Solid Oxide Fuel Cell Hybrid System for Distributed Power Generation

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Nguyen Minh
Principal Investigator
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GE Hybrid Power Generation Systems
19310 Pacific Gateway Drive
Torrance, CA 90502

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Abstract

This report summarizes the work performed by Hybrid Power Generation Systems, LLC (HPGS) under Cooperative Agreement DE-FC26-01NT40779 for the U. S. Department of Energy, National Energy Technology Laboratory (DoE/NETL) entitled "Solid Oxide Fuel Cell Hybrid System for Distributed Power Generation". The main objective of this project is to develop and demonstrate the feasibility of a highly efficient hybrid system integrating a planar Solid Oxide Fuel Cell (SOFC) and a gas turbine.

A conceptual hybrid system design was selected for analysis and evaluation. The selected system is estimated to have over 65% system efficiency, a first cost of approximately \$650/kW, and a cost of electricity of 8.4 cents/kW-hr. A control strategy and conceptual control design have been developed for the system.

A number of SOFC module tests have been completed to evaluate the pressure impact to performance stability. The results show that the operating pressure accelerates the performance degradation.

Several experiments were conducted to explore the effects of pressure on carbon formation. Experimental observations on a functioning cell have verified that carbon deposition does not occur in the cell at steam-to-carbon ratios lower than the steady-state design point for hybrid systems.

Heat exchanger design, fabrication and performance testing as well as oxidation testing to support heat exchanger life analysis were also conducted. Performance tests of the prototype heat exchanger yielded heat transfer and pressure drop characteristics consistent with the heat exchanger specification.

Multicell stacks have been tested and performance maps were obtained under hybrid operating conditions. Successful and repeatable fabrication of large (>12" diameter) planar SOFC cells was demonstrated using the tape calendering process. A number of large area cells and stacks were successfully performance tested at ambient and pressurized conditions.

A 25 MW plant configuration was selected with projected system efficiency of over 65% and a factory cost of under \$400/kW. The plant design is modular and can be scaled to both higher and lower plant power ratings.

Integrated gasification fuel cell systems or IGFCs were developed and analyzed for plant sizes in excess of 200 MW. Two alternative integration configurations were selected with projected system efficiency of over 53% on a HHV basis, or about 10 percentage points higher than that of the state-of-the-art Integrated Gasification Combined Cycle (IGCC) systems.

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Executive Summary

This report presents the work leading to the conceptual design of the full-scale hybrid system, incorporating a Honeywell Parallon 75 microturbine and planar solid oxide fuel cell (SOFC). The system design with anode and cathode recycle was downselected. This system concept is estimated to have over 65% efficiency and the highest reliability and availability compared to the other four system concepts evaluated. The down selected system is estimated to have a first cost of \$646/kW and a cost of electricity of 8.4 cents/kW-hour. Part-load analysis of this system showed that the efficiency of the system remains over 60% down to power levels of 200 kW. Several trade studies were completed for the control of this system. These trades helped develop an understanding of the controllability of the system and investigate strategies for controlling the proposed system.

This report also summarizes the results from the phase 1 SOFC barrier resolution task. In particular, the experimental work related to the impact of pressure on cell performance degradation and the carbon formation boundary and the results from the high temperature heat exchanger barrier resolution task.

A number of SOFC module tests have been completed to evaluate the pressure impact to performance stability. The results show that the operating pressure accelerates the performance degradation. The potential causes of the performance decay include oxidation of the metal interconnect, chromium poisoning to electrodes, interface resistance increase, electrode microstructure changes, back diffusion/leakage, and/or electrode conditioning process. The dominant degradation mechanisms remain unclear. Both interconnect oxidation and Cr transport to cathode were evident based in post-test analysis.

Several experiments were conducted to explore the effects of pressure on carbon formation. While carbon formation is favored at pressurized conditions, the conditions at which carbon formation occurs in a functioning cell is difficult to predict. Experimental observations on a functioning cell have verified that carbon deposition does not occur in the cell at steam-to-carbon ratios lower than the steady-state design point for hybrid systems. These results are in good agreement with theoretical analysis. These experiments were conducted on 3-inch and 4 3/8-inch SOFC cells operating at 4 atmospheres.

The high temperature heat exchanger barrier resolution task included heat exchanger design, fabrication, and performance testing, as well as oxidation testing to support the heat exchanger life analysis. Performance tests of the prototype heat exchanger yielded heat transfer and pressure drop characteristics consistent with the heat exchanger specification. Coupon testing on Inconel 625 and Haynes 230 samples completed over 3000 hours of exposure to SOFC conditions. A correlation of the oxide layer thickness as a function of time has been determined from the coupon sample data at three temperature.

Several SOFC stacks built with 6.3-inch diameter cells have been tested under hybrid conditions to assess their operating characteristics and the scalability of their stack design for hybrid applications. Also, theoretical and experimental study of the cell fabrication process led to the successful and repeatable fabrication of large (>12" diameter) planar SOFC cells using the tape calendaring process. A number of large area cells and stacks were successfully performance tested at ambient and pressurized conditions.

System scaleup strategies and integrated gasification fuel cell (IGFC) system concepts were also developed (see Appendix A and Appendix B). The impact of two advanced technologies on IGFC systems, Ion-Transfer Membrane and mid-temperature cleanup technology, has been evaluated and estimated to improve the overall system performance of the baseline system by 0.7% and 0.2%, respectively. Several system concepts have been identified that may have the potential of achieving the 60% system efficiency target. To reach 60% efficiency, an IGFC concept could require "revolutionary" performances from an SOFC not under development at this time.

Experimental

All experimental work performed on the program was contained in Technical Barrier Resolution effort. Experimental methods employed in this task is described and discussed in that section.

Results and Discussion

1 SYSTEM DESIGN

1.1 Conceptual System Design

Four different system configurations were considered in the trade study from which one configuration is down selected for further analysis and optimization. The down-selection to one system concept is based on system efficiency, reliability, cost, and the cost of electricity (COE). COE provides a means to trade system efficiency against reliability and cost.

The approach undertaken in the trade study consists of the following steps: (i) The efficiency of all system concepts is analyzed as functions of system parameters; (ii) A local maximum of the resulting system efficiency function is determined for each system concept; (iii) System components are identified for each candidate system concept (some components may be common across the candidate systems); (iv) System cost and reliability are estimated; (v) COE models are created and system COE is estimated; (vi) The system design point is adjusted if necessary to improve system COE at the acceptable expense of system efficiency; (vii) Steps (i) through (vi)) are repeated until an optimized system design is found for each candidate; (viii) The system with the "best" optimized solution is down selected.

System performance (efficiency), reliability, first cost, and COE analysis are described in the following sections.

1.1.1 Performance Analysis

A steady state system performance model is the basis for estimating system power and efficiency for a set of component performance assumptions. The components are modeled in a sequential modular fashion using the ASPEN PLUS steady state platform and its thermodynamic database. The model has been optimized for speed, thus enabling efficient system optimization using a large number of model runs.

The system is optimized at the peak power of the Parallon 75 microturbine, corresponding to its peak speed, while the SOFC design and operating points are varied. Thus, the air flow and pressure is fixed, but the SOFC cell operating voltage, fuel utilization, and total electrical power output is varied to maximize system efficiency while maintaining all design constraints of the system.

In general, efficiency increases as the ratio of the power output from the SOFC stack to that of the turbine increases. This is because the fuel cell is more efficient than the turbine. Thus, at this peak condition there is no firing of fresh fuel to increase the turbine inlet temperature. The stack exhaust temperature and the fuel utilization determine the turbine inlet temperature. However, in order to achieve high component reliability, the turbine inlet temperature is restricted. Similarly, temperature limitations are set for the burners, the recuperators, the high temperature air preheaters, and the stack.

The sensitivity of cell voltage (at 80% fuel utilization) and stack fuel utilization (at cell voltage of 0.75V) on systems efficiency and net system power output is shown in

for concept 1. The resulting net power output for concept 1 is in the range of 500 to 900 kW depending on the cell voltage and fuel utilization. The net power output increases as the cell voltage or the fuel utilization increases. This is because the efficiency of the stack increases as the cell voltage or the fuel utilization increases, enabling the power output to be increased. The turbine power output also increases due to increased total inlet fuel flow. The system efficiency is in the range of 60 to 65%, and it increases as the cell voltage or fuel utilization increases.

Figure 1 Sensitivity of System Efficiency and Power due to Cell voltage and Cell Utilization for Concept 1 (at 80% fuel utilization and 0.75 V)

The sensitivity of cell voltage and stack fuel utilization on systems efficiency and net power output for concepts 2, and 4 are shown in **Error! Reference source not found.** and **Error! Reference source not found.**. A summary of the analysis results for the four concepts is shown in Table 5 and Table 6. The range of cell voltage and fuel

utilization considered for each concept remained the same at 0.7 – 0.8 volts and 0.70 – 0.85, respectively.

Figure 2 Sensitivity of System Efficiency and Power due to Cell Voltage and Utilization for Concept 2 (at 80% fuel utilization and 0.75 V)

Figure 3 Sensitivity of System Efficiency and Power due to Cell Voltage and Utilization for Concept 4 (at 80% fuel utilization and 0.75 V)

	Concept 1	Concept 2	Concept 3	Concept 4
Net Power, kW	500-900	520-820	200-250	450-650
System Efficiency, %	60-65	60-65	60-65	64-70

Table 5 System Performance Analysis Results Summary

	Concept 1	Concept 2	Concept 3	Concept 4
SOFC Power (kW)	563	581	222	495
Turbine Power (kW)	86	98	17	78
Parasitic Power (kW)	35	53	15	56
Net System Power (kW)	614	626	224	517
System Efficiency (%LHV)	63	62	64	67

Table 6 Systems Performance Analysis Results (at 75% fuel utilization and cell voltage of 0.75V)

The power output from concept 3 is in the range of 200 to 250 kW, less than half of the power achieved by the other concepts. This is mainly due to the low power output from Parallon 75 caused by a low utilization of air. Because of its low power output, this concept is not analyzed as extensively as the others.

The size of the stack depends on the operating voltage and cell utilization in addition to its power level. Operating cell voltage increases as the current density decreases. Similarly, operating fuel utilization can also be increased by decreasing the operating current density while maintaining the same cell voltage. As the current density decreases, the stack size has to be increased to maintain the same power output, thus increasing the total systems cost. Therefore, there is a trade off between the system efficiency and the total systems cost, which directly affects the cost of electricity.

1.1.1 System Reliability Analysis

The reliability of a plant is the probability that the system will perform its intended function without failure under stated conditions for a specified period of time. It can be defined as

$$reliability = 1 - FOF, \quad (1)$$

in which *FOF* is the forced outage factor, the fraction of time the system is forced to shut down and does not include scheduled shutdowns. The system availability is the probability that the system will perform its intended function at anytime, when used under stated operating conditions. Thus it is the fraction of time the system is operational and includes both forced outages and scheduled outages.

Detail life, maximum and minimum repair time, and service interval data was gathered for all major components for each system concept. This data was used to estimate component reliability and availability. These estimates were then rolled up to determine the overall plant reliability for one year of continuous operation and plant availability over 10 years of operation. The number of forced shut downs as well as service maintenance intervals were also estimated to determine operation and maintenance costs. Table 7 and Table 8 summarizes plant level and subsystem level reliability and availability results for each concept. For these calculations the cell voltage and the cell utilization are fixed at 0.75 and 80%, respectively. This assumption is necessary to determine the heat and mass balance through the system and therefore the operating conditions of all the components.

The reliability of Concept 3 is 0.91, the lowest among the four concepts. The reliability of this concept is driven by its need for a high temperature air preheater.

Concept	Total	Subsystems				
		Fuel Processing	Microturbine	Power Electronics	SOFC	BOP
1	0.995	0.998	0.999	0.999999	0.999	0.999
2	0.994	0.998	0.998	0.999999	0.999	0.999
3	0.907	0.999	0.999	0.999999	0.999	0.910
4	0.997	0.999	0.999	0.999999	0.999	0.9998

Table 7 Reliability Results (at cell voltage of 0.75V and fuel utilization of 80%)

Concept	Total	Subsystems				
		Fuel Processing	Microturbine	Power Electronics	SOFC	BOP
1	0.93	0.99	0.98	0.996	0.995	0.97
2	0.94	0.99	0.98	0.996	0.995	0.98
3	0.94	0.99	0.98	0.996	0.995	0.98
4	0.95	0.99	0.97	0.996	0.995	0.996

Table 8 Availability Results (at cell voltage of 0.75V and fuel utilization of 80%)

Component life information was gathered from several sources. The life of the major components of the turbo-machinery was extracted from the detail life analysis reports on the Parallon75 components. The heat exchanger and the other balance of plant data were gathered from publications on previous development programs, including the advanced DOE micro-turbine program and heat exchanger programs. This data has been supported with information gathered from the Internet. The stack and fuel processor life numbers are based on current engineering knowledge and expert opinion.

The basic assumption behind the reliability model is that the components realize wear in time and the probability of failure is not constant over time. Therefore, the Weibull probability of failure is used. The Weibull slope for all components in this study is between 1 and 4, indicative of increasing probabilities of failure as time increases. In such cases, it has been shown in practice that a scheduled replacement may be cost effective. Consequently, in availability calculations when a maintenance interval is identified for a certain component, the component is assumed to be replaced with a new unit and the component life used in the overall plant reliability is assumed to be reset.

A simple rule of thumb was used to estimate the life of high temperature components and also the temperature dependence of their life. The nominal life of these components were assumed to reduce to half their original life for every 25 degree F increase in operating temperature. This rule of thumb assumes no change in materials from the baseline case and is based on expert judgment.

The availability spreadsheet uses a Monte Carlo simulation to estimate the expected plant availability, the standard deviation of the availability calculation, and the expected number of forced and scheduled plant outages. This analysis assumes a normal probability distribution for the life, i.e. a Weibull distribution with beta set at 1.0. This spreadsheet also requires input of the repair time duration for the availability calculation.

1.1.2 System Cost Analysis

The first cost of the overall system for each system concept was assessed using a cost roll-up model. Component specifications, obtained from system performance analysis, were used to estimate component sizing and then cost. Cost estimates have been made on all major components including the micro-turbine, stack, fuel processor, and the Balance-of-Plant (BOP). The BOP includes the thermal management sub-system, the air, fuel and water delivery sub-systems, and the controls and power electronics subsystems.

All cost estimates have been based on a production volume of 500 units/year or approximately 250 MW per year.

For sub-systems requiring significant technology development, a bottom-up cost model approach was used. For example, a dedicated stack cost model was constructed with the capability to conduct sensitivity analyses. The cost model itemizes the cost into four major components: materials, labor, equipment, and facilities costs. These costs are estimated based on the number of cells and stacks to be build. The total number of cells and stacks are estimated based on a stack performance model. The model makes a projection of the stack performance in the near future. The fuel processor cost was estimated to be the same as that of the stack. This is consistent with TIAX estimation.¹

Cost estimates for sub-systems containing, to a large degree, currently available technology were derived from quotes obtained from vendors. Adjustments were subsequently made to these quotes to ensure a consistent cost basis. These adjustments were based on engineering judgment. Cost information contained in vendor catalogs were also used as a guide. This approach was used on the micro-turbine, air, fuel and water delivery subsystems and other balance-of-plant (BOP) components. Quotes from vendors, compiled previously for all Parallon 75 micro-turbine, were used as the basis for many of the BOP parts.

Concept No.	1	2	3	4
Systems Cost, \$/kW	545	563	855	646

Table 9 System Cost Results (at cell voltage of 0.75V and fuel utilization of 80%)

Results from the cost model are summarized in Table 9. The cost ranges from \$545/kW to \$855/kW. Concept 3 has the highest cost per kW because its power output is low compared with the other concepts.

1.1.3 Cost of Electricity Analysis

The Cost of Electricity (COE) model integrates the performance, reliability, and the cost models. A schematic of the information flow is shown in **Error! Reference source not found.** COE is composed of three parts: the cost of fuel, capital cost, and the cost of operation and maintenance.

¹ "Scale-up Study of 5 KW SECA Modules to a 250 kW System", TIAX LLC Final Report to DOE/NETL, Reference: 74313, June 10, 2002

Figure 4 Schematic of the Integrated COE Model

The capitalized cost of the system includes the system manufacturing cost, the installation cost, the other costs associated with installation and operation startup (such as foundation and civil works, and safety systems, etc.) and the owner's costs (owner's engineer, finance costs, etc). The system manufacturing costs, including the microturbine, the stack, the fuel processor, and the Balance-of-Plant (BOP), are estimated in the cost model. The other costs are estimated to be about the same magnitude as the manufacturing costs.

The fuel cost is computed from the total fuel used and the unit cost of fuel. The total fuel used is related to the power plant system efficiency as well as the total power output. The efficiency and net kW output are estimated using ASPEN Plus steady state model. The cost of the natural gas depends on many factors and fluctuates day to day. The historical price of natural gas is shown in **Error! Reference source not found..** The price ranges from \$2 to \$16/MMBtu-HHV. In this analysis it is assumed to be \$6/MMBtu-HHV.

Figure 5 Historical Natural Gas Price²

The operation and maintenance cost are related to the cost and frequency of part replacement and maintenance. The frequency of service is computed using the component availability model described above. The frequency of parts replacement was obtained statistically and the replacement costs were estimated.

Results from the COE model are summarized in Table 10. The cell voltage and the cell utilization are fixed at 0.75, and 80%, respectively for this analysis. The COE of Concept 3 is the highest as a result of its highest capital cost.

Concept No.	1	2	3	4
COE, ¢/kW	8.4	8.1	11.8	8.4

Table 10 Summary of COE Results (at cell voltage of 0.75V and fuel utilization of 80%)

Cell voltage and fuel utilization sensitivity on COE have been evaluated. The results for concept 1, 2 and 3 are summarized in **Error! Reference source not found.**, **Error! Reference source not found.**, and **Error! Reference source not found.**. The cell voltage sensitivity trend on COE is the same as that of cell utilization. At low voltage or utilization the system efficiency is low and the cost of fuel drives the COE high. At high voltage or utilizations the capital cost of the system drives up a high COE. The capital cost is high because more cells has to be added in order to compensate performance

² Source: Bmarkenergy.com

penalty as a result of high fuel utilization. In the case of high cell voltage, more cells have to be added because of decreased power density as a result of increased cell voltage. The minimum COE occurs approximately at a voltage of 0.75 V and a fuel utilization of 0.8 for concept 1 and 2. For concept 4, the minimum occurs at a fuel utilization of 0.75. However, the COE difference at fuel utilization of 0.75 or 0.8 is very small.

Figure 6 Cell Voltage and Fuel Utilization Sensitivity on COE for Concept 1

Figure 7 Cell Voltage and Fuel Utilization Sensitivity on COE for Concept 2

Figure 8 Cell Voltage and Fuel Utilization Sensitivity on COE for Concept 4

1.1.4 Concept Downselection and Summary

The system concept down-selection is based on system efficiency, reliability, cost, and the cost of electricity (COE). The analysis results are summarized in

	Concept 1	Concept 2	Concept 3	Concept 4
SOFC Power (kW)	563	581	222	495
Turbine Power (kW)	86	98	17	78
Parasitic Power (kW)	35	53	15	56
Net System Power (kW)	614	626	224	517
System Efficiency (%LHV)	63	62	64	67
Reliability	0.995	0.994	0.907	0.997
Availability	0.925	0.940	0.938	0.948
System Cost (\$/kW)	545	563	855	634
Cost of Electricity (c/kWH)	8.4	8.1	11.8	8.4

Table 11. Results are reported with minimum COE. Concept 2 has the lowest COE, but concept 4 has the highest system efficient.

Design for six sigma tools were used trading cost with efficiency. The system variables used to down select a system concept included systems efficiency, availability, and COE. As shown in Table 12, concept 4 received the highest score

	Concept 1	Concept 2	Concept 3	Concept 4
SOFC Power (kW)	563	581	222	495
Turbine Power (kW)	86	98	17	78
Parasitic Power (kW)	35	53	15	56
Net System Power (kW)	614	626	224	517

System Efficiency (%LHV)	63	62	64	67
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System Cost (\$/kW)	545	563	855	634
Cost of Electricity (c/kWh)	8.4	8.1	11.8	8.4

Table 11 Summary of System Results

DOE Hybrid Conceptual Design			System
<div> <div>Pass</div> <div>Marg</div> <div>Fail</div> </div> <div> <div>1</div> <div>0.5</div> <div>0</div> </div>			☐	☐	☐
			Efficiency	Availability	Cost of Electricity
Importance			1	1	1
USL					8
LSL			65	96	
Tolerance			2	2	0.5
Units			%	%	cents/kwh
No.	Concepts	Score			
☐ 4	Concept4	0.67	66	94.8	8.4
☐ 1	Concept1	0.33	63	92.5	8.4
☐ 2	Concept2	0.33	62	94	8.1
☐ 3	Concept3	0.17	64	93.8	11.8

Table 12 Design Trade Off Results

1.1.5 Part-Load performance analysis of the down-selected concept

Following the down selection of the concept system, its performance at partial loads was analyzed. The system performance model was amended for this purpose by inclusion of component performance maps or estimates, particular to the down-selected concept.

