y_{oo} - z/s)}$$

where y_{oo} is the mole fraction of oxygen at inlet to the stack. The partial pressure of oxygen in the anode gas stream is found using an effective equilibrium constant for the methane derived fuel mixture,

$$p_{oA}(z) = \left\{ \frac{(1 - y_F(z) - y_N)K}{y_F(z)} \right\}^2$$

The Nernst potential based on the concentration in the gas streams is

$$E_n(z) = \frac{RT}{4F} \ln\left(\frac{py_o(z)}{p_{oA}(z)}\right)$$

The anode diffusion polarization is

$$\eta_A = \frac{RT}{4F} \ln\left(\frac{p_{oA}(z)}{p_{oAE}(z)}\right)$$

where

$$p_{oAE}(z) = K^2 \left(\frac{1 - y_N - y_{FAE}(z)}{y_{FAE}(z)} \right)^2$$

$$y_{FAE}(z) = y_F(z) - j(z)/S_3$$

$$p_{oA}(z) = K^2 \left(\frac{(1 - y_N - y_F(z))}{y_F(z)} \right)^2$$

The cathode diffusion polarization is

$$\eta_c = -\frac{RT}{4F} \ln(1 - j(z) / S_1 y_o(z))$$

ISOTHERMAL ANALYSIS RESULTS

The example is based on representative properties of a solid oxide fuel cell at 1000°C. The fuel composition is: 96% methane, 2% nitrogen and 2% CO₂. The fuel composition at the first stage inlet is based on internal reformation with a recirculation ratio that gives an oxygen to carbon ratio, OCR, equal to 2.1. Results are obtained at two system fuel consumption levels, 85% and 95%. The corresponding in-stack fuel consumption values, F_c , are 0.6839 and 0.8946. The combined fuel mole fraction of the inlet fuel is $y_{Fo} = 0.6326$ for both fuel consumption levels.

The reference current density was selected to be the value that gives the maximum power density for a single stage cell configuration. A plot of terminal voltage and power density as a function of current density is shown in Figure A.1 for $F_c = 0.85$. The values of current density at the maximum power point are $j_{maxP} = 0.522$ and 0.498 A/cm², for the 0.85 and 0.95 fuel consumption values, respectively.

Figure A.2 compares the current density and voltage distributions for single stage and four stage configurations at maximum power density for 85% system fuel consumption. For the four stage configuration, the total area is divided into four equal parts. The upper two curves, the Nernst potentials, are almost indistinguishable. The middle set of curves show the uniform voltage of the single stage and the four individual voltages of the four stage configuration. The bottom set of curves compare the current density distributions. Note that the maximum to minimum change in current density is reduced for the four-stage case. Since the average stage voltage is higher than the single stage voltage at the same current density and fuel consumption, the ratio of voltage in the same as the ratio of average power densities of the staged versus single stage cells. The average power density for this current density is increased by 0.58% above the value for a single stage cell for the limiting case of 16 stages.

Figures A.3 and A.4 show the same comparison at reduced current densities, $j_{\text{maxP}}/5$ and $j_{\text{maxP}}/20$, respectively. The staged power density only increases significantly at the lower current densities. Note that current density scales were changed in these figures to separate the two sets of curves.

Figures A.5, A.6, and A.7 provide similar comparison for 95% system fuel consumption. Table A.1 compares improvements in power density at 16 stages to that at a single stage. This table illustrates that the power density can be increased by significant amounts at a specified current density only for power densities that are well below the maximum power point. Thus the benefit of increased efficiency due to staging is only available at low power densities. Since staging is only beneficial at low power per cell, many more cells are required to compensate for the low power operation. Figure A.8 shows the relative number of cells required for a fixed plant output as a function of the current density.

The analysis was also applied to a generic planar cell configuration with a significantly lower total effective resistance. The total resistance including concentration polarization losses was held constant at 0.20 ohm-cm^2 . The maximum power density for this single stage planar cell occurs at a current density of 2.10 A/cm^2 for $F_c = 85\%$ and at 2.05 A/cm^2 for $F_c = 95\%$.

The increase in power output for Siemens Westinghouse cylindrical cells is compared to that of planar cells in Figure A.9. The curves show the increase in power output as a result of staging. The increase is shown as a percentage of the power output of the corresponding single stage cell. The current density has been made dimensionless on the respective current density at the maximum power point for the single stage cell condition. It is seen that the use of normalized current density results in an excellent correlation of the improvement in efficiency due to staging. The slight difference in the curves for the two geometries is due to the change in diffusion polarization resistance with current density in the cylindrical cell model. Otherwise the gains appear to be a function of normalized current and utilization.

Although the curve shows significant increase in output for $F_c = 0.95$, the partial pressure of O_2 at the anode/electrolyte interface exceeds the limit which is two orders of magnitude less than the equilibrium partial pressure of O_2 for the Ni/NiO/O_2 reaction. Oxidation of the anode is considered likely under these conditions. The diffusion con-

ductance of Siemens Westinghouse Power Corporation cells was used to evaluate the partial pressure of fuel at the electrolyte interface. For recirculated fuel, the fuel utilization limit is $FU = 91\%$ when the partial pressure of O_2 is maintained below the limit given above.