The following assumptions were made:

- (1) The Parallon75 recuperator is used. The recuperator performance maps available from Parallon75 test results were included in the model. The recuperator pressure drop curves were also included in the model.
- (2) The fuel cell model was adjusted to reflect the recent changes to the cell baseline.
- (3) The natural gas compressor efficiency was adjusted to lower numbers at partial loads, shown in Table 13. Since no performance maps are available for this component, conservative estimates were adopted.
- (4) The effectiveness of the steam generator, the reformer and the reformat heaters were set to constant values equal to those at the peak power. This is a conservative assumption as the effectiveness of a heat exchanger is likely to improve at lower flow rates due to a longer fluid residence time in the heat exchanger. This assumption is only valid at steady state and for a limited range of power loads because the bypass flows required to control heat exchanger outlet temperatures will likely be significant at unsteady and/or low power conditions. This assumption will be revised and updated during the detailed design phase when heat exchanger maps are available.
- (5) Pressure drops through the system were calculated at partial loads using Aspen's pipe model. The pipe geometry will be adjusted when the piping system design is completed during the detailed phase.
- (6) The SOFC pressure drop was conservatively assumed to be about 4.5% of the inlet pressure.

Fuel Flow, kg/sec	Pressure Ratio	Isentropic Efficiency	Mechanical Efficiency
0.0011	1.45	0.55	0.85
0.0037	1.99	0.60	0.85
0.0097	2.89	0.65	0.85
0.0158	4.19	0.70	0.85

Table 13 Natural Gas Compressor Operating Points

The performance model was used to analyze the system performance at the following Parallon75 shaft speeds: 35 krpm, 45 krpm, 55 krpm, and 64.76 krpm (the design point corresponding to peak power). The system power plot is shown on Figure 9.

Figure 9 System Power as a Function of the Parallon75 Shaft Speed

Note that the net power is a steep function of the shaft speed. This implies that the system is likely to operate at a speed close to the design speed as long as the power load does not significantly deviates from the peak load.

The microturbine generator power on Figure 9 is negative for shaft speeds below approximately 40 krpm (however, the net system power is still positive). This has implications for the start-up strategy, as the Parallon75 system will have to be operated in the turbocompressor mode rather than the microturbine mode. In other words, the generator would be operated as a motor to drive the compressor. It is recommended to operate the system above 40 krpm at all times in normal operations, if possible to avoid complicated and expensive controls and power electronics solutions.

It should be noted that the control scheme at part loads may have some differences from the peak load control scheme. Particularly, the recuperator inlet temperature requirement was found to have an effect on the average fuel cell temperature at part loads. Since the system operating pressure declines at lower speeds, the turbine pressure ratio is lower as well, which results in higher turbine outlet temperatures. Also, the compressor outlet temperatures are lower because of the lower compressor pressure ratio. This results in a lower fuel cell cathode inlet temperature.

The system efficiency plot is shown on Figure 10. Note that the system efficiency remains relatively flat for the net power higher than about 250 kW (or Parallon75 shaft speed above 50 krpm). The system power in this region is a strong function of two parameters: (1) the single-cell voltage (the higher the voltage, the higher the system efficiency); and (2) the SOFC specific power, i.e. the ratio of the SOFC power to the cooling air flow rate (the higher the specific power, the higher the efficiency). The voltage rises at lower power loads because the SOFC stack operates at a lower current

density. However, this rise is tempered by the decreasing fuel cell temperature and pressure at part load as was discussed above. On the other hand, the specific power tends to decrease with decreasing speed due to the recuperator temperature constraint. The combination of these effects results in a relatively flat efficiency line at higher part loads. However, the SOFC specific power decreases at a faster pace with decreasing system power, which eventually results in lower efficiencies at low part loads.

Figure 10 System Efficiency as a Function of Net System Power

1.2 Control System Development

The control system development focused on the following two areas: control system trade studies and conceptual control system design. The objective of the control system trade studies was to develop a better understanding of the controllability of the proposed full-scale system and investigate strategies for controlling the proposed system. The control system trade studies identify key controllability issues and are used to develop the conceptual control system design.

For these studies, controllability goals and objectives were established and the impact of the system design on controllability was evaluated. In addition, the trade studies helped prioritize the controllability issues based on system requirements. These trade studies provided feedback to the overall system design so that it could be adjusted to improve its potential operation.

1.2.1 Control System Trade Studies

Several control system trade studies were brainstormed and prioritized using a Six Sigma tool, the Quality Function Deployment (QFD). This QFD prioritized the trade studies by ranking each with respect to their influences on major system design changes, system cost, system performance, and system reliability, as well as the maturity of the tools needed to execute the trade study. The transient response trade study and bypass temperature control trade studies were significant enough to warrant investigation early in the program. The stack performance sensitivity trade study was not addressed due to the immaturity of the system design and the availability of the part-load model needed for the study.

A dynamic system model developed in Matlab/Simulink, was used to conduct the control system trade studies. This model was assembled using GE HPGS' Dynamic Fuel Cell Component Library. The model was verified by comparing its results with the results from the Aspen steady state system performance model of the full system. The full system model or an appropriate subset of the system model was used in each of the control system trade studies discussed below. The combined conceptual control system design was subsequently implemented and validated with the full system model.

1.2.1.1 Bypass Temperature Control

The objective of the bypass temperature control trade study is to determine the location and necessity of bypasses for temperature control during start-up, shutdown, normal operation, and load transitions. During this trade study, the conceptual control system design was revisited to explore different concepts for temperature control. Control strategies were evaluated based on:

- Number of bypass valves
- Location and temperature of bypass valves
- Impact on system capital cost
- Impact on system efficiency
- Controllability/Stability at setpoint
- Response time

The bypass temperature control study will be discussed first in terms of cathode temperature control and then in terms of anode temperature control. These trade studies were performed on the downselected system, system concept 4.

1.2.1.1.1 Cathode Temperature Control

There are two strategies that may provide adequate control of the cathode inlet temperature:

- A. Bypass hot side of recuperator
- B. Bypass cold side of recuperator

Both strategy A and strategy B have the same influence on the cathode inlet temperature. However, since strategy B requires a lower temperature it was selected as a candidate for further investigation.

Simulation studies were conducted using the full dynamic system model to evaluate the performance of the bypass control and tune the cathode inlet temperature controller. While the cathode inlet temperature does not perfectly track the setpoint, the performance of the control strategy is considered acceptable. One risk identified is that the cathode inlet temperature would drop below the lower specification limit if the airflow is greater than 10% of its design point flow rate for more than 30 minutes. However, violation of the lower specification limit may be avoided by decreasing the fuel utilization setpoint (thereby increasing the temperature of air exiting the air preheater) or actively

heating the stack. These risk mitigation strategies will be investigated further in subsequent phases of the program.

1.2.1.1.2 Anode Temperature Control

Four strategies were considered to provide adequate control of the anode inlet temperature:

- A. Bypass cold side of fuel preheater
- B. Bypass cold side of fuel processor preheater
- C. Bypass fuel to the fuel processor
- D. Bypass cold side of reformat preheater

These strategies were prioritized using a Six Sigma tool. The strategies were qualitatively evaluated against key elements that are considered critical to quality (CTQ's) for the control strategies. Strategy D is expected to have a strong influence on temperature but requires a high temperature valve. Strategy A does not require a high temperature valve but has a weak influence on anode inlet temperature. Strategy B has a weak influence on anode inlet temperature, but may be important for fuel processor and reformat composition control. Strategy C may have limited utility unless the stack's internal reforming capability is large and the temperature impact on the stack is explicitly considered in the control algorithms. Strategies A and D were selected as the candidates for further analysis.

Figure 11 Rank for Anode Temperature Control Strategies

A Matlab/Simulink dynamic subsystem model containing the fuel preheater and bypass, fuel processor, reformat preheater and bypass, and stack is used to analyze the two down selected control strategies. For both strategies, the bypass valve is used to control stack anode inlet temperature with a proportional plus integral (PI) controller.

For Strategy A, the fuel preheater bypass controller is turned on at the full power condition and is not able to maintain an anode inlet temperature with the bypass valve at 100% open. For Strategy D, the reformat preheater bypass controller is turned on at the full power condition and is also not able to maintain an anode inlet temperature, but by a less margin than strategy A. The bypass valve is also 100% open, but direct control of anode inlet temperature provides a more favorable response without the

process time delays and thermal lags of the fuel processor that exist with Strategy A. While neither strategy was able to maintain the desired setpoint, strategy A is considered more effective and therefore selected.

Simulation studies were subsequently conducted using the full dynamic system model to evaluate the performance of the reformat preheater bypass control (strategy A) and tune the anode inlet temperature controller.

1.2.1.2 Transient Response

This trade study focused on SOFC/microturbine hybrid system transient behavior. The results from this trade study will be used to develop load sharing control techniques for the microturbine-generator (MTG) and SOFC power systems during load transients and disturbances. A simplified hybrid system model was used to examine the major system components with the goal of qualitatively characterizing gross dynamic interactions. A novel control method has been developed that allows the hybrid system, with power being generated by both the SOFC and MTG, to follow a demanded, transient power setpoint and thus follow load changes in a controllable fashion.

The system concept evaluated in this trade study was based on Concept #1. Major components such as the SOFC, turbomachinery, recuperator, stack heat exchanger, and depleted fuel oxidizer were included in the dynamic system model.

The following recommendations can be made from the results of this trade study.

Minimize Internal Volumes - Internal volumes have a direct impact on controllability, especially during load transitions. The larger the internal volumes of the components and piping of the hybrid system, the more drastic the system responses are to a load change or disturbance. The system can transition faster between load demand setpoints when the volume is minimized. Larger volumes will decrease the transition speed away from the entitlement ramp rate.

Minimize Component Pressure Drops - Each component pressure drop contributes to losses in system efficiency. In addition, larger pressure drops leave less surge margin for the compressor and slow system response—especially during a load decrease. A reliable load transition necessitates adequate margin from surge as disturbances not modeled in this study may require additional margin to keep the compressor from surging.

Incorporate Blow-off Valving - As stated earlier, the volumes associated with the working fluid have a profound effect on the transient performance of the hybrid system. The largest of these will likely be associated with the pressure vessel that will house the SOFC stack. Large volumes slow the system response in that they act as a pressure capacitance, storing working fluid mass. When the system reduces power, the pressure inside the vessel does not decay instantaneously as the MTG speed is reduced. This causes the MTG Compressor performance running line to move towards the surge line. The surging phenomenon is caused by the inability of the compressor discharge fluid momentum to overcome the pressure gradient across the compressor wheel (without momentum, the fluid will tend to flow from high to low pressure, backwards over the compressor wheel—the opposite direction required to operate the hybrid system). In

order to reduce the likelihood of surging, the pressure in the SOFC stack containment vessel should be reduced quickly upon MTG deceleration, so that the running line remains away from the surge line. One way to increase the speed in which the pressure vessel volume discharges would be to incorporate a blow-off, or pressure relief valve. This would, if sized correctly keep the compressor away from surge during a load reduction event. The correct placement of the valve, either before or after the stack, should be studied further as a way to increase the transition speed, at least for load reduction.

Incorporate Bypasses - Because of the nature of the single shaft MTG, the turbine expander outlet temperature (TOT) decreases with increasing airflow and system power. This will serve, due to recuperation of the system exhaust heat, to lower the SOFC stack inlet temperature as airflow increases and increase stack inlet temperature as airflow decreases. Thus, there will be one particular airflow and associated enthalpy that will match the SOFC stack for a given current drawn, precluding any variation in stoichiometric ratios. For flexibility in operation (i.e. variation of air stoichiometry), bypasses around the stack and/or recuperator may need to be incorporated to provide an increased temperature control capability for off-design conditions.

There are some risks associated with the recommendations of the transient trade study. It must be understood that this was a simulation with many simplifying assumptions. The goal, in this case, was to identify trends that should be explored further in the Preliminary and Detailed designs. The quantitative values may not be representative of an actual system, but were used to qualitatively identify hybrid system performance.

One simplification that might have an effect on performance was the transient response of the fuel reformer. This study assumed that the thermodynamics of the fuel reformation process is accounted for, but does not specify the fashion (external, internal, steam, partial oxidation, etc.). A significant limitation might be imposed on the fuel utilization algorithm should the reformer play a critical limitation in transient response.

Another risk was the stack performance model. The current neural network model does not vary performance based on fuel composition—only on overall fuel utilization. It is well known that SOFC performance is dependent on fuel composition as well as fuel utilization. Only hydrogen fuel diluted with nitrogen has been mapped in the current SOFC performance model. Therefore, in addition to the thermodynamic response of the reformer stated above, the fuel composition entering the SOFC stack may impose transient effects not modeled in this study.

The heat loss from the system was assumed to only take place through the SOFC stack pressure vessel wall. Both the value and the location of the actual system heat rejection may have a profound effect on operation and thus might affect the actual transient and steady-state performance of the hybrid system.

1.2.1.3 Control System Trade Study Summary and Conclusions

Several areas were investigated as part of the control system trade studies to identify controllability issues that could influence the conceptual system design and to begin to

define how to control the proposed hybrid system. The most important conclusion drawn from these trade studies is that Concept 4 can be controlled.

The transient response trade study looked at the hybrid system and established a preliminary estimated load-change ramp rate of 1500 W/s if the system volumes are minimized. There were many assumptions that went into this estimate and additional analysis with more detailed component information will be needed in subsequent phases of the program to determine the true entitlement of the system. The control strategy developed in the transient response trade study showed that the temperature of the SOFC can be adequately controlled using current draw, with the power demand being met based on feedforward setpoint controls for airflow and some trimming with fuel utilization. This strategy shows great promise in minimizing the temperature fluctuations of the SOFC stack.

1.2.2 Conceptual Control System Design

The conceptual control system design for the full-scale system was updated with the results of the control system trade studies. The control structure underwent few changes as a result of this update.

The control structure is divided into two subgroups, the supervisory controls and the active controls. The supervisory controls perform the following functions:

- Translates load demand, user commands, and system status into setpoints for active controls
- Manage operating modes
 - Startup
 - Normal Operation
 - Shutdown
 - Emergency Stop
- Manages Built-In Test (BIT) and alarms

The active controls perform the following functions:

- Feedforward control for fast transitions between setpoints
- Feedback control for disturbance rejection and improved setpoint tracking

The feedforward control takes advantage of a priori information on the system operation to improve the transient response and stability of the system. An example of this would be the setting for the cathode temperature control valve being scheduled as a function of fuel cell power. Therefore when the fuel cell power is commanded by the controller, the appropriate valve position is also commanded. The feedback action of the control system would then fine-tune the position of the valve to achieve the exact cathode inlet temperature setpoint. The feedback portion of the control therefore provides disturbance rejection and robustness to variation in the process or components.

Due to the complexity of the system and its numerous constraints, the supervisory controls must be coordinated with the lower level control loops must also watch all component and system constraints while maintaining the demanded load.

The conceptual control system design has been implemented, tuned, and verified in the full dynamic system model. The supervisory controls have only been implemented to the point of coordinating lower level control loops. The built-in test and health monitoring functions will be added in subsequent phases of the program. The simulation results show the temperature and pressure constraints around the system are satisfied during rapid load changes of 500 W/s. The fuel utilization is not well controlled with load changes causing large deviations from the targeted value of 80%. This will need further investigation in subsequent phases of the program, but is probably due to the fuel flow control and pressure dynamics associated with the long path fuel must travel before reaching the stack. These oscillations can be limited or removed if the load change ramp rate were further limited.

2 TECHNICAL BARRIER RESOLUTION

2.1 High Temperature Heat Exchangers

The objective of this task is to develop, design, fabricate, and test a high temperature heat exchanger capable of operating with high-temperature exhaust gases to heat up air before it is introduced into the fuel cell stack. This report outlined the design and construction of a demonstration heat exchanger for testing and the analysis of a high temperature heat exchanger for pressure containment and creep life. The pressure containment and creep life analysis is supported with oxidation tests on coupons of representative high temperature materials presented below. Further, the performance testing of a prototypical heat exchanger is also presented below.

2.1.1 Heat Exchanger Design and Construction

The focus of this task was on the development of a high temperature heat exchanger to understand the feasibility of using such a heat exchanger in SOFC-hybrid systems. Consequently, the prototypical heat exchanger design was based on the requirements of the demonstration system. The design point performance requirements and the thermal design of the high temperature heat exchanger was therefore based on the performance requirements of the demonstration system. The specifics of this design are presented in Table 14 and the design is shown in Figure 12.

Figure 12 Heat Exchanger Design

Heat Exchanger Operation		
	Hot Fluid	Cold Fluid
Flow Rate (lb/min)	9.33	9.52
Inlet Temperature (°F)	308	908
Exit Temperature (°F)	849	430
Inlet Pressure (psia)	38.6	35.4
Pressure Drop (in H ₂ O)	0.524	0.234
Effectiveness	90%	80%
Heat Exchanger Core Construction		
Construction Type	Inconel 625, Plate-fin, Counterflow	
Core Dimensions (inches)	10 x 8.7 x 4.4	
No of passages (hot side/cold side)	19/19	
Core Weight (lb)	21.1	

Table 14 Thermal design of a high temperature heat exchanger

2.1.2 Heat Exchanger Analysis

Analysis was conducted to evaluate the life of the air preheat heat exchanger for the full-scale system at design point operating conditions. The analysis included structural analysis for pressure containment and creep life assessment for the heat exchanger core fins and manifolds. An example of the Finite Element computation field is shown in Figure 13.

The heat exchanger core and four manifolds were all found to have acceptable stress levels for pressure containment. For the gas manifolds, the stresses were calculated from nonlinear stress analysis. Negative stress margins confirmed local yielding in the manifold. However, the strain levels were found to be within acceptable limits.

The life to creep rupture was also estimated for the heat exchanger core and manifolds. The gas inlet manifold pan was found to be the most critical component in terms of susceptibility to creep rupture. However, this can be alleviated with the use of thicker material in the pan. The heat exchanger core design was found to have substantially higher life to creep if made with Inconel 625. Based on an estimated applied stress of 3.5 ksi in the hot passage of the core, the use of Inconel 625 fins yields a life of over 5,000 hours. The results for this analysis is summarized in Table 15.

Figure 13 Finite Element Models of the Heat Exchanger Fins and Manifold

	18-2 material	Inconel 625
Rupture, LMP [2]	40.5	50.1
Temperature	1650°F	1650°F
Time (hours)	0.16	5547

Table 15 Preliminary Creep Life Analysis Results

2.1.3 Oxidation Tests

Sixty coupon samples were prepared and placed in three furnaces at different temperatures: 732 °C (1350°F), 800 °C (1475°F), and 900 °C (1650°F). The tested coupons were made with Inconel 625 and Haynes 230 rods. These samples were exposed to the SOFC combustion environment having gas compositions as described in Table 16. Samples were scheduled for removal from the furnace at different exposure times ranging from 250 to 5000 hours. Micrographs showing the oxidation of these samples at exposure times up to 315 hour at the three temperatures is shown in Figure 14. Figure 15 shows similar micrographs for samples removed from the 900 °C (1650 °F) furnace after up to 3000 hours of exposure.

Gas	Approx Partial Pressure at 800 °C (1475 °F)
N ₂	0.7922
O ₂	0.1039
CO ₂	0.0346
H ₂ O	0.0693
H ₂	9.73E-11

Table 16 Approx Gas Composition Used for Coupon Testing

1350F, 280 hours
(oxide layer hardly seen)

1475F, 285 hours
(oxide layer hardly seen)

1650F, 315 hours

IN 625

HA 230

Figure 14 Oxidation Tests of Inconel and Haynes Coupons at Various Exposure Temperatures

IN 625 exposed at 1650F for 3041 hours Haynes 230 exposed at 1650F for 3041 hours

Figure 15 Burner Rig Photographs Showing Oxide Scale, Depleted Metal, and Oxide Penetration

The oxide penetration was measured on samples removed from the furnace and correlated with exposure time. The plots showing attack as a function of time for the three exposure temperatures is provided in Figure 16 through Figure 18.

Figure 16 Oxidation Data as Function of Time at 1350 °F

Figure 17 Oxidation Data as Function of Time at 1475 °F

Figure 18 Oxidation Data as Function of Time at 1650 °F

The depth of metal attack used in these plots includes the thickness of the oxide layer and the thickness of the depleted zone from which alloying elements are lost. It should be noted that general literature on oxidation usually is based on the oxide layer alone.

Assuming a required design life of 5000 hours for a high temperature heat exchanger constructed using 0.003-inch thick metal sheets, the data suggests both Inconel 625 and Haynes 230 are suitable for applications up to 730 °C (1350 °F). However, at 800 °C (1475 °F) the Inconel 625 will achieve the design life but Haynes 230 will not.

It should be noted the plots shown in Figure 16 through Figure 18 have been fitted with relatively short exposure data, approximately 3000 hours. Consequently, caution must be exercised in extrapolating this data to longer durations.

2.1.4 Heat Exchanger Tests

A series of performance tests were conducted with the prototypical heat exchanger, described above and shown in Figure 12. The heat exchanger test matrix, which includes pressure drop and thermal performance measurements at various design and off-design operations is shown in Table 17.

Test	Hot Flow Rate (lb/min)	Cold Flow Rate (lb/min)	Hot Inlet Temp (°F)	Cold Inlet Temp (°F)
1	8	8	900	75
2	9	8	900	75
3	10	8	900	75
4	11	8	900	75
5	8	9	900	75
6	9	9	900	75
7	10	9	900	75
8	11	9	900	75
9	8	10	900	75
10	9	10	900	75
11	10	10	900	75
12	11	10	900	75
13	8	11	900	75
14	9	11	900	75
15	10	11	900	75
16	11	11	900	75
17	10	10	1200	75
18	10	10	1400	75
Repeated data points				
19	8	10	900	75
20	9	10	900	75
21	10	10	900	75
22	11	10	900	75

Table 17 Test Matrix for Heat Exchanger Performance and Pressure Drop Test

The heat exchanger test set up is shown in Figure 19. The tests were conducted according to the following process. Flow rates, temperatures, and pressures were measured at the inlet and exit streams of the heat exchanger. At each operation set point the unit was allowed for the unit to achieve steady state conditions. Temperature and pressure measurements were taken at at-least three points within the duct cross-section to ensure bulk average temperatures were being recorded. At the outlet of the heat exchanger a mixer was used to measure the mixed flow temperature. The pressure measurement was measured before the mixer, at the outlet of the heat exchanger. All the data was recorded using Lab View data acquisition system. During the test, no anomalies occurred.

Front view

Left side view

Right side view

Thermocouples in heat exchanger skin

Figure 19 Pictures of Heat Exchanger Test Setup

Table 18 and Table 19 show the sample data taken during the performance test and the results computed from the raw data.

Hot Flow Rate (kg/s)	Cold Flow Rate (kg/s)	Hot Inlet Temp (deg K)	Cold Inlet Temp (deg K)	Hot Outlet Temp (deg K)	Cold Outlet Temp (deg K)	Pressure Drop Hot side (psig)	Pressure Drop Cold side (psig)
0.062	0.061	755.224	299.417	389.619	607.909	1.0	0.4
0.068	0.061	755.716	299.470	389.857	611.529	1.3	0.39
0.077	0.061	755.726	299.435	399.017	617.246	1.6	0.39
0.084	0.061	755.022	299.487	416.106	624.106	1.8	0.4

Table 18 Sample Heat Exchanger Performance Test Data

Hot Stream Mean Temp (deg K)	Cold Stream Mean Temp (deg K)	Qmax (W)	Effectiveness	U
572	454	28304	0.68	20.26
573	456	28220	0.68	20.41
577	458	28332	0.69	20.01
586	462	28308	0.71	19.08

Table 19 Sample Heat Exchanger Performance Results

Figure 20 Heat Exchanger Results

Figure 20 summarizes the performance results of the heat exchanger. The heat transfer and pressure drop of the heat exchanger are in good agreement with the design specification of the heat exchanger.

2.2 Pressurized SOFC

Single cell pressurized testing has been performed using two identical pressure vessels that have a stamped vessel rating for 60 psig. These vessels have been used to

conduct pressurized carbon deposition experiments and performance degradation or life tests.

2.2.1 Carbon Deposition Experiments

The use of nickel based anode has many advantages over others in terms of cell support structure (anode-supported cells), fast electrochemical oxidation kinetics (high exchange current density) and high activity toward reformation of fuels (internal reforming of hydrocarbon fuels). Especially, the high internal reforming activity provided by nickel based anodes allows highly efficient cell operation with reduced degree of external reformation, leading to a smaller external reformer. However, one concern with the use of hydrocarbon containing fuel streams is carbon deposition. The problem of carbon deposition is present whenever hydrocarbon containing fuel is employed as anode fuel irrespective of pre-reformation method (such as steam, auto-thermal or partial oxidation reformation).

Carbon deposition is generally observed over non-noble metal catalysts such as Ni, Co and Fe. The carbon deposition rate is impacted by many factors, such as the nature of the catalyst material, the chain length of hydrocarbon fuels, carbon/hydrogen/oxygen (C/H/O) elemental ratio of the fuel stream, and temperature and residence time. Also, the nature of carbon that is deposited (whether it is nanotube, whisker, polymeric, etc..) is dependent on the above mentioned reaction conditions. Generally, higher hydrocarbon fuels exhibit considerably higher rates of carbon deposition.

A fuel stream that shows high carbon element percentage in its C/H/O elemental ratio shows higher thermodynamic driving force toward carbon deposition. Therefore, raising the H and/or O ratio in C/H/O elemental ratio helps to prevent or circumvent environments that lead to carbon deposition. This directly translates to adding more steam or oxygen (or air) to the anode fuel stream for either external or internal reformation. During fuel cell operation, the generation of current is equivalent to transferring oxygen from the cathode stream to the anode stream through the electrolyte. Therefore, the fuel stream composition changes gradually as it travels along the anode flow field. The anode inlet of the cell corresponds to zero current generation and therefore, is more likely susceptible to carbon deposition. Contrarily, fuel stream leaving the anode flow field is relatively rich in O in its elemental C/H/O ratio and therefore, it is far less likely to suffer from carbon deposition, or in other words, is far away from carbon deposition boundary.

The thermodynamic prediction of carbon deposition at 1 atm in a cell running on a simple mixture of methane and steam is depicted in Figure 21. The elemental compositions of the individual anode fuel streams were calculated and plotted in elemental ternary diagram. Also, the composition of anode fuel stream that went through 80% fuel utilization was calculated and plotted on the diagram for comparison. The carbon boundary line shown in Figure 21 was reproduced from literature and it was found to agree well with our thermodynamic calculations. From Figure 21, anode fuel stream of S/C 1.5 (i.e. a simple mixture of H_2O and CH_4 in the ratio of 1.5) is located far below the carbon boundary at $800^\circ C$ both at the anode inlet and outlet locations, with

fuel utilization at 80%. This implies that all anode regions are safe from carbon deposition. Contrarily, the inlet region of S/C 0.1 anode fuel stream is located above the carbon boundary indicating potential for carbon deposition.

As shown in Figure 21, the elemental ratio of C/H/O changes along the anode flow field such that it can cross the carbon deposition boundary. Moreover, the accompanying internal reforming reaction, aside from current generation, contributes to local temperature changes since the internal reformation reaction is strongly endothermic. These factors make the prediction of carbon deposition in a working fuel cell complicated. In addition, kinetic factors are not taken into account in this thermodynamic prediction. Therefore, it is imperative to experimentally verify the conditions under which fuel cells can be run on hydrocarbon fuels (methane in this study) without the risk of carbon deposition.

Figure 21: Carbon deposition prediction in fuel cell anode

In this study, the potential degradation of cell performance associated with carbon deposition was experimentally evaluated for both ambient and elevated pressure (44.1 psig). Single cell modules with interconnects operating on a controlled reformate composition were employed and the effect of carbon deposition was observed based on cell performance and stability.

2.2.1.1 Testing Approach

In this task, cells were operated on a number of controlled anode fuel streams and the cell performance was monitored to learn the effect of carbon deposition on cell performance and stability. The cell performance test was conducted at a constant cell load (current) and cell voltage was measured. A drop in cell voltage over time is attributed to carbon deposition over the anode electrode if it exceeds the expected steady state degradation.

Experiments were conducted using 3-inch and 4 3/8-inch circular cells with a sealless radial flow field configuration. In this configuration the anode fuel stream and cathode air stream exhausts were mixed together and combusted inside the vessel after leaving

the cell. Once the cell was assembled, mounted and plumbed for gas feed, it was reduced at two different conditions (25% H₂ and 64% H₂, respectively) for a predetermined period of time. Then, the cell was conditioned for its maximum performance by applying an electric load for about 20 hours. Once the cell exhibited acceptable performance through cell conditioning, cell testing with a number of simulated anode fuel stream was initiated. At the start of the test the anode fuel stream was wet enough (corresponding to high steam-to-carbon ratio of fuel stream) to ensure that carbon deposition did not occur. During this period the performance and stability of the cell was monitored for a predetermined period of time. Once the cell demonstrated stable operation with the “wet” fuel stream, the anode stream was transitioned to a “drier” (corresponding to low steam-to-carbon ratio) fuel composition and the cell performance and stability was monitored. This process was repeated until cell voltage drop attributed to carbon deposition was observed. The voltage was allowed to continue dropping for a prescribed period of time or until the voltage got as low as 0.55V. This was to ensure that sufficient carbon was deposited for verification purposes. Finally, the cell was put under open circuit condition (with no electric load) and subsequently cooled down under a controlled environment to preserve any carbon deposits. Subsequently, carbon deposits were verified through microscopic surface analysis.

2.2.1.2 Test Set-Up

The testing for carbon deposition was designed to mimic a cell operating within the framework of a hybrid system configuration as shown in Figure 22. Upstream of the cell, methane and steam, with a predetermined steam-to-carbon ratio, is mixed with the anode exhaust recycle stream. Then, this mixture flows through the “Steam Reformer” and “SOFC, Anode” for external steam reformation and power generation, respectively. The anode exhaust out of “SOFC, Anode” is then split into two directions. Part of the anode exhaust stream is directed out of the system for further processing and the other part of the anode exhaust stream is recycled back to the inlet fuel. A closer look at the system configuration reveals that the composition of the anode fuel stream that enters the “SOFC, Anode” is determined by many system design factors, such as steam-to-carbon ratio in the inlet fuel mixture, the steam reformer size and temperature, the cell current density and fuel utilization, and the anode exhaust recycle ratio. In determining system representative anode fuel stream compositions, cell current density and fuel utilization were set at 400mA/cm² and 80%, respectively. Steam-to-carbon ratio was found to be a major factor that determines the anode fuel stream composition and representative values of steam reformer size/temperature and anode exhaust recycle ratio were used in the calculation of anode fuel stream composition.

Figure 22: Configuration of fuel cell within the framework of hybrid system design

The anode fuel stream composition was computed using the process described above and calculated using process modeling software and is summarized in Table 20.

S/C	3.0	2.0	1.5	1.0	0.7	0.6	0.5	0.4	0.3	0.2	0.1	0.001
H ₂	26.3%	28.6%	30.0%	31.5%	32.0%	32.2%	32.0%	31.9%	31.1%	30.0%	26.6%	4.0%
O ₂	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%
H ₂ O	36.3%	31.4%	27.6%	22.2%	17.8%	15.7%	13.8%	11.3%	8.5%	5.0%	1.2%	0.0%
N ₂	0.0%	0.1%	0.1%	0.1%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%
CO ₂	24.0%	21.7%	19.7%	16.7%	13.9%	12.6%	11.3%	9.4%	7.4%	4.5%	1.0%	0.0%
CH ₄	6.0%	10.0%	13.6%	19.4%	25.3%	28.0%	31.3%	35.2%	40.6%	47.4%	58.3%	92.6%
CO	7.3%	8.3%	9.1%	10.2%	11.0%	11.4%	11.6%	12.1%	12.4%	13.0%	12.9%	0.2%
C ₂ H ₆												2.8%
Total	100.0%	100.0%	100.0%	100.0%	100.0%	100.0%	100.0%	100.0%	100.0%	100.0%	100.0%	99.7%

Table 20: Anode fuel stream composition at different steam-to-carbon ratios

Overall, the anode fuel stream composition is enriched with methane, while the steam content drops as steam-to-carbon ratio decreases. The elemental composition corresponding to individual steam-to-carbon ratios was calculated based on the anode fuel stream composition in Table 20 and is listed in Table 21.

S/C	3.0	2.0	1.5	1.0	0.7	0.6	0.5	0.4	0.3	0.2	0.1	0.001
C	13.4%	14.1%	14.7%	15.6%	16.3%	16.7%	17.0%	17.4%	17.9%	18.5%	19.1%	19.9%
H	53.7%	56.5%	58.9%	62.3%	65.3%	66.6%	68.0%	69.6%	71.6%	73.8%	76.6%	80.0%
O	32.9%	29.3%	26.4%	22.1%	18.4%	16.7%	15.0%	13.0%	10.5%	7.7%	4.2%	0.0%
Total	100.0%	100.0%	100.0%	100.0%	100.0%	100.0%	100.0%	100.0%	100.0%	100.0%	100.0%	100.0%

Table 21: Elemental composition of anode fuel stream at different steam-to-carbon ratios

A schematic of test set-up for pressurized carbon deposition experiment is shown below in Figure 23.

Figure 23: Schematic of test set-up for pressurized carbon deposition experiments

A single cell module was housed inside a cylindrical pressure vessel, equipped with an electrical furnace for heating up of the cell module. Programmable temperature controllers were employed for precise and programmed temperature control of the vessel. Mass flow controllers were used for precise and controlled flow rates of individual fuel gases. These individual gas flows were mixed and then introduced to the anode flow field. The anode fuel stream was preheated as it travels through the fuel preheating line in the vessel. A separate water line from the liquid water pump entered the vessel and steam was generated as it traveled through the coiled water preheating line before entering the anode fuel preheating line as shown in Figure 23. Radial seal-less single cell SOFC modules of 3-inch diameter were used for ambient pressure experiments while 4 3/8-inch radial seal-less single cell SOFC modules were used for pressurized (44.1 psig) experiments. With the radial seal-less cell module, excess fuel in the anode exhaust stream that had not been consumed at the anode was combusted along the circumference of the cell. This helped to maintain the overall cell temperature. For pressurized tests, a back pressure regulator was employed to pressurize the cell. For safety purposes, the pressure vessel was equipped with a pressure relief valve. The pressure relief valve opened up to prevent any excessive and accidental pressure build-up inside the vessel. The test set-up was interfaced with a digital data acquisition system that logged time stamped information such as individual gas flows, temperature, cell voltage and current. A pressure transducer was installed upstream of the fuel preheating line outside the furnace and pressure vessel and the pressure in the fuel line was monitored to ensure there is no flow blockage.

2.2.1.3 Test Results

2.2.1.3.1 Ambient Pressure Experiments

Altogether, 10 cells were assembled and tested for carbon deposition experiments at ambient pressure during this reporting period. Initial tests were used for verifying and troubleshooting the test setup, and for rough mapping of the carbon deposition regions. Later tests were used to pinpoint the carbon deposition regions.

Figure 24 shows cell voltage and upstream fuel line pressure plotted as a function of time for different anode fuel stream compositions. As specified above, the cell was first reduced and then conditioned at 457 mA/cm² for 20 hours. This conditioning procedure increased cell voltage from 0.617V to 0.670V. Following the cell conditioning, the anode fuel stream was transitioned to a steam-to-carbon ratio of 1.0, with fuel composition as specified in Table 20. In previous cell tests, stable cell performance on steam-to-carbon ratio of 1.5 (S/C 1.5) had been demonstrated several times and so it was considered safe to start at this condition to establish stable operation without carbon deposition after cell conditioning. The total anode fuel flow rate after the cell conditioning was controlled throughout the whole test period to maintain fuel utilization of 28.6% at current density of 286mA/cm² at an operating cell voltage above 0.7V.

Figure 24: Carbon deposition experiment result at ambient pressure

As shown in Figure 24 the cell maintained stable cell voltage of 0.77 ± 0.015 V for 24 hours at anode fuel stream S/C of 1.0. Fluctuations in cell voltage was observed

generally on cells running on reformat fuel streams containing steam. These fluctuations were caused by fluctuations in the steam generation for the test. The degree of voltage fluctuation was experimentally observed to be higher as the anode fuel stream was enriched with steam.

After reaching 24 hours of cell operation on S/C 1.0 anode fuel stream, anode fuel stream was transitioned to S/C 0.5. At this condition the cell showed very slow and gradual voltage drop without increase in fuel line pressure, indicating carbon deposition on the cell. The cell voltage dropped from 0.777V to 0.769V over 24 hours (voltage drop rate 0.333mV/hr). The cell was subsequently operated for a second 24 hours duration on S/C 0.5 anode fuel stream. The cell voltage continued to drop gradually (from 0.769V to 0.743V) but at an increased rate of 1.08mV/hr. As the cell voltage drop rate measured was fairly low and the cell voltage remained above 0.7V, even after 48 hours of cell operation on S/C 0.5, it was decided to reduce the steam content further to S/C 0.4 for an extended period of time (another 48 hours) as shown in Figure 24. The cell voltage drop rate on S/C 0.4 anode fuel stream for 48 hours was measured to be approximately 0.90mV/hr and this rate was very close to the one measured on S/C 0.5 anode fuel stream. A slight increase in the fuel line pressure was observed (from 14.9 psia to 15.4 psia) while the cell was operated on S/C 0.4 anode fuel stream. The fuel line pressure rise was later found to be due to the clogging of the dry fuel line upstream of the steam mixing point due to carbon deposition. The amount of carbon left inside the fuel preheating line was measured and assuming that the carbon deposition inside the fuel preheating line occurred for 48 hours, it corresponded to less 2% of methane flow rate in S/C 0.5 anode fuel stream. As the fuel utilization level in this experiment was low (28.6%), methane consumed for carbon deposition in the fuel preheating line was considered to have very minimal effect on cell voltage drop observed on S/C 0.5 and 0.4 anode fuel stream.

The cell was cooled down after reaching 48 hours of operation on S/C 0.4 anode fuel stream and the cell assembly was removed from cell stand and the cell was tore open from the endplates to inspect for any changes in cell structure or to locate any carbon deposits that may have been formed over the anode flow field or the cell anode surface.

Figure 25 shows an image of cell anode after the conclusion of the ambient test described above. It appears that the anode is in a reduced state (gray color of metallic nickel) and no carbon species were visually found. The small black powder along the edge of anode cell pieces in Figure 25 is cathode material.

Figure 25: Image of cell anode taken out after carbon deposition experiment at ambient pressure

The cell anode was further inspected using SEM-EDX surface analysis techniques. Various points across the anode layer were inspected. Figure 26 shows a typical SEM-EDX spectrum obtained from the cell anode of Figure 25. A small C peak is present in the spectrum on both YSZ (left) and Ni (right) particles. Anode with visible carbon deposit is reported to yield a C peak that is comparable in size to the Zr peak. Considering the little C peak on SEM-EDX spectrum in Figure 26, the carbon deposition was considered to occur fairly slowly on both S/C 0.5 and 0.4 anode fuel streams and that slow carbon deposition rate resulted in low cell voltage drop rate as shown in Figure 24.

Figure 26: SEM-EDX spectrum taken from anode support layer after carbon deposition experiment at ambient pressure

Thermodynamic calculations predict that the carbon deposition boundary lies between S/C 0.6 and S/C 0.7 anode fuel stream composition of Table 20 at ambient pressure and 800°C. The results of the experiment (carbon deposition at S/C of 0.5 and 0.4) are consistent with the thermodynamic calculations.

2.2.1.3.2 Elevated Pressure (44.1 psig) Experiments

Two pressurized tests were conducted on 4-3/8 inch circular seal-less SOFC single-cell modules. The assembled cell went through cell reduction and cell conditioning before pressurization. The cell voltage and fuel line pressure data collected during this test is shown in Figure 27.

Figure 27: Carbon deposition experiment result at 44.1 psig pressure (numbers in parentheses refer to flow conditions detailed in Table 22)

Anode Fuel Streams employed in Carbon Deposition Test at Elevated Pressure	Current Density (mA/cm ²)	Fuel Utilization (%)	Carbon Deposition Predicted	
			Open Circuit	Under Load
(1) Cell Conditioning, H ₂ 0.64 SLPM + N ₂ 0.36 SLPM	370	29.4		
(2) H ₂ 0.51 SLPM + N ₂ 0.29 SLPM	286	28.6		
(3) H ₂ 0.51 SLPM + N ₂ 0.11 SLPM + H ₂ O 0.177 SLPM	286	28.6		
(4) S/C 2.0 + N ₂ 0.363 SLPM,	286	28.6	NO	NO
(5) S/C 1.0 + N ₂ 0.598 SLPM	286	28.6	NO	NO
(6) S/C 0.7 + N ₂ 0.672 SLPM	286	28.6	NO	NO
(7) S/C 0.5 + N ₂ 0.723 SLPM	286	28.6	YES	NO
(8) S/C 0.3 + N ₂ 0.777 SLPM	286	28.6	YES	NO
(9) (S/C 0.5)x2.2 + N ₂ 0.362 SLPM	286	13.0	YES	NO

Table 22: Details of flow conditions used in pressurized carbon deposition experiment (see Figure 27)

The vessel was pressurized while the cell was operating on a hydrogen/nitrogen fuel stream. The anode fuel flow rate was maintained at 28.6% fuel utilization at 286mA/cm² current density. Pressurization of the vessel from 0 psig to 44.1 psig raised cell voltage from 0.713V to 0.764V.

After 24 hours of stable cell operation at these conditions, steam injection was initiated to understand the effect of steam on cell performance and stability at pressurized conditions. Liquid water flow rate was controlled to achieve 22.2% steam content by replacing part of the nitrogen flow in the anode fuel stream. The 22.2% steam content

was the same as that of S/C 1.0 in Table 20 and considered the representative steam content. A fluctuation in cell voltage ($\pm 15\text{mV}$) was observed for the same reasons discussed above. The total anode flow rate was subsequently increased to 1.024 SLPM without changing the mole percentage of hydrogen, nitrogen, and the steam in the anode fuel stream (64% H_2 , 13.8% N_2 and 22.2% H_2O). The flow rate was increased because at high pressures the linear velocity is decreased, making it difficult to achieve local flow uniformity and adequate velocity at the exit to prevent back diffusion. Subsequently, all testing was performed with an anode flow rate of 1.024 SLPM by adding nitrogen as needed, to maintain adequate linear velocity. The increase in total anode flow rate raised cell voltage by 0.015V (from 0.73V to 0.745V) through reduced fuel utilization (from 28.6% to 22%).

After 24 hours of stable cell operation with a wet (22.2% H_2O) hydrogen/nitrogen anode fuel stream, the anode fuel stream was transitioned to a S/C 2.0 reformat stream. The cell voltage remained stable at approximately 0.705V for 24 hours of operation and showed no measurable sign of cell voltage degradation. Cell operation was continued by transitioning the anode fuel stream to lower S/C ratios (S/C 1.0, S/C 0.7, S/C 0.5 and S/C 0.3), as shown in Figure 27, for 24 hours each and after confirming stable cell operation at each S/C setting. The cell logged a total of 120 hours of operation.

Followed by the stable cell operation on S/C 0.3 reformat stream, the anode fuel stream was transitioned to S/C 0.5 reformat stream with reduced nitrogen flow rate. To compensate for the reduced nitrogen flow, but maintaining a total anode flow rate of 1.024 SLPM, the S/C 0.5 reformat stream flow rate was raised by 120%, which resulted in a lowered fuel utilization from 28.6% to 13.0%. It should be noted that the reduction of nitrogen flow rate and increase in S/C 0.5 reformat stream flow rate increased the partial pressure of individual reactant gases in the reformat stream. The initial cell voltage measured with S/C 0.5 reformat stream was higher (0.773V) than the preceding S/C 0.5 reformat stream test (0.717V). This higher cell voltage was considered to result from the increased reformat stream flow rate (by 120% compared to the preceding one) and the resultant low fuel utilization (13.0% vs 28.6%).