The effect of once through fuel flow on the increase in output due to staging was also considered. With the same fuel supply, external reformation at oxygen to carbon ratio of 2.1 gives a fuel inlet mole fraction of 0.780. The in-stack fuel utilization value equals the system fuel utilization. Results for the planar cell with constant resistance are shown in Figure A.10 for 85% and 95% system fuel utilization. As expected, the gain due to staging increases significantly if one can push the operation to higher fuel utilization for once through fuel flow (external fuel reformation). Limiting the partial pressure of O_2 to two orders of magnitude below the equilibrium value restricts fuel utilization to $FU = 89\%$.

SUMMARY OF ISOTHERMAL ANALYSIS

The results demonstrate that the effect of fuel staging on fuel cell power output is very small for current densities near the maximum power point. This is shown to be valid also for other geometries that have considerably lower resistance than the cylindrical cell. With fuel recirculation, the output at the maximum power point increases by 0.58 and 1.15% at 85 and 95% fuel utilization, respectively. The improvement due to fuel staging increases as current density is reduced or as fuel utilization is increased. Oxidation of the fuel electrode is likely with recirculated fuel for utilization above 91%.

The improvement due to staging is greater for systems with once through fuel flow as compared to those with recirculated fuel. However, oxidation of the anode is likely for once through flow when fuel utilization exceeds 89%. Although fuel staging could increase output at low current densities by more than 10% at 95% utilization, the low power density would result in a large increase in system cost.

Nomenclature for Appendix A

A_o	Total cell area, [cm ²]
E_n	Nernst potential, [V]
F_c	In-stack fuel consumption, [-]
F_{cs}	Overall system fuel consumption, [-]
I_o	Total cell current generated by n stages = $j_{av} A_o$.
i	Stage index, [-]
j	Local current density, [A/cm ²]
j_{av}	Average current density for the stage, [A/cm ²]
j_{maxP}	Current density that gives maximum average power density for a single stage Cell, [A/cm ²]
K	Effective equilibrium constant for the methane derived fuel mixture, [-]
n	Number of stages, [-]
p	Pressure of the fuel cell gases, [Atm]
P_d	Average power density, [W/cm ²]
$p_{OA}(z)$	Local partial pressure of oxygen in the anode gas stream, [Atm]
$p_{OAE}(z)$	Local partial pressure of oxygen at the anode electrolyte, [Atm]
R_c	Cell resistance index, [ohm-cm ²]
S_1	Diffusion conductance of the air electrode at one Atm, [A/cm ² -Atm]
S_3	Diffusion conductance of the fuel electrode at one Atm, [Atm/cm ² -Atm]
V_i	Terminal voltage of stage i, [V]
x	Non-dimensional cell area, = A/A_o , [-]
Δx	Non-dimensional area per stage, [-]
$y_F(z)$	Mole fraction of fuel in the anode gas at fuel consumption z, [-]
$y_{FAE}(z)$	Mole fraction of fuel at the anode electrolyte at fuel consumption z, [-]
y_{Fo}	Fuel mole fraction in anode gas at inlet to the stack, [-]
y_N	Mole fraction of nitrogen in the fuel at inlet to the stack, [-]
$y_O(z)$	Mole fraction of oxygen in the cathode gas stream at fuel consumption z, [-]
y_{OO}	Mole fraction of oxygen in the cathode gas at inlet to the stack, [-]
z	Local in-stack fuel consumption, [-]
z_i	In-stack fuel consumption at exit of stage i, [-]
η_A	Anode diffusion polarization, [V]
η_C	Cathode diffusion polarization, [V]

Figure A.1 — Power density and terminal voltage for a single stage cell.

Comparison of Single and 4 Staged Fuel @ max Power

Figure A.2 — Emf and current density distributions in a 4 staged fuel cell.

Comparison of Single and 4 Staged Fuel @ $j=j_{\max P}/5$

Figure A.3 — Emf and current density for single and 4 staged cell.

Comparison of Single and 4 Staged Fuel @ $j=j_{\max P}/20$

Figure A.4 — Emf and current density for single and 4 staged cell.

Comparison of Single and 4 Staged Fuel @ max Power

Figure A.5 — Emf and current density for single and 4 staged cell.

Comparison of Single and 4 Staged Fuel @ $j=j_{\max P}/5$

Figure A.6 — Emf and current density for single and 4 staged cell.

Comparison of Single and 4 Staged Fuel @ $j=j_{\max P}/20$

Figure A.7 — Emf and current density for single and 4 staged cell.

Table A.1
Effect of Staging on Average Voltage or Power Density

j_{av} , [A/cm ²]	F_{CS} , [-]	P_d for 1 stage, [W/cm ²]	$\Delta P_d/P_d$ for 16 stages, [%]
0.5220	0.85	0.2268	0.58
0.1044	0.85	0.0785	1.53
0.0261	0.85	0.0205	4.58
0.4980	0.95	0.2099	1.15
0.0996	0.95	0.0707	5.06
0.0249	0.95	0.01786	11.87

Figure A.8 — Relative number of cells required for fixed power output as a function of current density.

Comparison of Increase in Power Output Due to Staging.

Figure A.9 — Increase in the power output due to staging for planar and cylindrical cells vs. dimensionless current density.

Once Through Flow Compared with Recirculation Fuel Flow, Staged Planar Cells

Figure A.10 — Comparison of effect of once through and recirculation fuel flow on the benefits of cell staging.

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