Once the cell voltage reached 0.773V, then it started to drop at the rate of approximately 4.1mV/hr, indicating carbon deposition. The cell voltage drop was accompanied by a rise in fuel line pressure. The fuel line pressure rise was moderate (less than 2 psi) and was not considered high enough to substantially impact anode flow. Cell operation was terminated under a controlled environment after the cell voltage dropped by approximately 0.1V. The cell module was cooled down under a controlled environment to conserve any carbon deposits that may have been formed.

No carbon deposits were found in the fuel preheating line upstream of the cell anode. Therefore, it is inferred that the gradual rise in fuel line pressure resulted from the accumulation of carbon deposits in the anode flow field, as seen in the figure.

SEM-EDX analysis was conducted on the cell. This revealed a morphology of carbon deposit over the anode surface that was found to be bulky carbon fibers and aggregated spherical carbon, as shown in Figure 28.

Figure 28: SEM image of carbon deposits over anode surface

The profile of carbon deposits across the anode layer was further investigated to determine whether carbon deposition occurred inside the anode layer. Figure 29 shows an optical microscopic image of the test cell cross-section in the regions where carbon deposits were observed at the anode surface. Dark black spots were found on certain areas in the active (electrochemical) anode layer. There were very few signs of carbon deposition in the support anode layer.

The carbon deposit profile was also checked at the electrolyte layer, active anode layer, and four other locations in the support anode layer. Figure 30 shows the location and the SEM-EDX spectra results. The C peaks in Figure 30 increase as the location moves toward the electrolyte layer. The active anode layer has the highest C peak. No noticeable C peak was found on the electrolyte layer. This observation does not support the general understanding that the anode inner layer is more enriched with H_2O and less enriched with methane compared to the anode outer layer. It is probable that carbon deposits closer to the anode/gas interface might have been removed through reaction with hydrogen during the cool down period of the test since low hydrogen flow was bled into the nitrogen stream during this period.

Figure 29: Optical microscopic image of test cell cross-section

Figure 30: SEM-EDX spectrum across test cell cross-section

2.2.1.4 Discussions and General Conclusions

Along with the ambient and pressurized carbon deposition experiments as shown in Figure 24 and Figure 27, respectively, thermodynamic calculations were performed to understand the operating conditions that are likely to promote carbon deposition in an operating cell. Calculations for ambient pressure operation indicate that S/C 1.0 anode

fuel stream in Figure 24 does not deposit carbon under open circuit operating conditions (i.e. with no electric load drawn from the cell) while S/C 0.5 does deposit carbon. When oxygen transfer via current generation from the cell is accounted for, S/C 0.5 anode fuel stream is not predicted to deposit carbon.

Thermodynamic calculations at conditions employed in Figure 27 (elevated pressure) predicts that carbon boundary lies between “(6) S/C 0.7 + N₂ 0.672 SLPM” and “(7) S/C 0.5 + N₂ 0.723 SLPM” anode fuel streams. Current generation on “(7) S/C 0.5 + N₂ 0.723 SLPM” anode fuel stream at fuel utilization of 28.6% was shown to add enough oxygen so that no carbon deposition is predicted at the anode outlet region. In Figure 27, cell voltage drop was clearly observed with “(9) (S/C 0.5)x2.2 + N₂ 0.362 SLPM” anode fuel stream which is also predicted to deposit carbon under open circuit.

In general, the experimental observation of carbon deposition was in line with thermodynamic calculations at open circuit conditions. The higher the steam-to-carbon ratio in anode fuel stream of Table 20 or the higher the O content in C/H/O elemental composition in anode fuel stream of Table 21, the less likely carbon deposition would occur. The experimental results show that cell operation on anode fuel of S/C 1.0 reformat stream (both ambient and pressurized) was stable without any sign of cell voltage drop. This observation suggests that anode fuel compositions corresponding to S/C 1.0 or higher are in a safe operating zone.

Thermodynamically, carbon deposition is favored at pressurized condition. For example, the anode fuel stream composition of S/C 0.7 in Table 20 is predicted to deposit carbon at 3 atm (29.4 psig), but not at 2 atm (14.7 psig). However, its pressure dependence is considered to be very weak and, therefore, the carbon boundary in elemental C/H/O diagram of Figure 1 may not be moved significantly with pressurization. In a working cell, kinetic parameters (such as mass transfer, rate constants, etc) also play important roles along with thermodynamic driving force toward carbon deposition. All these factors are reflected in experimental observations and the contribution of individual factors would be gained more clearly through extensive work over wide ranges of experimental conditions.

2.2.2 Life Test

2.2.2.1 Task Objectives

The objective of this task is to determine the impact of pressurized operation on the SOFC performance stability. Prior to this task, most of the SOFC development and evaluation was focused on ambient conditions. No long-term data was available on planar SOFC's operated at elevated pressures. The goal of this task is to characterize the SOFC performance stability under pressurized operation and to compare it to that of ambient pressure operation.

2.2.2.2 General Approaches

In the early stages of this program, due to limitations in available test stands, SOFC performance had been evaluated at 2 atm with 3-inch cells. The preliminary data

indicated that SOFC performance appeared to decay faster than that under ambient conditions. In this reporting period, several long-term tests were completed. Most of the tests used 4 3/8-inch circular cells with radial sealess design and were conducted at nominal 800°C with 64% H_2 /36% N_2 as fuel and air as oxidant. These cells consisted of YSZ electrolyte separating Ni/YSZ anode and LSM/YSZ cathode. Commercially available ferritic alloy was used as interconnect. To reduce interference of fuel utilization, these tests were conducted under moderate fuel utilization (~40%). During tests, the cells were operated under a given current density and pressure, and the cell voltage was monitored over the testing period to evaluate its performance. Cell area specific resistance (ASR) was also estimated from the cell open circuit voltage, operating voltage, and current density. In some tests, AC impedance was also taken before and after the test to evaluate the change in cell impedance. Post test analysis was conducted to characterize the cell microstructure, oxide scale on metallic interconnect, and chromium contamination to electrodes.

2.2.2.3 Fuel Cell Test Results

Figure 31 shows the performance (cell voltage and ASR) of Cell 122 as a function of time under test conditions. After cell reduction and initial polarization, the cell was held at 0.435 A/cm² and 1 atm (ambient pressure). The cell performance continued rising in the first 200 hours and it appeared to be steady between 300-500 hours. The initial performance increase with time is typical for these cells. However, it appears to take longer for the performance to stabilize or start to decay. This process has not been well understood and it is often referred to as “conditioning” of the cell, especially the electrodes. Since the “conditioning” process was long, a higher current density (0.473 vs 0.435 A/cm²) was used to accelerate the process. However, the cell performance started to decay rapidly, even when the cell returned to the original current density (0.435 A/cm²).

Nevertheless, the cell module was pressurized to 4 atm and held at 0.435 A/cm² for further testing. Again, the cell decayed and the degradation rate appeared to be much faster (voltage loss of ~170 mV within 20 hours) than that at 0.435 A/cm² and ambient pressure. Afterward, the cell performance could not be sustained at a lower current density (0.323 A/cm²). At this point, the cell test was terminated. It seemed that the cell was damaged during the current transition period even though the exact causes remained unclear.

Figure 31: Performance of cell 122 at 800°C with 64% hydrogen balance nitrogen

In parallel to Cell 122, Cell 110 was also tested with the same objective. Figure 32 shows the performance with time of Cell 110. Similar to the Cell 122, the performance of Cell 110 improved over time in the first 640 hours, showing the cell voltage increase and ASR decrease under a fixed current density of $0.3\text{A}/\text{cm}^2$. After approximately 650 hours, the cell was pressurized to 4 atm and held at current density of $0.379\text{A}/\text{cm}^2$ for about 100 hours before the furnace malfunctioned. The furnace cooled down unexpectedly while the electronic load continued to draw current from the cell, which damaged the cell. In the approximately 100 hours testing period under 4 atm, the cell voltage appeared steady or slightly decreased in comparison to that observed at ambient pressure (Figure 33).

Figure 32: Performance of cell 110 at 800°C

Figure 33: Performance data of cell 110 showing the transit between ambient pressure and pressurized conditions

Since the module performance increased with time in the initial testing period and this conditioning process under ambient pressure was relatively slow as indicated in Cell 110 and Cell 122 tests, a test plan was devised to obtain the performance stability trend under the pressurized conditions first and then switch to ambient conditions. Figure 34 shows the test history of Cell 104. It starts with a 4 atm hold, followed by several steps: at 1 atm, 4 atm, 3 atm, 1 atm and then back at 4 atm. The total duration of this test is more than 1000 hours. In the late stage of this test (from ~900 hours), the pressure fluctuated as much as 5 psi due to a malfunction of the back pressure controller.

During the whole testing period, the voltage appeared somewhat noisy. Despite this noise, the data clearly shows that pressure had a significant impact to the cell performance stability. Cell performance decayed under pressurized conditions in comparison to cell performance improvement (or “conditioning”) at 1 atm in the testing period. Close examination of the test data and conditions indicated some of voltage fluctuation arose from the temperature fluctuation of the furnace. The error introduced by the temperature fluctuation was corrected with a transfer function of voltage and temperature obtained in the temperature range between 790 and 815°C. The temperature adjusted data is shown in Figure 35.

Figure 34: Testing history of cell 104 showing performance and temperature fluctuation

Figure 35: Performance of cell 104 after correction of temperature effect

As shown in Figure 35, the cell voltage decayed under pressurized conditions while the cell performance increased with time once it was switched back to ambient pressure. These results clearly demonstrate the pressure impact to the cell performance stability with time. Based on the data shown in Figure 35, the cell ASR (which includes both ohmic and non-ohmic losses) increased by $\sim 350 \text{ mohm-cm}^2$ after being tested for about 1165 hours.

The AC impedance was also measured on Cell 104 under open circuit conditions at ambient pressure before and after the test. The impedance data (Figure 36) indicates that ohmic resistance of the module increased by $\sim 182 \text{ mohm-cm}^2$ (from $\sim 274 \text{ mohm-cm}^2$ at the beginning of the test to $\sim 456 \text{ mohm-cm}^2$ at the end of the test after 1165 hours). This ohmic resistance increase is about half of the total ASR increase ($\sim 350 \text{ mohm-cm}^2$) observed in the performance evaluation.

The impedance analysis also shows that the electrode polarization increases. While the electrode polarization can be estimated from the impedance analysis in principle, the data shown in Figure 36 can only be used as reference because it represents the conditions under open circuit voltage (OCV) at ambient pressure. Due to instrument limitations, AC impedance could not be carried out with current load.

Figure 36: AC impedance taken at 800°C under OCV before and after test

2.2.2.4 Post Test Analysis

Post-test analysis on Cell 104 (Figure 37) did not reveal micro structural anomalies in the cell cathode, electrolyte, and anode in comparison to the cells tested under ambient pressure. The oxide scale was about 2 micron thick, for both the cathode and anode side interconnect (Figure 38). The oxide thickness is similar to those cells tested at ambient pressure for a comparable duration. Silicon-containing material was also detected under the chromium-rich oxide scale. The silicon is believed to be an impurity in the interconnect metal.

Figure 37: Fracture surface of cell 104 after being tested for more than 1000 hours under pressure

Figure 38: Cross-sections of cell 104 showing oxide growth at both cathode and anode side interconnect

Using an EDX line scan, Chromium was detected between the LSM/YSZ cathode and the metallic interconnect. A slight chromium pile up at the electrolyte/cathode interface was also noticed in the analysis. It appears that the chromium species of metallic interconnect transported and interacted with the cathode materials.

Detailed composition analysis was also conducted on Cell RJ019, which was tested for about 1000 hours. Most of the test time of this cell was at 2 atm pressure. Electron probe micro-analysis (EPMA) was performed in the areas under the interconnect fin (direct current path) as well as the areas open to the gas channel. There are several interesting observations from this data:

- The local chromium pileup at the cathode/electrolyte is clearly seen. In the center and middle portions of the cells, ~1.5-1.6% Cr is observed at the cathode/electrolyte interface. Lower Cr (~0.5%) is present in the edge areas of the cell.
- Chromium presence in a majority of the cathode is low except at the two interface areas.
- Higher chromium (3-5%) is present in the cathode

The profile looks similar to that under the fin area with two exceptions:

- Chromium presence is lower in the gas channel area compared to the area under fin
- At the cathode/electrolyte interfaces, Cr is less than 0.7% and there is little difference from cell center to edge

It is reasonable to believe that the local current densities are higher under the fin areas than the neighboring areas under the gas channels. Since both fuel and air flow is radial from the center to the edge, current density at the center is likely to be higher than that at the edge because of the fuel concentration difference (assuming the temperature is similar). In conjunction with the chromium profile observation, it seems that the Cr

transport and interaction with cathode materials may be current related. The higher current density may favor Cr deposition.

2.2.2.5 Hypothesis on Degradation Mechanisms

Major potential causes to cell performance decay include oxidation of metal interconnect, Cr poisoning to electrodes, interface resistance increase, electrode microstructure changes, back diffusion/leakage, and/or electrode conditioning process.

For oxide growth, the parabolic kinetics gives

$$x^2 = k_p t$$

or

$$x = k_p^{0.5} t^{0.5}$$

where x is the thickness of the oxide scale, k_p is the parabolic rate constant (thickness based) and t is the oxidation time.

For Cr-containing ferritic alloys, the parabolic rate constant depends on the self-diffusion coefficient of Cr and self-diffusion of oxygen. It has been found that the Cr outward diffusion to the oxide scale surface dominates the oxidation kinetics. At a given elevated temperature, the oxygen partial pressure, P_{O_2} dependence of the parabolic rate constant k_p can be estimated with diffusion coefficient of Cr ion in Cr_2O_3

$$k_p \propto D_{Cr} \propto P_{O_2}^{3/16}$$

Thus, the oxide growth thickness based on parabolic kinetics can be estimated as

$$x \propto P_{O_2}^{3/32} t^{0.5}$$

Assuming this relationship is valid for the testing conditions, the thickness of oxide scale in oxidizing environment is expected to increase by approximately 7%, 11%, 14% when pressure increases from 1 to 2, 3, and 4 atm, respectively. As indicated in Figure 38, the oxide scale was about 2 micron after 1165-hour test, which is not much different from those samples tested for similar duration under ambient pressure. It is inconclusive on whether or not the oxide scale grows faster under higher pressures due to the following reasons:

- The resolution is low for moderate growth rate increase under pressures tested, based on parabolic kinetics and the thickness measurements
- The thickness of the oxide scale observed is not representative since the oxide layer can also react with oxygen and water vapor to form vapor species

Another possible degradation mechanism is the chromium poisoning of the cathode, as observed in post-test analysis. This could be electrochemical reaction driven and/or chemical reaction driven. Considering chromium oxide and vapor formation reactions:

Reaction 1 describes the chromium oxide scale formation on Cr-containing ferritic alloys, which will increase the oxide scale thickness as described with parabolic oxidation kinetics. Reaction 2 and 3 describe the possible chromium vapor formation, which could poison the cathode. Under the test conditions, the most possible chromium species may be chromium oxyhydroxide from reaction 3 because of the inevitable presence of H_2O impurity in the air used. The exact mechanism of the chromium contamination through chromium vapor species is unclear at this stage. However, higher operating pressure favors higher chromium-species pressures. Thus, the driving force for chromium vapor species diffusion and subsequent reaction with active cathode materials will be higher under higher pressure.

Other possible degradation mechanisms could also come in play, such as interface resistance increase, electrode microstructure changes, back diffusion/leakage, and/or electrode conditioning process. Exploration of those reaction mechanisms is beyond the scope of this task.

2.2.2.6 Summary

In summary, a number of SOFC module tests have been completed to evaluate the pressure impact to performance stability. The results clearly demonstrate that the operating pressure accelerates the performance degradation. The potential causes to the performance decay include oxidation of the metal interconnect, chromium poisoning to electrodes, interface resistance increase, electrode microstructure changes, back diffusion/leakage, and/or electrode conditioning process. The dominant degradation mechanisms remain unclear. However, current density appears to have an enhanced effect on chromium interaction with cathode material. Both interconnect oxidation and Cr transport to cathode were evident based on post-test analysis. To reduce the degradation related to metallic interconnect, interconnect materials have to be improved or protected with a coating. Future work is needed to understand the degradation mechanisms and the impact of pressure on the electrode conditioning processes.

2.3 Cell Scalability

2.3.1 Introduction

The objective of this effort was to demonstrate cell area scaleup based on the tape calendaring method. Both theoretical analysis and experimental work was conducted. The theoretical analysis involved literature search to determine fundamental limitations, if any, on the size that can be sintered. Experimental work as well as theoretical studies were performed to identify the key process parameters that affect cell scalability. These key factors were optimized to demonstrate process scaleup to large area cells (> 10" diameter). Testing was conducted on representative large area cells to determine their performance under typical hybrid operating conditions.

2.3.2 Background

The tape-calendering method was used to fabricate the cells. Figure 39 shows the basic tape-calendering sequence for producing a thin electrolyte supported on an anode support. In this process, the starting materials (YSZ electrolyte, NiO/YSZ anode support electrode, organic binders, and plasticizers) are mixed in a high-shear mixer. The heat generated by the mixing process softens the organics to form a homogeneous plastic mass. Each plastic mass is then rolled to produce a flexible sheet on the order of 0.25 to 2.5 mm thick. The electrolyte and anode tapes are then laminated together to form a bilayer and subsequently rolled to reduce the thickness of each layer. Another anode tape is added to the bilayer and again rolled to a thin tape. The addition of anode tapes has the effect of reducing the electrolyte thickness while keeping the bilayer at a handleable thickness. Such process is repeated until the desired thickness of electrolyte is reached. At this point, the bilayer is cut to the desired shape and size. The green tape is then fired at elevated temperatures in an air atmosphere to sinter the bilayer tape. The cathode is then screen printed on the electrolyte surface of the sintered bilayer to produce a complete cell (Figure 39).

Figure 39 Schematic showing the process steps in the tape calendering process

Fabrication of the SOFC ceramic components by tape calendering utilizes the basic principles and processes of ceramic engineering to meet the structural and material requirements for a high-performance SOFC. Each of the components of the ceramic cell has different requirements. A thin, dense electrolyte is crucial as a gas separator membrane in the fuel cell and for high performance as the oxygen ion conductor. In the electrodes, however, a porous structure is required to reduce polarization losses. The anode has three roles, electrochemical, structural and gas delivery to the electrochemical reaction sites. Optimum electrochemical properties are only needed near the anode/electrolyte interface, thus a thin layer with these properties can be designed into fabrication process to meet these requirements. The structural and gas transport requirements are met through a multilayer anode approach. By incorporating

layers of increasing pore size on top of the electrochemical layer the structural needs can be traded off with the gas transport need to create a multilayer anode.

Some of key requirements that the fabricated cells must meet are: a) strength-cell should be able to withstand handling and stacking stresses; b) flatness-flatter cells are better to facilitate to provide uniform contact during stacking; and c) performance-cell material and microstructure should be able to provide high performance in order to reduce overall cost. As shown in figure 1, cells sizes of 16cm (6.3") meeting the above requirements are routinely fabricated at GE HPGS.

State-of-the-art in cell size reported for planar stacks is as follows: Mitsubishi Heavy Industries, Japan (20 cm x 20 cm); Fuel Cell Energy, USA (15 cm or 5.9 inch diameter); Julich, Switzerland (20 cm x 20 cm); Ceramic Fuel Cells Limited, Australia (13 cm diameter); General Electric, USA (20.32 cm or 8 inch diameter); and Indec, Switzerland (15.6 cm or 6.13 inch diameter). (21). As evident from the above information, the focus of this work, i.e., fabrication and testing of cell sizes greater than 10" diameter, would be beyond that demonstrated to date in SOFC planar cell technology.

2.3.3 Large Cell Fabrication

Initial attempts on large cell fabrication involved rolling larger tapes of baseline composition and firing the cells in the standard manner used for baseline tapes. However, two successive attempts on all 6 tapes resulted in severe cracking of the type typically seen in early stage failures (Figure 40).

Figure 40 Photograph illustrating failures during firing of a large cell

It was also decided to prepare some tapes with a modified composition based on the results from experiments and prior experience. Key processing steps were also modified. When these changes were incorporated, the first firing run showed excellent results (Figure 41).

Figure 41 Photographs of successfully fired large cells up to 12.75" in size.

After the first firing run demonstrated that 12" bilayers can be fired, additional runs were made to ensure the repeatability of firing step. The average firing yield on 20 cells from 5 runs was about 67%.

One of the other major issues encountered was the breakage of cells either during firing or in handling due to defects. A distinctive feature of these defect-related failures was that the cell broke into two-or more large pieces (Figure 42). Visual and optical microscopic examination of the cross-section of cells broken in this configuration revealed large particles with slight bluish hue. SEM examination of the cross-section revealed that these defects were about 100-200 microns in size and were rich in alumina. It is possible that the bluish color observed may be due to the formation of a nickel aluminate compound.

Figure 42 Surface view of a broken cell and x-sectional view indicating the presence of a defect and its chemical composition

Possible source of these defects include contamination in the raw material or particles in ambient atmosphere. A procedure was instituted to reduce the possibility of raw material contamination with large particles. Subsequent cells firing runs with the tape made using the sieved powders did not exhibit the defect-related failures. These results, though preliminary, do indicate the potential process improvements that can be made by addressing issues related to special cause defects.

2.3.4 Large Cell Performance Test

During this reporting period, two large-area cells were tested by using radial sealless single cell test vehicle to assess the capability of testing large area cell in our current test vehicle and facility, and evaluate the effects of hybrid environment on the performance of large cell, particularly, the effects of pressure on cell temperature distribution, and cell flow uniformity.

DH02: 11-3/4 Inch Diameter Radial Seal-less Cell Module Test

DH02 Test Setup

The 11-3/4" single cell module was based on radial sealless design that was developed at GE HPGS. It should be understood that the assembly process of the large-area cell is not a mature process at this stage and it has been faced with a number of technical challenges so far. The assembled large-area single cell module is shown in Figure 43. The cell active area of the single cell module is 532 cm².

In order to mimic the hybrid SOFC operation conditions, two preheaters that compose of a serpentine design of 1/4" tubing were used to heat the air and fuel to the SOFC operating temperature under pressurized conditions. Both air and fuel preheaters were sized based on their maximum flow rates, the air and fuel could be heated up to 750°C at the single cell module inlets.

Figure 43 Assembled Large-area Single Cell Module with 532 cm² Active Area

DH02 Test Results

The 11-3/4" radial sealless single cell module was heated up to 800 °C for normal SOFC operations. The anode reduction was started after the single cell module

temperature reached 800 °C. The test sequence for the 11-3/4” radial sealless single cell module is listed in Table 23.

Table 23 11-3/4” Radial Sealless Single Cell Module Test Sequence

•	Initial polarization curve at fixed flows with dilute hydrogen (64% H ₂ /balanced with N ₂) (at 14.7 psia. and 800 °C);
•	Hold for 20 hours under constant current for conditioning (at 14.7 psia. and 800 °C);
•	Polarization curve at fixed flows with dilute hydrogen (at 14.7 psia. and 800°C);
•	Polarization curves at fixed fuel utilization with dilute hydrogen (at 14.7 psia. and 800°C);
•	Polarization curves at fixed fuel utilization with simulated SR fuel (at 14.7 psia. and 800°C);
•	Polarization curve at fixed flows with dilute hydrogen (at pressurized condition and 800°C);
•	Polarization curves at fixed fuel utilization with dilute hydrogen (at pressurized condition and 800°C);
•	Polarization curves at fixed fuel utilization with simulated SR fuel (at pressurized condition and 800°C);
•	Polarization curves at fixed fuel utilization with dilute hydrogen (at pressurized condition and different temperature);

a. Cell performance with diluted hydrogen fuel at fixed flow rate

The performance of 11-3/4” radial sealless single cell module running on fixed flow with 64% H₂ fuel stream is shown in Figure 44 at different pressurization levels. The peak power density at ambient pressure (14.7 psia) was measured to be 152 mW/cm² and it reached 214 mW/cm² at 61.5 psia. From ambient pressure condition, the peak power density is calculated to increase by 19% at 29.4 psia, 34% at 44.1 psia and 41% at 61.5 psia, respectively. Under low fuel and air utilization and high cell temperature (about 845 °C), the peak power density at 61.5 psia reached 253 mW/cm².

Figure 44 Performances of the 11-3/4" Radial Sealless Single Cell Module, DH-02 as a Function of Operation Pressure with 64% H₂ and Fixed Flow at 800°C

b. Effects of fuel composition (64% H₂ vs. simulated SR fuel steam)

The performance of 11-3/4" radial sealless single cell module running on 64% H₂ fuel stream was compared with that on SR fuel in Figure 45. The peak power density measured on SR fuel was 89 mW/cm² and this is way lower than that measured on 64% H₂ fuel stream (152 mW/cm²).

Figure 45 Performances of the 11-3/4" Radial Sealless Single Cell Module, DH-02 with Diluted H₂ and SR Fuel and Fixed Flow at 800°C

c. Post-mortem test and root cause analysis

Post-mortem test was performed after the first 11-3/4" radial sealless single cell module, DH-02 test to find the evidences that may lower the stack performance. Crack and burning marks were found about 1 inch away from the cell perimeter. A root cause analysis was performed to identify causes of the cell crack and high cell

ASR. Recommended actions to mitigate the risks of cell crack and high cell ASR were also identified.

DH05: 12.5 Inch Radial Seal-less Single Cell Module Test

DH05 Test Setup

To improve the performance of large-area radial sealless single cell module, reduce the ASR of the module, a 12.5" radial sealless single cell module, DH05 was assembled by using Ni foam as the anode flow field and reducing the diameter of the endplate to about 12.3 inch so that the edge of the 12.5" cell can hang out of the flow field and avoid the cell crack induced by touching the endplate (see Figure 56). The assembled large-area radial sealless single cell module, DH-05 is similar to 11-3/4" radial sealless single cell module, DH-02. The cell active area of the 12.5" radial sealless single cell module is 670 cm². Preheaters that compose of a serpentine design of 1/4" tubing were also used to heat the air and fuel to the SOFC operating temperature.

Figure 46 Reduce the Diameter of Endplate to Avoid the Cell Crack Induced by Touching Endplate

DH05 Test Results

The 12.5" radial sealless single cell module was heated up to 800 °C for normal SOFC operations. The anode reduction was started after the single cell module temperature reached 800 °C. The test sequence for the 12.5" radial sealless single cell module is listed in Table 23.

a. Cell performance with diluted hydrogen fuel at fixed flow rate

The power density curves of 12.5" radial sealless single cell module, DH05, running on fixed flow, 64% H₂ fuel and SR fuel stream, respectively, are shown in Error! Reference source not found.. The peak power density of DH-05 cell on 64% H₂ fuel stream was higher than that on SR fuel stream by around 34 – 50% throughout the whole pressurization levels employed.

Figure 47 Performances of the 12.5” Radial Sealless Single Cell Module, DH-05 with Diluted H₂ and Simulated SR Fuel at Fixed Flow at 800°C

Compared with the performance of 11-3/4” Radial Sealless Single Cell Module, DH-02, the DH-05 showed approximately 20% increase in power density at fixed flow with 64% H₂ fuel stream (see **Figure 49**). This improvement might be attributed to (i) “flatter” DH-05 cell compared to DH-02 cell since there is significant improvement in the preparation of “flatter” large-area cells, which is explained in detail in other part of this report, and (ii) the modified assembly processes that is expected to have reduced the ohmic resistance originating from poor contact between the electrode surface and flow field. Along with the improvement in cell and assembly processes, the thickness of Ni foam for anode flow field was reduced in DH-05 cell assembly. This reduced Ni foam thickness translates to increased pressure drop along the radial fuel flow path and is considered to promote convective mass transfer of fuel species within anode flow field.

Root Cause Analysis on Cell Performance

As mentioned earlier, part of the work on the cell scalability effort involved testing large-area cells (~12” diameter) and evaluating the effects of hybrid environment on the performance of large cells. Though the testing of cells of this size (~12”) was demonstrated, the performance was lower than that observed on smaller cells (0.3 W/cm² vs 0.25 W/cm² at 0.7V, 46% fuel utilization)

A root cause analysis study was undertaken to identify the reason for lower performance of large cells.

The potential reasons for the low performance were divided into two broad categories: (i) Cell related; and (ii) Stack design and assembly related. Each of these categories was further drilled down for all potential causes as explained below.

- (i) Cell related: Potential causes in this category include: (a) Cell Microstructure; (b) Cell strength; and (c) Cell flatness/ripples
 - (a) *Cell Microstructure*: Potential causes in this sub-category include differences in cell component (electrolyte, anode, and cathode) thickness, porosity, or interfacial microstructure from that of a baseline cell. For instance higher electrolyte thickness may lead to lower performance due to increased ohmic resistance or a denser electrode may experience increased diffusion polarization, and hence, lower cell performance. Cross-section samples from a tested large-area cell were prepared and thickness, porosity, and interfacial microstructure were examined using optical and scanning electron microscope (SEM) at various points along the cells. For comparison, the microstructure of a baseline cell was also examined. Slight differences in component thickness were seen between the large cell and the baseline cell. These differences, however, may not be significant enough to lead to observed low performance in large cells.
 - (b) *Cell strength*: If the cell cracks due to poor strength, it may result in lower than expected performance. However, the fact that both large cells showed lower electrochemical performance but the second cell did not exhibit any cracks in the tested area suggests that cell cracking may not be a significant factor in observed low performance.
 - (c) *Cell flatness/ripples*: Presence of ripples or uneven areas on the cells may lead to poor contact of cell with interconnect in those areas, which, in turn, may result in lowered cell performance. A design of experiments, as described below, was performed to get quantitative information on ripples or unevenness in the cell.
- (ii) Stack design and assembly related: Potential issues in this category include: cathode and anode bond paste thickness, thickness uniformity, and amount, poor contact between cell and interconnect, flow uniformity issues related to interconnect design.

To isolate the influence of stack design and assembly related issues on performance, two smaller cells (4-3/8" diameter) were laser cut from different areas on a 12.75" bilayer. Cathode was applied by hand painting (as done for large cells) and by the standard screen-printing process, respectively. The cells were tested on diluted hydrogen fuel stream. The Ni mesh fuel flow field design used especially for the large cells testing was also adopted for the small cell testing. The performance of both tested

cells was measured to be similar to the performance of standard baseline cells under similar testing conditions and test fixture (Figure 3).

Figure 48 Electrochemical Testing Data on: (i) 12"; (ii) 4.375" Baseline; and (iii) 4.375" Cell Cut from a 12" Cell.

The above results strongly suggest that performance loss in large cells is largely an artifact of the test vehicle, and properties of the cell itself, such as cell size or cell microstructure, may have a small role in the observed lower performance.

Figure 49 Performances of the Large-area” Radial Sealless Single Cell Module, DH-05 and DH-02 with Diluted H₂ and Fixed Flow at 800°C

2.3.5 Cell Fabrication Improvements

Fabrication of 12” bilayers was demonstrated in this program. More than 50 large bilayer tapes were produced and fired. The average firing yield on the critical bilayer firing step was 70%. Since post bilayer firing failures were observed due to presence of large particles and/or foreign impurities, a sieving step on raw material powders was introduced to remove foreign impurities or large particles. This resulted in an improvement in post-firing yields during bilayer cleaning and cathode application from 55% to about 80%.

Due to the presence of an edge curl on the bilayer, cathode application was done on the first few cells using hand painting. Laser cutting the cell removes the edge curl, and hence, facilitates the screen-printing operation for cathode application. The screen-printing process is desirable for the control and uniformity of the layer thicknesses and for having the potential of being a low-cost, high throughput manufacturing process. With the availability of multiple large cells a new fixture was ordered for the lasercutting operation. Over 20 large bilayers were successfully cut to 12” diameter with a lasercutting yield of almost 100%. With the success in laser cutting these cells, screen-printing process development work for cathode application was initiated. A large screen was designed and procured and larger screen print equipment was commissioned. Several process parameters were optimized to overcome debonding issues in the green and fired stages. The optimized process resulted in successful application of cathode on large bilayers using the screen-printing process; thus validating the applicability of screen-printing to fabricate scaled-up cells.

As cell sizes get larger, stiffness of bilayer and its strength become more important for handling and stacking purposes. One of the ways to increase bilayer strength and stiffness is to increase the thickness of anode, and hence, the total thickness. Initial attempt to increase the green tape thickness resulted in severe cracking and delamination of electrolyte during firing. With the modification of the firing process, the firing run was a success (both bilayers 100% intact). Further, the bilayers looked quite flat with very little edge curl. Success in firing these thicker tapes opens up ways to improve the strength, flatness, and edge curl of large-area bilayers. However, any potential decrease in performance with increased cell thickness, especially at high fuel utilizations, needs to be determined.

2.3.6 Design of Experiments

As mentioned earlier, certain cell properties such as flatness affect the stacking behavior and performance of the cells. However, exact correlation between these parameters and cell performance is not known. The stack requirements that are affected by cell properties are: cell edge sealability; electrical contact between the cells and interconnect; and mechanical integrity. It is therefore important to understand the manufacturing process parameters that affect the above-mentioned properties. As part of this task, a design of experiments was initiated to understand bilayer characteristics as a function of size and other parameters.

After sintering, a laser based measurement system (Acugage, Manchester, NH) was used to generate a two dimensional scan of the bilayer surfaces. Line scans were used to measure the edge curl. The output from the program is a series of x, y and z measurements for each location on the cell surface. The measurements were made on the bilayer as it sat freely on the Acugage table. A representative scan is shown in Figure 8.

Figure 50 Surface Profile of a Scanned Bilayer. (The color scale shows the variation in the z height over the surface.)

A schematic of the circumferential surface ripple and the edge curl are shown in Figure 51. The surface ripple typically spreads out from the center, with the wavelength and amplitude increasing as we move away from the center. As can be inferred, in a fuel cell stack, as the cell and interconnect surfaces mate in the stacking process, the points which are out of flatness in the cell will be the main points of contact, with the other areas having lesser contact. The edge curl is a steep camber at the very edge of the cell, which can severely affect sealing ability. As mentioned earlier, the screen printing operation typically requires removing the edge curl; presently by a laser cutting process.

Figure 51 (a) Surface Ripple and (b) Edge Curl in a Bilayer.

As the extraction of the amplitude and wavelength from the unconstrained laser scan is not straightforward, a Matlab program was developed to analyze the scanned surface profiles for surface ripple using Fast Fourier Transform analysis. The average amplitude for a set of predetermined wavelengths was then calculated (Figure 52). As the values shown are a weighted average over the whole cell, they are lower than localized high ripples that might be of more concern. Hence, it should be noted that a direct comparison of these average values may not translate into performance difference between cells. The edge curl values are plotted in Figure 53.

Figure 52. Average Amplitude of Surface Ripple in Different Bilayers

Figure 53. Edge Curl Data for Various Cells

The data was analyzed using GE's Design for Six Sigma (DFSS) Design of Experiments optimizing tool and a transfer function was generated to correlate the most significant factors affecting surface ripple and edge curl. It was seen that thickness and size were the most significant factors affecting surface ripple. Though the above results provide preliminary trends on variation of bilayer properties with several key process parameters, more detailed studies are needed to generate accurate correlations.

2.3.7 Literature review on theoretical limitations

Literature review on theoretical models to understand fundamental limitations on the maximum sizes that can be made using the tape calendaring process was completed. During the course of review, it became clear that theoretical limitations on cell size couldn't be predicted using available binder burnout and sintering models. Current models need to be modified to take into account effect of external forces (thermal, gravity, frictional etc.). Though there are no obvious theoretical limitations, practical limitations viz. thermal gradients, uniform binder removal, friction between tape and setter plates, and availability/warpage of setter plates may dictate the upper limit on cells sizes that can be made.

2.4 Stack Hybridization

The objective of this effort was to demonstrate operation of planar SOFC stacks and their operating characteristics under hybrid environment and assess scalability of stack design. SECA derived technology was leveraged in this task. The work focused on stack testing of various sizes under hybrid conditions, such as temperature, pressure, and gas composition, which are critical to the development of megawatt class hybrid demonstration systems. Stack performance as a function of fuel utilization was determined. Analysis of design features that influence stack reliability, performance, and scalability (area and height) was performed. The dynamic performance of stacks under hybrid conditions including startup and shutdown was evaluated. Designs and operating procedures to facilitate stack scaleup were identified.

2.4.1 Stack Tests

A 2-cell and two 5-cell stacks were tested in hybrid conditions. All the stacks were circular half sealed design and with 142 cm² active area. The performances of the stacks as a function of pressure, temperature, fuel utilization, and fuel composition was evaluated. The results of these tests are summarized below.

DH01: 2-Cell Stack Test

During this reporting period, a 2-cell stack, DH-01, was tested to evaluate the effects of hybrid environment on the performance of SOFCs, particularly, the effects of pressure on cell seal, cell temperature distribution, and cell flow uniformity since the 2-cell stack contains all of the design features of the n-cell stack. The functionalities and safety of SOFC pressurized test stand were also checked during this test.

DH01 Test Setup

The 2-cell stack was based on the stack that was designed for SECA program. In order to mimic the hybrid SOFC operation conditions, preheaters that compose of coils of ¹/₈" tubing were used to heat the air and fuel to the SOFC operating temperature under

pressurized conditions. The assembled 2-cell stack is shown in Figure 43. The cell active area of the 2-cell stack is 142 cm².

Figure 54 Assembled 2-cell stack with 142 cm² active area per cell

DH01 Test Results

The 2-cell stack was heated up to 900 °C for curing the manifold, anode and cathode cell seals. Then, the stack was cooled down to 800 °C for normal SOFC operations. The anode reduction was started during the stack heating up when stack temperature reached 600 °C. The test sequence for the 2-cell stack is listed in Table 24.

Table 24 2-Cell stack test sequence

• Initial polarization curve at fixed flows with dilute hydrogen (64% H₂/balanced with N₂) (at 14.7 psia. and 800°C); • Hold for 20 hours under constant current for conditioning (at 14.7 psia. and 800°C); • Polarization curve at fixed flows with dilute hydrogen (at 14.7 psia. and 800°C); • Polarization curves at fixed fuel utilization with dilute hydrogen (at 14.7 psia. and 800°C); • Polarization curves at fixed fuel utilization with simulated SR fuel (at 14.7 psia. and 800°C); • Polarization curve at fixed flows with dilute hydrogen (at pressurized condition and 800°C); • Polarization curves at fixed fuel utilization with dilute hydrogen (at pressurized condition and 800°C); • Polarization curves at fixed fuel utilization with simulated SR fuel (at pressurized condition and 800°C); • Polarization curves at fixed fuel utilization with dilute hydrogen and simulated SR fuel (at pressurized condition and different temperature);

a. Effects of fuel composition (64% H₂ vs. simulated SR fuel steam)

The performance of the 2-cell stack, DH-01 running on diluted hydrogen fuel (64% H₂, balanced with N₂) and simulated steam reformat (SR) fuel stream are shown in Figure 44. The fuel flow rate was fixed throughout the whole current density range and therefore, fuel utilization increases linearly with current density as shown in Figure 44. The performance curves (I – V curves) were measured at both ambient pressure (14.7 psia) and elevated stack operation pressure (58.5 psia). Cells running on 64% H₂ fuel stream show cell voltages higher than SR fuel stream by approximately 0.5 Volt (both ambient and pressurized condition) when compared at same current density. Pressurization of the stack operation pressure from ambient pressure to 58.8 psia raises the cell voltages and the voltage gain through pressurization is seen generally to grow with current density (compare solid line and dotted line in Figure 44). It should be noted that the open circuit voltage of cells running on SR fuel is raised (from 0.971V to 1.026V on cell #1 in Figure 44) when the stack was pressurized from ambient pressure to 58.8 psia. This increase in open circuit voltage is in good agreement with thermodynamic prediction. With SR fuel stream, thermodynamic calculation predicts that pressurization decreases equilibrium O₂ composition in fuel stream and leads to increased open circuit voltage.

The low cell voltage (or performance) observed with SR fuel stream vs. 64% H₂ fuel stream can be explained primarily by internal reformation. The SR fuel stream contains 13.6% CH₄ (the CH₄ slip comes from steam reformer for external reformation), and the CH₄ content is dropped to less than 1.0% through internal reformation within SOFC anode compartment (the high conversion of CH₄ through internal reformation within SOFC anode compartment has been measured in separate studies at HPGS and is close to thermodynamic equilibrium predictions). The occurrence of this internal reformation, which is highly endothermic (approximately 205 kJ/mole), lowers the stack temperature (especially cell temperature) and leads to reduced cell voltages. The local cell temperature inside the stack is anticipated to vary widely and more precise measurement of local cell temperature is required to clearly understand internal reformation driven local cooling effect.

Figure 55 Performances of the stack, DH-01 with diluted H₂ and steam-reformate fuel

b. Effects of fuel utilization

Cell performances were also measured with fixed fuel utilization. Fuel flow rates were controlled at given current densities to match fuel utilization specification (i.e., fuel flow increases linearly with current density). Figure 45 shows cell voltages measured with 75% fuel utilization on 64% H₂ fuel and simulated SR fuel streams, respectively. The starting current density of these measurements was higher than 300 mA/cm² because the requirement of the low fuel flow rate for low current density reached the limitation of the mass flow controllers.

As shown in Figure 45, the voltage gain through pressurized operation was most significant at initial pressurization level (e.g., when going from 14.7 psia to 29.4 psia), the high pressurization level (e.g., from 44.1 psia to 58.8 psia) did not raise cell voltages appreciably (less than 20 mVs). It is seen that 64% H₂ fuel maintains higher cell voltage over simulated SR fuel at 75% fuel utilization condition.

Figure 56 Performances of cell #2 of DH-01 stack at 75% fuel utilization with dilute H₂ and steam reformat fuel

The I-V curves of cell #2 shown in Figure 45 are plotted for power density in Figure 57. With 64% H₂ fuel stream, the power density of cell #2 at ambient pressure is measured to be 0.32 mW/cm² and it reaches 0.39 mW/cm² at 58.8 psia. With simulated SR fuel stream, power density at ambient is measured to be 0.27 mW/cm² and 0.38 mW/cm² at 58.8 psia, respectively. The increases of power density with simulated SR fuel stream from ambient pressure (14.7 psia) is calculated to be 18% at 29.4 psia, 31% at 44.1 psia, and 40% at 58.8 psia.

Figure 57 Power Density of Cell #2 of the stack, DH-01 at 75% fuel utilization with diluted H₂ and steam reformate fuel

c. Effects of SOFC operation temperatures

The effect of temperature on SOFC performance was measured under pressurized condition (58.8 psia). Figure 57 shows cell performance on 64% H₂ fuel at 75% fuel utilization. Vessel temperature was maintained at 750°C, 800°C and 850°C, respectively, during performance measurement.

As shown in Error! Reference source not found., stack operation temperature brings about noticeable change in cell performance. When compared to power density at 800°C for cell #1, 13% increase in power density was observed at 850°C and 27% drop in power density was observed at 750°C, respectively.

The effect of temperature on cell performance could also be clearly observed with simulated SR fuel stream as shown in Figure 58. The power density measured at 850°C on cell #2 running on simulated SR fuel stream (Error! Reference source not found.) reached 0.41 W/cm² and this power density translates to 62% increase in power density over that (0.25 W/cm²) at 750°C.

Figure 58 Performances of the stack, DH-01 as a function of temperature at 75% fuel utilization with diluted H₂

Figure 59 Performances of the stack, DH-01 as a function of temperature at 75% fuel utilization with simulated steam reformate fuel

d. Stack performance degradation

DH-01 stack has been tested for over a week period of time and has gone through 6 pressure cycles. During the test period, cell performance at fixed test condition (64% H₂ fuel stream, fixed flow, 800°C, 14.7 psia) was repeated to check stack performance day-to-day stability. The cell voltages measured at certain current densities are shown in Figure 46. So far, it was observed that pressure cycle (pressurizing and de-pressuring of the stack) does not apparently affect individual cell stability. However, it is too early to quantify cell stability at this stage because of limited test data. In this task, not much attention has been given to long-term cell stability issue. For clear understanding of cell stability issues within the framework of hybrid system configuration, more stability test with long-term time frame is considered necessary.

Figure 60 Performance stability of the stack, DH-01 with dilute H₂ at 800°C and 14.7 psia

DH06: 5-Cell Compliant Feed Tube Stack Tests

After the successful test of the 2-cell stack, two 5-cell stacks, DH-06 and DH-07, were tested under hybrid conditions. The purpose of the two 5-cell stack tests was to evaluate the performance of the SOFC stack as a function of pressure, temperature, and fuel utilization and the capability of the compliant feed tube manifold system (compliant feed tube and manifold sealant) at hybridization conditions. Simulated steam-reforming fuel was used in the tests.

DH06 Test Setup

Because the stack design showed several major risks, such as manifold seal leakage, manifold shorting, and tolerance stack-up for interconnect during the stack test, the compliant feed tube stack design that was also developed under SECA program was used in the 5-cell stack tests. The advantages of the compliant feed tube stack design are (i) using feed tubes replace the bellows to provide the compliancy between stack core and manifold, and avoid the loss of the compression on manifold due to the bellows material anneals at the SOFC operation temperature; (ii) capable of presealing cell to interconnect and reducing cell that allow inspection of the cell crack, cell shorting, cell seal, pressure drop of anode flow field, and elimination of bad cell module before it is used in the stack.

The assembled 5-cell compliant feed tube stack is shown in Figure 61. The cell active area of the 5-cell stack is 142 cm². In order to mimic the hybrid SOFC operation conditions, two preheaters that compose of a serpentine design of 1/4" tubing were used to heat the air and fuel to the SOFC operating temperature. Both air and fuel preheaters were sized based on their maximum flow rates, the air and fuel could be heated up to 750°C at the stack inlets. An Isometric view of air preheater design is shown in Figure 62.

Figure 61 Assembled 5-cell compliant feed tube stack with 142 cm² active area per cell

Figure 62 Isometric view of the air preheater design

DH06 Test Results

The 5-cell stack was heated up to 900 °C for curing the manifold and cathode cell seals (anode cell seal was already formed during the pre-seal/reduction process). Then, the stack was cooled down to 800 °C for normal SOFC operations. The test sequence for the 5-cell stack is listed in Table 25.

Table 25 5-Cell Stack Test Sequence

- Initial polarization curve at fixed flows with dilute hydrogen (64% H₂/balanced with N₂) (at 14.7 psia. and 800°C);
- Hold for 20 hours under constant current for conditioning (at 14.7 psia. and 800°C);
- Polarization curve at fixed flows with dilute hydrogen (at 14.7 psia. and 800°C);
- Polarization curves at fixed fuel utilization with dilute hydrogen (at 14.7 psia. and 800°C);
- Polarization curves at fixed fuel utilization with simulated SR fuel (at 14.7 psia. and 800°C);
- Polarization curve at fixed flows with dilute hydrogen (at pressurized condition and 800°C);
- Polarization curves at fixed fuel utilization with dilute hydrogen (at pressurized condition and 800°C);
- Polarization curves at fixed fuel utilization with simulated SR fuel (at pressurized condition and 800°C);
- Polarization curves at fixed fuel utilization with dilute hydrogen (at pressurized condition and different temperature);

a. Effects of fuel composition (64% H₂ vs. simulated SR fuel steam)

5-cell compliant feed tube stack, DH-06 has been tested with 64% H₂ fuel and SR fuel stream, respectively. The initial performance of individual cells in the 5-cell stack, DH-06 at 800 °C and 14.7 psia are shown in Figure 63. As shown in Figure 63, cell #3 is relatively under-performing compared to other cells in the stack and the current density employed for testing of the 5-cell stack, DH-06 was limited in most cases by the low cell voltage of cell #3 (generally, current densities are adjusted to maintain all cell voltages above 0.55 V).

Figure 63 Performances of Individual Cells in the Compliant Feed Tube Stack, DH-06 with Diluted H₂ at 800°C and 14.7 psia

The average performance of cells in 5-cell stack, DH-06 running on 64% H₂ fuel stream under fixed flow condition was compared with that on SR fuel stream in Figure 64. Compared to 64% H₂ fuel stream, average cell voltages measured on SR fuel stream are lower by more than 50 mVs at stack operation pressures up to 61.5 psia.

Figure 64 Performances of Average Cell in the Compliant Feed Tube Stack, DH-06 as a Function of Operation Pressure with Diluted H₂ and SR Fuel, Fixed Flow at 800°C

b. Effects of fuel utilization

The I-V curves of average cell of 5-cell stack, DH-06 at 60% fuel utilization are shown in Figure 65 and they are compared with those obtained with fixed fuel flow

condition. As anticipated, cell voltages with 64% H₂ fuel stream at 60% fuel utilization appear to approach those under fixed flow condition with the increase in current density. With SR fuel stream, cell performance data at 60% fuel utilization could only be obtained at 61.5 psia due to the under-performing cell #3.

Figure 65 Performances of Average Cell in the Compliant Feed Tube Stack, DH-06 as a Function of Operation Pressure with Diluted H₂ and SR Fuel at 800°C, Fixed Flow vs 60% Fuel Utilization

c. Effects of the transition of stack operation pressure

Compliant feed tube stack, DH-06 did go through 2 pressure cycles and further testing at pressurized condition could not be made because of the abnormal rising of the temperature (up to 900 °C) in fuel exhaust manifold during the stack pressurization. The temperature rise was also accompanied by gradual drop in OCV (open circuit voltage) of cell #5 during pressurization (by as much as 80 mV). Upon depressurization, the fuel exhaust manifold returned to normal temperature, but the OCV of cell #5 remained low (lower than previous OCV by 20 mV).

Chronological analysis of observations and events showed that the initiation of cell #5 OCV drop preceded the rise in fuel exhaust manifold temperature and stack pressurization. The cell #5 OCV drop coincided with the transition of stack pressurization, i.e. fuel stream from “by-pass of back pressure regulator” to “through backpressure regulator”. The fuel exhaust manifold temperature was found to rise approximately 10 minutes after the stack pressurization started. Figure 66 shows the excursion of individual cell OCVs and temperature measurements of the 5-cell stack, DH-06.

Post-mortem test was performed after 5-cell stack, DH-06 test to find the root causes of the abnormal rising of the fuel exhaust temperature and low OCV of cell #5 and evidences that may lower the cell performance (cell #3). Crack and burning marks were found near the edge of cell #5 between the flow field and seal area (see Error! Reference source not found.). The post-mortem test also indicates that the

cathode bonding is weak, there is no bonding strength between cathode and cathode interconnect, especially of cell #3. A root cause analysis was performed to identify causes of the cell crack that resulted in anode exhaust temperature increase and OCV of cell #5 decrease during the stack pressure transition. Based on information gathered so far, it was suggested that the transition of fuel stream from “by-pass of back pressure regulator” to “through backpressure regulator” generated certain level of differential pressure between anode and cathode, and led to crack the cell #5. The differential pressure recorded during fuel stream transition is lower than 1 psia, but it is highly probable that local differential pressure at very short period of time during fuel stream transition may be much higher than the recorded differential pressure. The crack in cell #5 is considered responsible for the OCV drop and might have led to abnormal temperature rise in fuel exhaust manifold through accelerated crossover of fuel and air upon pressurization.

Figure 66 Observation of OCV of Cell #5 in the Compliant Feed Tube Stack, DH-06 and Fuel Exhaust Manifold Temperature during the Transition of Stack Operation Pressure

Recommended actions to mitigate the risks of cracking cell during the stack operation were identified during the root cause analysis. Since one of the potential causes for the cell crack during the operation pressure transition is the unbalanced anode and cathode seal that is introduced by cell preseat/reduction process, a one-step stack assembly process was suggested to confirm the hypothesis.

DH07: 5-Cell Compliant Feed Tube Stack Test

DH07 Test Setup

To confirm the hypothesis that cell crack is induced by unbalanced anode and cathode seal, a 5-cell stack, DH-07 was assembled by using compliant feed tube interconnect and following the one-step stack assembly process, i.e. omit the preseal/reduction process and assemble the stack in one step and curing the anode and cathode seal and manifold seals all together. The assembled 5-cell compliant feed tube stack is similar to 5-cell stack, DH-06 shown in Figure 61. The cell active area of the 5-cell stack is 142 cm². Preheaters that compose of a serpentine design of 1/4" tubing were also used to heat the air and fuel to the SOFC operating temperature.

DH07 Test Results

The 5-cell stack was heated up to 900 °C for curing the manifold, anode and cathode cell seals. Then, the stack was cooled down to 800 °C for normal SOFC operations. The anode reduction was started during the stack heating up when stack temperature reached 600 °C. The test sequence for the 5-cell stack is listed in Table 25.

a. Effects of fuel composition (64% H₂ vs. simulated SR fuel steam)

DH-07 5-cell stack has been tested without going through fuel flow transition to avoid cell cracking problem, which has experienced in DH-06 stack test. Testing of DH-07 stack revealed that the upper 3 cells had consistently low OCVs and pressurization led to increased OCVs with the upper 3 cells. The performances of individual cells in the stack, DH-07 with 64% H₂ and fixed fuel rate at 800 C are shown in Figure 67. As shown in Figure 67, the ASR (area specific resistance) of these 3 cells was comparable or better than the other 2 cells (which showed normal OCVs). It was suggested that shorting either across the cell originating from defective cells or through manifold seal might have occurred. Post-analysis of the DH-07 stack is planned for clear explanation of the shorting problem.

The performance of cell #1 and cell #2 in the 5-cell stack, DH-07 running on 64% H₂ fuel and SR fuel stream are shown in Figure 67, respectively. The fuel flow rate was fixed throughout the whole current density range. The voltage gain of 64% H₂ fuel stream over SR fuel stream is measured to be approximately 100 mV and that voltage gain translates to power density gain in the range of 15% - 18% at current density of 300 mA/cm².

Figure 67 Performances of Individual Cells in the Compliant Feed Tube Stack, DH-07 with Diluted H₂ at 800°C and 14.7 psia

Figure 68 Performances of Cell #1 and #2 in Compliant Feed Tube Stack, DH-07 as a Function of Operation Pressure with Diluted H₂ and SR Fuel Stream, Fixed Flow at 800°C

b. Effects of fuel utilization

The performance of cell #1 and cell #2 in the 5-cell stack, DH-07 with 60 % fuel utilization is shown in Figure 68. It appears that pressurization brings I-V curves of fixed fuel utilization (60% fuel utilization in Figure 69) closer to that of fixed flow condition. Pressurization is understood to reduce mass transfer over-potential in current generation. Therefore, cell voltage gain attained through low fuel utilization

(compared to high fuel utilization) would be much less pronounced at high pressurization level.

One thing to note is that under pressurized (61.5 psia) condition, cell performance of the 5-cell stack, DH-07 was not stable at 60% fuel utilization. The cell voltage under load at 60% fuel utilization was highly noisy, unstable and continued to drop at 61.5 psia stack operation pressure. At this stage, the reason for the voltage dropping and fluctuations is not clearly understood. One possible explanation is that there might be serious leak through fuel or air manifold upon pressurization, which leads to measurable effect on cell voltage under load.

Figure 69 Performances of Cell #1 and #2 in Compliant Feed Tube Stack, DH-07 as a Function of Operation Pressure with Diluted H₂ and SR Fuel Stream at 800°C, Fixed Flow vs 60% Fuel Utilization

DO11 5-Cell Compliant Feed Tube Stack Test

DH11: Test Setup

The compliant feed tube stack design was used in this 5-cell stack test. Because the root cause analysis indicate that unbalanced glass seal formed during the cell preseat/reduction process may induce cell edge cracks, the 5-cell stack, DH11 was assembled by using compliant feed tube interconnect and followed the one-step stack assembly process. The cell active area of the 5-cell stack was 142 cm². Preheaters based on a serpentine design of ¼" tubing were also used to heat the air and fuel to the SOFC operating temperature. Both air and fuel preheaters were sized based on their maximum flow rates, the air and fuel could be heated up to 750°C at the stack inlets.

Test Results

The stack was tested at 800 °C for normal SOFC operations. The anode reduction was started during the stack heating up when stack temperature reached 600 °C. The test plan for the 5-cell stack is given in Table 9

Table 26 5-Cell Stack Test Plan

• Initial polarization curve at fixed flows with dilute hydrogen (64% H₂/balanced with N₂) (at 14.7 psia. and 800°C);• Hold for 20 hours under constant current for conditioning (at 14.7 psia. and 800°C);• Polarization curve at fixed flows with dilute hydrogen (at 14.7 psia. and 800°C);• Polarization curves at fixed fuel utilization with dilute hydrogen (at 14.7 psia. and 800°C);• Polarization curve at fixed flows with dilute hydrogen (at pressurized condition and 800°C);• Polarization curves at fixed fuel utilization with dilute hydrogen (at pressurized condition and 800°C);• Polarization curves at fixed fuel utilization with simulated steam reforming or SR fuel (at pressurized condition and 800°C);• Polarization curves at fixed fuel utilization with simulated SR fuel (at 14.7 psia. and 800°C).

a. Effects of fuel utilization (64% H₂ fuel)

The performance of individual cells in the 5-cell compliant feed tube stack, DH-11 running on diluted hydrogen fuel (64% H₂, balanced with N₂) fuel stream at 800 °C and 14.7 psia are shown in Figure 9. The fuel flow rate was fixed throughout the whole current density range and therefore, fuel utilization increases linearly with current density as shown in Figure 9. A power density of 307 mW/cm² was achieved at 45% fuel utilization and average cell voltage of 0.607 V. A typical baseline cell performance was achieved in this test.

Performances of individual cells in the 5-cell stack were also measured with 70% fixed fuel utilization at 800 °C and 14.7 psia. During this measurement, fuel flow rates were controlled at given current densities to match fuel utilization specification (i.e., fuel flow increases linearly with current density). The starting current density of these measurements was higher than 260 mA/cm² because the requirement of the low fuel flow rate for low current density reached the limitation of the mass flow controllers.

As shown in Figure 10, this 5-cell compliant feed tube stack achieved power density of 281 mW/cm² (0.588 V @ 0.479A/cm²) with 64% H₂, 70% fuel utilization at 800°C, but, performance of the 5-cell stack at 70% fuel utilization is lower than those obtained with fixed fuel flow. There is about 16% performance difference between

fixed flow rate and fixed fuel utilization condition at low current density. However, as anticipated, cell performance with 64% H₂ fuel stream at 70% fuel utilization appears to approach those under fixed flow condition with the increase in current density, which indicates that mass transfer does limit the cell performance.

Figure 70 Performances of Individual Cells in the Compliant Feed Tube Stack, DH-11 with Diluted H₂ and Fixed Flow at 800°C and 14.7 psia.

Figure 71 Performances of Individual Cells in the Compliant Feed Tube Stack, DH-11 with Diluted H₂ at 800°C and 14.7 psia. Fixed Flow vs 70% Fuel Utilization.

b. Effects of the differential pressure between anode and cathode

The test of the 5-cell compliant feed tube stack, DH-11 at pressurized condition was terminated due to the abnormal rising of the temperature (up to 850 °C) in fuel exhaust manifold and the open circuit voltage (OCV) drop of cell #5 during the stack pressurization.

The analysis of the recorded temperature, pressure, and cell voltage data and events showed that the initiation of cell #5 OCV drop and the temperature rise in the fuel exhaust manifold were associated with the change of differential pressure between anode and cathode. Because the inlet pressure control scheme was used during the operation of the stack pressurization and the sealless design was used in the cathode flowfield of the compliant feed tube stack, this phenomenon (cell OCV drops and the temperature of the fuel exhaust manifold raises when cathode pressure is higher than anode pressure) implied that there was leak path between anode of the cell #5 and the cathode. When the anode pressure was higher than the cathode pressure, the fuel leaked into the cathode and caused fuel combustion in the cathode or pressure vessel. This type of fuel leakage affected stack performance at high fuel utilization condition, but did not damage stack and cause catastrophic failure of the stack since the heat was quickly carried out by rapid flow of air functioning as an oxidant and a thermal management medium. The cell OCV was not dramatically affected. However, when the anode pressure is lower than the cathode pressure, the air will leak into the anode and cause fuel combustion in the anode compartment. The consequences of this type of air leakage are cell OCV drop, temperature of the fuel exhaust rising, and catastrophic failure of the stack.

After the test, post-mortem test was performed to determine the root causes of the abnormal rising of the fuel exhaust temperature and the OCV drop of cell #5. A burning mark was found near the seal area cell #5. This finding confirms the leak path hypothesis. It is believed that this edge leakage caused the low OCV and high fuel exhaust temperature at pressurized conditions.

A root cause analysis was performed to identify causes of the swing of the differential pressure between anode and cathode during stack pressurization. An issue associated with the setup of the pressure control hardware was identified. The pressurized test stand was then modified to ensure that there is always a + 0.25 offset pressure add to the anode pressure to avoid the air leaks into the anode compartment.

2.4.2 Large-area Cell Stack Test

After the successful test of the two large-area single cell modules discussed earlier, two 3-large area cell stacks were tested under hybrid conditions. The purpose of the tests of 3-large area cell stacks was to evaluate the performance of the large area cell as a function of pressure, temperature, and fuel utilization. The cell flow uniformity and the capability of the compliant feed tube manifold system (compliant feed tube and manifold sealant) were also examined at hybridization conditions. Simulated steam-reforming fuel was used in the tests.

DH12: 12" 3-cell Radial Stack

Test Results

The 12" 3-cell radial stack was tested down to 800 °C for normal SOFC operations. The anode reduction was started after the 3-cell stack temperature reached 800 °C. The test plan for the 12" 3-cell radial stack is given in Table 10.

Table 27 3-Cell (12") Radial Stack Test Plan

- Initial polarization curve at fixed flows with dilute hydrogen (37% H₂/balanced with N₂) (at 14.7 psia. and 800 °C);
- Hold for 20 hours under constant current for conditioning (at 14.7 psia. and 800 °C);
- Polarization curve at fixed flows with dilute hydrogen (at 14.7 psia. and 800°C);
- Polarization curves at fixed fuel utilization with dilute hydrogen (at 14.7 psia. and 800°C);
- Polarization curve at fixed flows with dilute hydrogen (at pressurized condition and 800°C);
- Polarization curves at fixed fuel utilization with dilute hydrogen (at pressurized condition and 800°C);
- Polarization curves at fixed fuel utilization with simulated SR fuel (at pressurized condition and 800°C);

a. Cell performance with diluted hydrogen fuel at fixed flow rate

The performance of 12" 3-cell radial stack running on fixed flow with 37% H₂ fuel stream is shown in Figure 17 at different pressurization levels. The peak power density at ambient pressure (14.7 psia) was measured to be 213 mW/cm² and it reached 290 mW/cm² at 61.5 psia. From ambient pressure condition, the power density at similar operation condition (36.4% fuel utilization and 0.342A/cm²) is calculated to increase about 4% at 30 psia, 10% at 45 psia and 19% at 60 psia, respectively. At 60 psia, the stack achieved total power of 532 W (0.592 V at 0.489 A/cm², power density of 0.290 W/cm²) at 52% fuel utilization with 37% H₂ fuel stream

Figure 72 Performances of the Average Cell in the 12" 3-Cell Radial Stack, DH-12 as a Function of Operation Pressure with 37% H₂ and Fixed Flow at 800°C

b. Effects of fuel composition (37% H₂ vs. simulated SR fuel steam)

The performance of 12" 3-cell radial stack running on 37% H₂ fuel stream at 60% fuel utilization and 60 psia operation pressure was compared with that on simulated SR fuel in Figure 18. The peak power density measured on SR fuel was 216 mW/cm² and this is lower than that measured on 37% H₂ fuel stream (247 mW/cm²). However, at same operation condition the difference of the performance between operated at 37% H₂ fuel stream and simulated SR fuel steam is less than 6%.

Figure 73 Average Cell Performance of 3-Cell (12") Radial Stack, DH-12 with Diluted H₂ and SR Fuel and 60% Fuel Utilization at 800°C

c. Effects of fuel utilization

Cell performance was also measured with fixed fuel utilization. Fuel flow rates were controlled at given current densities to match fuel utilization specification (i.e., fuel flow increases linearly with current density). Figure 19 shows cell voltages measured with 60 %, 70%, and 75% fuel utilization on simulated SR fuel streams at 60 psia operation pressure. The starting current density of these measurements was higher because the requirement of the low fuel flow rate for low current density reached the limitation of the mass flow controllers. As anticipated the average cell performance decrease as the increase of fuel utilization due to the limitation of the mass transfer. At 60 psia operation pressure, the stack achieved power density of 0.157 W/cm² at 75% fuel utilization (0.635 V at 0.245 A/cm²) with simulated SR reformat (stack efficiency: 38%).

The performance of this 3-cell stack is much better than the large area single cell module tested before (about 28% performance increase, 0.229 W/cm², 0.672 V/cell at 340 mA/cm², vs. 0.179W/cm², 0.665V at 268.7 mA/cm² under similar operating conditions); However, the performance of this 3-cell stack is still lower than the performance of 6.3" cell stack (0.335 W/cm² at 75% fuel utilization with 64% hydrogen), the possible causes for the these may be due to the contact resistance and poor flow distribution.

Figure 74 Average Cell Performance of 3-Cell (12") Radial Sealless Stack, DH-12 as a Function of Fuel Utilization with Simulated SR Fuel at 800°C

R14: 12" 3-cell Half Seal Stack Test

Test Results

The 3-cell (12") half seal stack was tested at 800 °C for normal SOFC operations. The anode reduction was started after the 3-cell stack temperature reached 800°C. The anode flowfield used in this stack was designed to improve the fuel flow uniformity. To explore the effect of cell microstructure on the cell performance, three different type of 12" cells were used, (1) cell made of 0.025" thick baseline bilayer tape, (2) cell made of 0.017" thick baseline bilayer tape (this is GE HPGS's baseline cell thickness), and (3) "Option 1" cell made of 0.025" thick "Option 1" bilayer tape, were used in this stack.

The initial performance of 12" 3-cell half seal stack running on fixed flow with 64% H₂ fuel stream is shown in Figure 23. After 18 hours of conditioning, the performance of the 3-cell half sealed stack achieved power density of 0.273 W/cm² (0.675 V/cell at 405 mA/cm²) with 64% dilute hydrogen and 35% fuel utilization at ambient pressure (14.7 psia), which is better than the 1st 3-large area cell radial stack tested before (241 mW/cm², 0.592V at 408 mA/cm²) under similar operating conditions. This is about 13% of performance improvement. However, the stack cannot perform stably at pressurized condition due to the issue of pressurized test stand. The pressure regulator of anode side was unable to stabilize the anode pressure. The fluctuation of differential pressure between cathode and anode (up to 2 psid at pressurized condition) caused or accelerated the reactants crossover in the 3-cell stack and damaged cells due to internal combustion. This issue has been resolved by retuning the pressure regulator and setting new control parameter for the pressure regulator.

Figure 75 Individual Cell Performance of 3-Cell (12'') Half Seal Stack, R14 with 64% H₂ and Fixed Flow at 800°C

3 SOFC SCALEUP FOR HYBRID AND FUEL CELL SYSTEMS

See Appendix A

4 INTEGRATED GASIFICATION FUEL CELL SYSTEM STUDY

See Appendix B

5 INTEGRATED GASSIFICATION FUEL CELL SYSTEM EXTENSION STUDY

See Appendix C

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Appendix A

Solid Oxide Fuel Cell Hybrid System for Distributed Power Generation

SOFC Scaleup for Hybrid and Fuel Cell Systems
Final Topical Report

April 2004 to September 2004

David Deangelis

Rich Depuy

Debashis Dey

Georgia Karvountzi

Nguyen Minh

Max Peter

Faress Rahman

Pavel Sokolov

Deliang Yang

October 2004

Performed under DOE/NETL Cooperative Agreement

DE-FC26-01NT40779

GE Hybrid Power Generation Systems

19310 Pacific Gateway Drive

Torrance, CA 90502

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Abstract

This report summarizes the work performed by Hybrid Power Generation Systems, LLC (HPGS) during the April to October 2004 reporting period in Task 2.3 (SOFC Scaleup for Hybrid and Fuel Cell Systems) under Cooperative Agreement DE-FC26-01NT40779 for the U. S. Department of Energy, National Energy Technology Laboratory (DOE/NETL), entitled "Solid Oxide Fuel Cell Hybrid System for Distributed Power Generation". This study analyzes the performance and economics of power generation systems for central power generation application based on Solid Oxide Fuel Cell (SOFC) technology and fueled by natural gas. The main objective of this task is to develop credible scale up strategies for large solid oxide fuel cell-gas turbine systems. System concepts that integrate a SOFC with a gas turbine were developed and analyzed for plant sizes in excess of 20 MW. A 25 MW plant configuration was selected with projected system efficiency of over 65% and a factory cost of under \$400/kW. The plant design is modular and can be scaled to both higher and lower plant power ratings. Technology gaps and required engineering development efforts were identified and evaluated.

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Executive Summary

The Solid Oxide Fuel Cell (SOFC) technology is regarded as one of the most promising future power generation technologies and perceived to have a range of advantages over competing technologies. The study presented in this report explores the possibility of the SOFC technology to challenge the technologies that dominate the central power generation market, the combined gas turbine-steam turbine cycle and the traditional coal powered steam plant. A minimum plant size of 20 MW was chosen for this study. The plant uses natural gas as fuel and delivers AC power at the grid voltage. The objective of the SOFC scale up task is to develop credible scale up strategies for large SOFC-gas turbine (GT) hybrid systems, and particularly, to understand the effects of system and stack architecture on plant scale up and performance.

The study identified the product requirements for a 20 MW central generation power plant based on the standards of competing GE products. The system is also required to meet or exceed the Solid State Energy Conversion Alliance (SECA) factory cost goal of \$400/kW and to target system efficiencies in the 55-75% (LHV) range.

The large system size requirement provides an opportunity to integrate SOFC stacks with a gas turbine to achieve high system efficiency. In a hybrid with an SOFC, the gas turbine extracts useful work and generates power from the SOFC by-product heat, which leads to system efficiencies unmatched by either SOFC or gas turbines alone. A large number of system and stack concepts were generated in the program to address issues associated with the integration of SOFC stacks with a gas turbine. Four concepts were selected for conceptual analyses.

The four concepts were analyzed for system efficiency, factory cost, and reliability. The analysis results were compared to the product specification, and Concept 1 was selected because it was projected to have the best chance of meeting the product requirements. Concept 2 was chosen for a risk mitigation strategy.

The down selected concept was further analyzed. A 25 MW plant design was completed, and its performance, factory cost and projected reliability were estimated. The design first focused on the SOFC stack subsystem architecture. The study determined that a 19.9 MW SOFC stack subsystem operating at a pressure of 5 atm would be required for the 25 MW plant. The subsystem architecture included multiple stacks placed in pressurized modules. An individual stack size was determined through cost and reliability studies while factoring in power electronics constraints.

The study identified that the size of individual cells in a stack and the number of cells in a stack can be determined independently of each other. The cell size is found through a stack cost and reliability optimization. Using projections of the circular planar cell manufacturing technology capabilities, the stack cost as a function of cell diameter was found to be minimized over a wide range of cell diameters, 30-60 cm (12-24 inches). Reliability considerations favor large cell diameters and a smaller cell count. However the cell reliability may decline rapidly with the cell diameter thus placing a constraint on the cell size. Due to the lack of available reliability data on large cells operating in a

stack, the cell size constraints were not quantified, and instead the cell size of 45.7 cm (18 inches) in the middle of the cost-minimizing cell range was chosen for the remainder of this study. The number of cells in a stack was determined through the optimization of the SOFC power conversion subsystem. The optimal maximum stack voltage of 400V was identified, which translates into the optimal number of cells per stack of 400. Given the cell size and the number of cells per stack, the optimal stack building block for the 25 MW plant was estimated to have a nominal power rating of about 320 kW.

Plant level reliability analyses were performed on the 25 MW hybrid plant including the selected plant architecture and the stack building block. This analyses identified that an operability scheme with eight active and two redundant modules each containing eight stack building blocks would achieve the reliability targets of the product specification. Therefore, pressurized modules of eight stacks were chosen for the 25 MW plant SOFC subsystem architecture.

The 25 MW plant factory cost and performance analyses were conducted next. The factory cost was determined to be about \$7.9M, or \$317/kW, compared to the product specification target of \$400/kW. The calculated plant efficiency of 66% on natural gas also exceeds the product specification target of 65%. The plant projected layout shows that the projected plant footprint area is about 1580 m² (17000 ft²), about twice the size of comparable combined-cycle plants. The plant design is also suitable for operation with coal gas with minimal design modifications. With coal, plant power was de-rated to 18 MW, and the efficiency on coal gas was determined to be about 69%. This translates into a gasified coal plant efficiency of 48-55% after factoring in the efficiency of a gasifier.

The 25 MW hybrid plant design developed in the study can be scaled to both higher and lower power levels using the same stack building block. Power plants of higher size can be built from the same eight-stack pressurized modules used in the 25 MW plant design. A 250 MW hybrid plant was projected to have 67.6% plant efficiency and \$260/kW plant factory cost. Both parameters exceed the 25 MW plant capabilities due to plant economies of scale. A 5 MW hybrid plant can use the same 320 kW stack building blocks as the 25 MW hybrid plant, however the stacks would have to be arranged in smaller pressurized modules, with two stacks per module. The 5 MW hybrid plant projected efficiency and cost were projected to be 65.1% and \$512/kW respectively.

In conclusion, the study developed a 25 MW SOFC-gas turbine hybrid power plant design for central power generation on natural gas. The projected plant performance, cost and reliability exceed the product specifications. The plant design is modular and can be scaled to higher and lower power ratings. To realize this plant concept, several technology goals must be met, primarily in the SOFC stack, including the cell scale-up and high-temperature stack seals. In addition, the turbomachinery and balance of plant (BOP) components will require significant re-engineering.

Experimental

No experimental work was performed as part of this task.

Results and Discussion

6 INTRODUCTION

6.1 Background

The Solid Oxide Fuel Cell (SOFC) technology is regarded as one of the most promising future power generation technologies. The SOFC technology is perceived to have a range of advantages over competing technologies including high electrical efficiencies, low emissions, modularity, potential for low cost, etc. SOFC-based products are considered for a variety of applications, such as distributed power generation, auxiliary power units for transportation, combined heat and power for residential and commercial use. Most of these applications focus on products of low power rating due in large part to the challenges of scaling up existing development efforts to larger sizes. The study presented in this report explores the possibility of the SOFC technology to challenge the technologies that dominate the central power generation market, the combined gas turbine-steam turbine cycle and the traditional coal powered steam plant.

The products that the study targets will have to be an integral part of the existing electrical transmission and distribution system. This system was designed to take advantage of the steam power plant economies of scale, which drive the power plant size requirements into tens and even hundreds of megawatts. As the transmission and distribution system is an infrastructure of an enormous size that is unlikely to be modified, the SOFC products for central power generation must have a similar power-rating requirement. A minimum plant size of 20 MW was chosen for this study.

6.2 Program Objectives

The objective of the SOFC scale up program is to develop credible scale up strategies for large SOFC- GT hybrid systems. In particular, the task is designed to understand the effects of system and stack architecture on plant scale up and performance.

The task focuses on three inter-related areas: system architecture, stack building block, and cell size. System architecture has a significant impact on the SOFC subsystem and therefore on scalability and modularity of the SOFC stacks. Consequently, after determining the top-level plant requirements, several system architectures or concepts are to be considered. These concepts are then evaluated based on their ability to meet the plant requirements and their implications on the stack. A down selected system concept is to be used for further analysis.

Subsequently, plant performance, cost, and reliability are considered on the down selected concept to evaluate the optimum cell size. Simultaneously, cell size ranges that allow plant requirements to be met are determined.

Finally, the above analysis is integrated to ascertain whether a common stack building block suitable over a range of plant sizes can be identified. The optimum building block

stack is estimated and the range of plant sizes over which this building block is applicable is evaluated.

The above approach necessitates a plant product focus. This study is limited to plant sizes in excess of 20 MW that are applicable to central station power generation applications. The system is also required to meet or exceed the Solid State Energy Conversion Alliance (SECA) factory cost goal of \$400/kW and to target system efficiencies in the 55-75% (LHV) range.

Since SOFC stack and system scalability strategies are driven by the functional requirements of the end user, in this case the power generation industry, the first task in this study is to develop a functional product specification that would form the basis for all subsequent trade studies. Within the bounds of this product specification, system architectures that exploit planar SOFC technology are identified, and the SOFC stack scale up and technology development risks are discussed. For example, the minimum cell size and modular stack building block size are estimated based on performance, cost, and reliability considerations. Conversely, the applicable plant power range is estimated for a particular stack building block size. Finally, stack technology gaps are articulated, and a top-level technology development strategy is developed.

6.2.1 Product Specification

As mentioned above, the approach to this task is product focused. This focus is driven through the use of a functional product specification that was drafted at the start of the task and maintained through the duration of the task.

The functional product specification is a document that lists all the necessary top-level technical requirements of the plant. The product specification consists of a list of variables that are critical to the quality (CTQ) of the plant. The range over which each requirement can vary is also provided.

The product specification is typically determined by interacting with the end user or customer of the plant. Since this approach was not practical for this study, the product specification was established based on the performance standards of competing GE products. The product specification was also further divided into two sections. The top section, titled "Contractual Requirements" lists the plant attributes that are necessary to meet the contractual requirements of this task. In particular, the three attributes in this section included the power output, the plant efficiency, and the plant manufactured cost. The second section, titled "For Information Only" lists the plant attributes that are seen to be necessary for the plant to be competitive and successful in the targeted market but which are used in this study for guidance purposes only. Due to scope limitations of this study, system analysis is restricted to a conceptual level, and several attributes in this section cannot be determined with low variability at this level of analysis. They are listed to ensure that system concepts that are considered and down selected have the capability of achieving levels of performance listed in this section of the product specification.

As mentioned above, centralized power applications are the primary target of this study. This application is particularly pertinent to fuel cell studies since it necessitates base-loaded operation at low cost or high system efficiencies.

7 SYSTEM TRADE STUDY

7.1 System Concept Selection

System architecture has a significant impact on the design of the SOFC stack and consequently, on the scalability and modularity of SOFC stacks and hybrid systems. The system concept must be selected to maximize the advantages of integrating SOFC stacks with gas turbines while satisfying the product requirements. Therefore, the system concept selection process started with brainstorming concepts having high probability of satisfying the product requirements. The initial long list of concept ideas was then reduced to a few most promising concepts based on top-level system analyses and engineering judgment. The remaining concepts were then analyzed in more detail to determine the final hybrid plant conceptual design that has the highest potential to satisfy the product requirement. The plant conceptual design was then used as a starting point of plant-level trade studies to find an optimal plant design solution.

7.1.1 Initial Concepts

SOFC stacks operate at high temperatures, generally above 650°C, due to the ionic conductance characteristics of the electrolyte. A mixture of hydrogen, carbon monoxide and methane can be used as the fuel, while air usually supplies oxygen. Though very efficient power producers, SOFC's still generate much by-product heat that needs to be removed to avoid overheating the fuel cell. In high-temperature fuel cells, such as the SOFC, systems are normally designed so that the by-product heat is removed with airflow through the fuel cell. Usually, the cooling requirement imposed on the airflow results in a much higher airflow rate than that required for the fuel cell reaction, due to the poor heat transfer characteristics of air and, equally importantly, the inability of the SOFC stack to withstand large temperature rise from stack inlet to stack exhaust due to thermal stresses. The presence of the large temperature gradients is detrimental to both structural integrity and reliability of the stack.

Therefore, the stringent heat rejection and SOFC thermal gradient constraints result in high airflow requirement and a necessity to preheat the air to a temperature nearly equal to the stack temperature before it enters the stack. Efficient, reliable and inexpensive airflow thermal management is thus the key to SOFC system design. Other important considerations may also affect the system design, such as the choice of the fuel pre-reformer and its thermal integration as well as water management.

The system size is an important factor in system design. Technological limitations for both the cell size and number of cells in a stack, discussed later in the report, result in a feasible stack power in order of hundreds of kilowatts, while the system power requirement is above 20 MW. Hence, a number of stacks, rather than a single stack, have to be integrated into the system. This fact poses both a challenge and an opportunity to the system designer. On the one hand, the air and fuel flows between

the multiple stacks have to be managed to achieve high system efficiencies while maintaining reasonable system cost and reliability. On the other hand, the system may be designed to offset the air preheat and heat rejection requirements between multiple stacks, thus enabling high system efficiencies.

The large system size requirement provides an opportunity to integrate SOFC stacks with a gas turbine to achieve high system efficiency. As the SOFC stacks convert the chemical energy of fuel to electrical energy, it generates by-product heat that can further be converted to electricity. In the SOFC simple cycle, this thermal energy is rejected as waste heat. In a hybrid system, however, the thermal energy is recovered within the gas turbine generator, which converts thermal energy into electrical energy. A 4-5 MW gas turbine would provide a good match for a 20 MW hybrid system design. Gas turbines in this power have reasonable component efficiencies and are available commercially. Therefore, system designs considered in this study were focused on hybrid designs due to their high efficiency potential.

A large number of system and stack concepts were evaluated to address issues discussed above. Some of the ideas explored are noted as follows

- Due to the challenging system cost target, system design simplicity was stressed during concept evaluation. Concepts with multiple gas turbines and complex air preheating schemes were eliminated early due to their limited low-cost potential.
- SOFC stacks can be arranged either parallel or in series with respect to the airflow. In the parallel flow arrangement, flow is split equally among all the stacks. In the series or staged flow arrangement, the exhaust of one stack is the inlet to a subsequent stack. The staged arrangement potentially results in higher system efficiencies with a small cost penalty, as the by-product heat of the upstream stage is used to preheat the subsequent stack air inlet. A combination of staged and parallel arrangements is also possible.
- Solid-carbon-producing reactions limit the extent of internal reforming. When the ratio of molar concentrations of carbon and water reaches a certain limit (that depends on pressure and temperature), the chemical equilibrium shifts towards solid carbon, severely limiting the SOFC performance.
- Similarly, SOFC stacks can be arranged either parallel or in series with respect to the fuel flow. A staged arrangement may result in a high fuel utilization in the stacks even when each stack's fuel utilization is low, thus resulting in a high system efficiency.
- Both recuperated and un-recuperated systems were considered. A recuperator placed at the turbine exhaust to heat the gas turbine compressor outlet may improve the system efficiency. It is unlikely that the recuperator by itself can heat the compressor outlet to the SOFC operating temperature. Therefore, additional air preheat means is required.

- The additional means of air preheat to the SOFC operating temperature can be accomplished by burning additional fuel in the SOFC air inlet or using the by-product heat of the SOFC reaction.
- Different kinds of fuel reformers were also considered, namely steam reformers, partial oxidation fuel processors and auto-thermal reformers.
- Steam is required for fuel processing (both steam reformers and auto-thermal reformers). There are three options for steam supply: (1) outside water supply with steam generation; (2) a water pump with a steam generator providing steam from a water tank with a condenser at the system exhaust; and (3) recycling of the SOFC anode outlet containing product water to the reformer. Option 2 and 3 maintain water neutrality and are preferred if the plant site has fresh water supply limitations.

Over twenty system concepts were generated using the ideas outlined above. These concepts were ranked consistent to the criteria outlined in the product specification using simple conceptual analyses and engineering judgment.

7.1.2 Results of Initial Screening

The results of initial screening analyses and some general observations are outlined below:

- A considerable amount of internal reforming within the SOFC stacks is desirable to achieve high system efficiency. Internal reforming occurs within the SOFC stacks when methane is present in the SOFC stacks fuel inlet. The reaction product water can be used for converting the methane to hydrogen and carbon monoxide. The methane reforming reaction is endothermic and absorbs the SOFC reaction by-product heat, thus lowering the gas turbine airflow requirement and improving the system efficiency. The fuel processor does not have to convert 100% of the inlet fuel to hydrogen and carbon monoxide and therefore becomes a pre-reformer.
- Steam reformers require no airflow and therefore result in higher system efficiency than that of systems that contain partial oxidation and auto-thermal reformers. The power required to pressurize the pre-reformer airflow reduces the auto-thermal-reformer-based system efficiency by about 5 percentage points below the steam-reformer-based system. Similarly, partial oxidation reformers result in a 10-percentage point efficiency disadvantage compared to steam reformers. Since there is no clear cost or reliability advantage for any of the reformer types, steam reformers are a better choice for hybrid systems.
- Air thermal management design can drive the SOFC stack design requirements, and vice versa.
- There is likely a limit of the number of SOFC stack stages in staged arrangements. This limit may be driven by either the minimum pressure drop

through SOFC stacks or the minimum reactant concentration required for stack operation.

7.1.3 Down Selected Concepts

Considering the initial screening observations outlined above, four concepts were selected for further analyses to identify the system with the best chance of meeting the product requirements. All concepts incorporate multiple SOFC stacks with a steam reformer and a gas turbine (Figure 76). The gas turbine compressor supplies pressurized air to the SOFC stacks. The air is preheated to the SOFC operating temperature. All systems are recuperated. Steam reformers partially convert the fuel to a hydrogen-containing gas (so-called reformat) and supply it to the SOFC stacks. The differences between the concepts are in the ways each of them accomplishes air pre heat and water management.

Figure 76 Generic SOFC-GT hybrid system

7.1.4 System Analysis Approach

7.1.4.1 Performance Modeling

Thermodynamic performance models were developed for all four concepts using Aspen PlusTM, a thermodynamic and chemical processes analysis tool. The thermodynamic performance models calculate the heat and material balances around each system component with appropriate component performance parameter assumptions and integrate the components into the system flow sheet. The performance model outputs are the calculated system efficiency and component performance specifications. The latter serve as the inputs to the system cost and reliability models. The component performance assumptions are outlined below.

7.1.4.1.1 SOFC Stacks Performance Assumptions

The SOFC stack performance assumptions are based on GE's experience in the development of SOFC stacks. The assumptions are as follows.

- SOFC stacks are constructed out of planar fuel cells. The cells are ceramic tri layers with metallic interconnects and flow fields.
- The maximum air temperature rise through the stack is fixed. This assumption drives the airflow requirement through the stack. The air temperature rise is measured from the stack inlet air manifold to the stack outlet air manifold. The air temperature rise is usually limited by the cell temperature gradients. The relationship between the air temperature rise (or the airflow) and the maximum allowable cell temperature gradient is highly dependent on the stack and cell design.
- Maximum SOFC air outlet temperature is kept at a fixed value.
- Single-cell voltage at full load is 0.7V. The single-cell voltage drives the SOFC stack and the overall system efficiency. The single-cell voltage must normally be traded with the SOFC power density (or current density) to achieve the most favorable system efficiency-system cost balance. This trade-off requires knowledge of the SOFC stack polarization curve, showing the average single-cell voltage as a function of average current density. This polarization curve is a reflection of the cell technology and stack design. Rather than predicting what the stack polarization curves will be at the time when the SOFC technology matures to the point that a 20MW+ plant sizes are feasible, we chose in this study to fix the average cell voltage and determine the required power density to achieve the system cost goal. The single-cell voltage of 0.7V at full load is a reasonable assumption based on GE's previous SOFC system designs and other benchmark activities.
- SOFC subsystem air pressure drop is 5% of the total system pressure. The SOFC subsystem includes the SOFC stacks and the associated valves and piping.

- SOFC subsystem heat loss is equal to 2% of the SOFC power output. This is a relatively conservative assumption for a large system. This assumption will be revised in the detailed plant design phase.
- The concepts proposed in this study consider two types of SOFC stacks: stacks that use cells with cathode-to-anode exhaust seals, and stacks designed without these seals. We assumed that stack performance assumptions do not differ between the sealed and seal-less stacks. The stack designs for sealed and seal-less cells could potentially be drastically different. However, there are two reasons to believe that the performance assumptions should be identical for both stack types from the system perspective. First, while there can be significant differences in performance entitlements between the two types (for example, the fuel utilization entitlements), the performance assumptions made in this study are likely to be sufficiently below the entitlements that identical assumptions should be made. Second, the stack designs at high power levels considered in this study have not been developed. It is unclear which stack type will result in a higher entitlement for each particular assumption, therefore an identical assumption for both cell types should be made.
- The assumed baseline power density as a function of pressure was assumed. An assumption was made that the power density at 5 atm is 0.5 W/cm².

7.1.4.1.2 Gas Turbine Assumptions

Gas turbine compressor and turbine efficiencies are assumed to be functions of component size. For recuperated cycles, the recuperator operating conditions in the hybrid cycles under consideration are likely to be similar to those in the gas turbine cycle. Therefore, the recuperator effectiveness and pressure drops through the cold and hot sides are assumed to be similar to the values observed in typical gas turbines.

7.1.4.1.3 Pre-Reformer Assumptions

There are several options for sizing the fuel pre-reforming subsystem: (1) one central pre-reformer feeding all stacks in the system; (2) one pre-reformer per group of stacks; or (3) one pre-reformer per stack. The most appropriate configuration should be determined through performance-cost-reliability trade offs. Other assumptions are as follows.

- All pre-reformers are steam reformers.
- Pre-reformers are assumed to be in chemical equilibrium.
- Pre-reformers require minimum amount of steam in feed that corresponds to a steam-to-carbon (S/C) ratio of 1.5. The S/C ratio is defined as (mole flow of water in the pre-reformer feed)/(mole flow of carbon atoms in the pre-reformer feed). For example, the S/C ratio is $S/C = \frac{M_{H_2O}}{M_{CH_4} + 2 \cdot M_{C_2H_6} + M_{CO} + M_{CO_2}}$ in the case of the pre-reformer feed being a mixture of steam, methane, ethane, carbon monoxide, and carbon dioxide. This requirement ensures that reactions that can

lead to carbon deposition in the pre-reformer are prevented. The minimum S/C value of 1.5 is consistent with empirical data.

7.1.4.2 Cost Analysis Approach

The scope for all the costing is manufactured cost and not installed cost. A manufactured-cost model has been developed for the four system concepts presented earlier. The system cost model consisted of component cost models that were parameterized to allow analyses of component costs with respect to their duties. This modeling approach makes system cost model parametric to allow studies of the effects of system design and size on cost. It was assumed that all the system components, with the exception of the SOFC stacks, are acquired at their market prices at a volume corresponding to an annual production volume of 50 system units per year. The SOFC stack cost is analyzed separately with an SOFC stack cost model that is described in Section 4.2.3.2. Since the component cost models are duty-based, sensitivity studies, such as the effect of system pressure, effect of amount of recycle, various levels of internal reforming, etc., can be performed.

7.1.4.3 Reliability Analysis Approach

The approach to modeling the system reliability is to develop component roll-up tools, followed by detailed redundancy studies. For the purposes of trade studies between the concepts presented above, the same redundancy scenario was selected for all concepts. The optimal reliability scenario was then determined by system reliability and cost trade studies on the down selected concept.

Mean Times Between Failures (MTBF) and Mean Times To Repair (MTTR) numbers for all the system components, except the SOFC stack, were collected from published data (ref. 1). Required SOFC stack MTBF and MTTR numbers are outputs of the analysis that are found from the condition of satisfying the availability and reliability targets. Component MTBF and MTTR numbers were then combined on the system level into the plant MTBF and MTTR using the selected redundancy scenario.

A reliability analysis tool for the hybrid SOFC plant has been developed. The model was designed to serve as a trade-off tool, so that the sensitivity of plant reliability to various design parameters could be studied.

7.1.4.3.1 Approach

A two-part modeling approach was undertaken:

1. A detailed model was set up that could investigate the effect of stack arrangement, part load, maintenance intervals etc. This model concerned only the SOFC stack subsystem.
2. A roll-up tool was developed that included the stack assembly results from the first model, combined that with reliability data of the Balance Of Plant (BOP), and calculated the plant reliability and availability as a result.

The first model was done using the commercial software BlockSimTM while the second model was implemented in FORTRAN.

7.1.4.3.2 Definitions and assumptions

The term reliability is generally used to indicate the ability of a system to continue to perform its intended function (ref. 1). The key parameters in defining system reliability are MTBF and MTTR. The MTBF is defined as the mean exposure time between consecutive failures of a component. The MTTR is defined as the mean time to repair or replace a failed component.

The following are the major assumptions in the analysis.

- Only the principal failure mode of each component is considered
- Effects of specific failure modes are not investigated
- The failure rate is constant
- Failure is defined by stack outage, not de-rating

The last bullet requires explanation. In the context of reliability, failure needs to be adequately defined. For the purpose of this study, failure was considered at two levels: (1) the failure of the plant, and (2) the failure of the individual stacks that make up the SOFC subsystem of the plant. The two are not necessarily the same, since the plant could have redundant stacks, and thus the failure of one stack does not cause a plant failure. Plant failure is defined as the inability of the plant to provide the output power, within a margin of the rated power (25 MW). The margin is not explicitly defined in this study. It should be noted that for the purpose of this study, the plant power is being guaranteed, and not the plant efficiency.

Stack level failure is harder to define. It is a well-known fact that fuel cells degrade in power over their lifetime, and SOFC is no exception. The level of degradation, usually expressed as a percentage of power loss for every 1000 hr operation, is a known quantity for any specific fuel cell design. Since it is known, within a margin of uncertainty, degradation is accounted for in the plant operation. Certain failure modes might cause the stack to produce lesser power than what the degradation schedule allows for. If this power reduction is small in magnitude, it will not warrant repair till the next scheduled repair opportunity. However, if the power reduction is large then immediate attention will be warranted. This last event is termed a stack failure.

Figure 77 summarizes this effect. The blocks to the right entitled “fmN” refer to the probability of individual failure modes. In a well-designed stack the failure modes that cause outage will have very small probabilities.

Figure 77 Stack failure definition

7.1.4.3.3 BOP Reliability Assumptions

The data shown in Table 28 have been used for reliability calculations. The data source is the standard IEEE STD 493 –1997. The SOFC stack mean time between failure (MTBF) data are scarce. An MTBF of 13,400 hrs has recently been reported (ref. 2). The MTBF numbers for SOFC stacks are expected to improve as the technology is developed. In this study, SOFC stack MTBF was varied to achieve the required reliability and thus is considered a study output.

Components	MTBF (hrs)	MTTR (hrs)
SOFC Stack	Variable	150
Pre-reformer	40000	150
Pipe inside pressure vessel	40000	36
Hot pipe	40000	36
Cold pipe	40000	36
Hot valve	40000	72
Cold valve	40000	72
Inverter	200000	126
Transformer	200000	130
Step - up transformer	200000	130
Mass addition	50000	72
Heat exchanger	40000	72

Table 28 Expected mean time between failure (MTBF) for system components

7.1.5 Results of Analyses of Down Selected Concepts

Concepts 1-4 were analyzed for performance, cost, and reliability with the goal of selecting the best concept based on these three measures. The analysis used the parameterized performance, cost and reliability models to identify the concept with the best chance of meeting the product requirements. The analyses assumed that the initial hybrid plant power is 25 MW to account for possible power degradation.

The comparison function for each concept was formed based on the analysis results:

$$S = \frac{\eta}{0.65} + \frac{\$400 / kW}{C} + \frac{R}{0.985} + \frac{MTBF}{4380hrs}$$

where S is the dimensionless concept score used to compare concepts against each other, η is the system efficiency, C is the concept manufactured cost in \$/kW, and R is the system reliability. Each concept was then optimized with respect to the system operating pressure to find the maximum score S, and the system with the highest score was then selected for further plant optimization analyses.

Analyses showed that the four concepts optimize approximately at the same system operating pressure of about 5 atm. This result is expected because similar recuperated gas turbine cycles optimize in efficiency at about 4-5 atm. The addition of manufactured cost as an object function to the optimization problem shifts the optimal value of the operating pressure to the upper end of this range.

Table 29 below shows the results of the concept analyses. These results will be used to select one concept for the plant optimization analyses.

Parameter	Concept 1	Concept 2	Concept 3	Concept 4
System Power (kW)	25000	25000	25000	25000
Fuel Cell Power (kW)	20315	18656	19811	19959
Gas Turbine Power (KW)	6083	7557	5664	5348
System Efficiency, η , %	65.53	60.54	67.19	66.69
Cost	\$13,385,686	\$13,990,218	\$13,633,862	\$13,551,625
Specific Cost, C , \$/kW	\$535	\$560	\$545	\$542
Reliability, R	94.75%	94.75%	94.75%	94.75%
MTBF, hrs	1258	1258	1182	1146
Total Score, S	3.0045	2.8954	2.9990	2.9876

Table 29 Results of preliminary analyses of selected concepts

7.1.6 Final system concept selection results

Based on the assessment and analysis of the various concepts, Concept 1 was selected for further planned design.

7.2 Hybrid System Design

7.2.1 Plant description

The system concept down selected for plant design consists of the following subsystems:

- Gas Turbine;
- SOFC Stack;
- Fuel Delivery;
- Thermal Management;
- Water Management;
- Power Electronics and Controls.

SOFC Stack Subsystem is made up of several modules. A module contains SOFC stacks, pre-reformers, and the associated piping and valves. The modules operate at an elevated pressure, and therefore they must meet the requirements of pressure vessels.

The main characteristics of the conceptual design presented are as follows:

- A number of steam reformers for fuel pre-reforming;
- A recuperator is used to transfer heat from the turbine exhaust to the compressor exhaust.

During the detailed analysis phase, this conceptual design was optimized for performance, cost and reliability.

The system lay out is shown in Figure 78. The approximate footprint area of the plant is 1580 m² (17000 ft²). The footprint is about twice the area of a combined cycle plant with a comparable power rating.

Figure 78 Twenty-five MW SOFC-GT hybrid plant layout

7.2.2 25 MW plant performance projections

The plant conceptual design is based on Concept 1. Detailed plant design is an iterative process because the outputs of the performance analysis serve as the inputs to the cost and reliability optimization studies, however the results of the cost and reliability optimization studies can also have an effect on the plant performance analysis through changing pressure drop, heat loss and other performance parameters. This subsection presents the final plant performance analysis results.

The Aspen PlusTM plant performance model developed in the previous sections was again used in the plant performance analyses. The performance assumptions for a 25 MW plant are listed in Table 30 below. The assumptions are similar to those defined during the concept down selection process. One notable exception is the SOFC heat

loss assumption that was re-calculated given the results of the module sizing analyses. The updated heat loss calculations given the SOFC stack module size and assumed levels of thermal insulation resulted in about 100 kWth of total system heat loss, which is about 0.4% of the system power and is below the 2% value assumed in the system down selection calculations (the change would not have had any effect on the concept down selection results).

Variable	Value	Comment
Cell Design	Planar	
Single-cell voltage at full load	0.7V	Will not be optimized through performance-cost analyses because stack polarization curve is unknown
Stack fuel utilization	70%	
System operating pressure	Optimized	
SOFC subsystem pressure drop	5% of system pressure	
Total system heat loss	100 kWth	Computed based on projected module size and length of hot piping

Table 30 Performance model assumptions

The results of the pressure optimization study are shown on Figure 79. Note that the maximum efficiency occurs at the system pressure of between 5 and 6 atm. The combined system efficiency-cost optimization leads to approximately the same optimal pressure, as the cost gains from the fuel cell cost improvements are almost offset by the higher cost of balance of plant.

The maximum system efficiency of the 25 MW plant is equal to about 66%. Figure 79 also shows the effect of the SOFC subsystem pressure drop assumption on the system efficiency. The pressure drop assumption drives the SOFC stack flow field design and, as Figure 79 illustrates, has a major effect on system efficiency. In fact, if the stack pressure drop is reduced to 1.5% of the total system pressure then the maximum efficiency point moves to a slightly lower pressure, between 4-5 atm, and the maximum efficiency is 68%, almost 2 percentage points higher than the 5% pressure drop case. The uncertainty of the optimal pressure selection is quite small, and it appears that the optimal pressure is around 5 atm. Hence, we chose this value as the system operating pressure. Additionally, electrical power loads produced by the system power producers (SOFC stacks and the gas turbine) and consumed by major power parasites are shown in Table 31.

Figure 79 Results of 25 MW plant performance optimization

Component	POWER PRODUCED (CONSUMED), KW
SOFC Stacks	19941
Gas Turbine	7779
Blower	(1554)
Natural Gas Pump	(1159)
Net	25008

Table 31 Power produced and consumed in the 25 MW plant

7.2.3 25 MW plant SOFC building block design

The SOFC subsystem of the 25 MW plant design has a power output of 19.9 MW as shown in Table 31. It is impractical from a cost and reliability standpoint to build a 19.9 MW SOFC stack. Therefore, multiple stacks will be involved in the system configuration. The following chapter discusses an approach to determine the optimal number of stacks in the 25 MW plant and the associated stack power.

7.2.3.1 Approach

The approach to determining the optimal number of stack and the stack size undertaken in this study looks at stack size effects on the performance, cost and reliability of the plant. A bottoms-up analysis of the stack performance, cost and reliability results in an explicit dependence of the system cost, efficiency and reliability on the cell size, and number of cells in the stack. An optimization problem is then set up to minimize cost and maximize efficiency and reliability by varying the cell size and the number of cells. The stack size is the solution of this optimization problem.

Note that this problem appears to be under-defined, as three dependent variables are optimized with only two dependent variables. This is a reflection of the fact that system level parameters cannot be optimized with just stack level variables, as the rest of the system will affect the system level dependent variables. For example, stack grouping methods and redundancy scenarios will also have an impact on the system cost and reliability. In this study, some system level constraints will be unknown or uncertain due to the uncertainty of many assumptions and available data. Therefore, the solution for the stack size is not necessarily a fully optimized solution but rather an approximate solution and a guide to future system design efforts.

7.2.3.2 SOFC stack cost model

The cost model for this program was based on SECA stack cost model. This a complete cost estimation tool that uses a series of performance inputs and design assumptions and generates a breakdown of materials, equipment, labor and facilities costs associated with SOFC stack manufacturing.

The stack cost model has been designed to accommodate sensitivity analyses through flexibility, modularity and user-friendliness. It is therefore an easily modified tool, where progresses can be recorded as the design matures. Ultimately, a manufacturing cost, including equipment, labor and facility costs, is generated and is added to the materials cost to yield the total stack costs.

7.2.3.2.1 Stack Configuration

The GE SECA conceptual stack configuration was assumed in the study.

7.2.3.2.2 Assumptions

Material cost assumptions were based on DOE recommendations for the SECA program stack cost. **Error! Reference source not found.** The stack manufacturing process was divided into two parts: (1) cell manufacturing, and (2) stack assembly. Yields were assigned to both sub-processes. Cell manufacturing yield is assumed to vary from 90% to 60% for a cell size variation of 15-50 cm (6-20 inches), Figure 80. These values are based on typical yields of mature ceramic manufacturing processes observed in the industry. Therefore, the cell yield assumptions are subject to the condition that the cell manufacturing process has a similar entitlement to existing ceramic processes.

Generally, ceramic process manufacturing yields are usually described with a Weibull distribution similar to the blue line on Figure 80. The cell manufacturing process yield

however may not be accurately predicted with the Weibull distribution, and a more conservative yield distribution was chosen. Because of the uncertainty of the yield assumption, the impact of a lower yield was also examined.

Stacking yield was varied from 95% to 75% for the same cell size variation, Figure 81. Similarly, it was assumed that a straight line describes the yield distribution as a function of cell size. Sensitivity to a lower stacking yield assumption was also analyzed.

Figure 80 Cell manufacturing yield assumption

Figure 81 SOFC stacking yield assumption

7.2.3.3 Optimal cell size selection

Stack cost analysis results showed that the stack cost is a strong function of the cell size, Figure 82, in the 15-30 cm range. This result is intuitively obvious, since the cell-manufacturing yield (Figure 80) is expected to be a strong function of cell size. However, stack cost is a weak function of the number of cells in the stack, if the number of cells is over 50, because the cost of non-repeat parts in the stack, such as current collector plates, is much lower than the cost of repeat parts, such as cells, manifolds and flow field assemblies.

This is an important result to optimal cell size selection because it leaves the stack cost a function of the cell size only. Since the system cost is a weak function of number of stacks, as will be shown later in chapter 4.2.6, the system cost is also only a function of cell size and not number of cells in the stack. Therefore, the problem of determining the optimal cell size reduces to a simple minimization problem of the stack cost. Note that system reliability considerations may also become

Figure 82 shows that the stack cost has a minimum around cell diameter of 50 cm (about 20 inches). In fact, the stack cost function is relatively flat in the 30-60 cm (12-24 inches) region. Varying yield assumptions reduces optimal cell diameter values to about 40 cm (about 16 inches). Since the yield assumptions are highly uncertain, a range of acceptable cell diameters should be specified rather than an optimal value. From Figure 82, the 30-60 cm (12-24 inches) range appears to be acceptable, since the stack cost function is flat in this region for a variety of yield assumptions.

The reliability analyses discussed below favor larger cells and therefore, a smaller cell count in the system. This result argues for the selection of cells of 60 cm in diameter or

even larger. However, cell reliability as a function of cell size is not well understood, and a safer cell size should be chosen. In the following system cost and reliability projections, a cell diameter of 45.7 cm (18.0 inches) was assumed in all calculations for illustration purposes.

Figure 82 SOFC stack cost as a function of cell diameter

7.2.3.4 25 MW power electronics subsystem description

7.2.3.4.1 Power electronics design

For the design of stationary, utility-grade power conversion equipment, efficiency and availability are the predominant design requirements. The power conversion system linking the fuel cell stacks and the grid provides voltage and frequency transformation, galvanic isolation, and control over the power flow, but excludes grid interconnection. For a 20 MW SOFC stack subsystem as part of the hybrid 25 MW power generation plant, the low voltage, high current DC power provided by the fuel cell stacks must be converted to a balanced, three-phase medium voltage AC power. Uni-directional DC-to-AC power conversion is required. The specific electrical characteristics and requirements of the SOFC and utility grid must be considered while designing the conversion system. Since only base-loaded applications are being considered, energy storage is not required. The power converters feature small energy storage elements to balance instantaneous fluctuations in the equilibrium of input power and output power.

7.2.3.4.2 Semiconductor Technology

The choice of the stack voltage level will affect primarily semiconductor technology and voltage class choices. Both, technology and voltage class will ultimately influence the converter losses and the converter efficiency.

For a given voltage class, higher or lower power can be converted using more or less parallel connected devices. As long as the current density in each device remains the same losses scale linear with power. The same holds true for the losses in the bus bar arrangements.

This section presents an optimization in terms of efficiency for the stack and converter voltage level. The grid voltage level is solely affected by the transformer design and will not be further discussed. Any medium voltage grid voltage level can be achieved by an adequate transformer design. State-the-art semiconductors are silicon based and are generally manufactured with uni-polar or bipolar device structures. Nominal blocking voltages of 100V, 200V, 300V, 600V, 1200V and 1700V are the pre-dominant voltage classes for low voltage power conversion systems. Blocking voltage ratings other than these values exist but are not included in the comparison. Devices up to 500V feature primarily a uni-polar structure (MOSFETs), devices starting at 600V are typically of bipolar structure (IGBTs, IGCTs, GTOs, thyristors). Devices that use one single semiconductor chip to obtain large current ratings are latching devices, such as thyristors, GTO's and IGCTs. Blocking voltages start above 800V. The maximum voltage/current ratings commercially available are summarized in Figure 84 for device structures mentioned above. Short circuit current turn-off capability of the non-latching devices, e.g. MOSFET or IGBTs, is advantageous in case of a converter shoot-through. Latching devices such as IGCTs or GTOs require additional di/dt limiters partially eliminating the on-state loss advantages in comparison to non-latching devices.

The availability of device blocking voltages in discrete steps in combination with the maximum device utilization prescribes a unique number of cells for each semiconductor device class. For a maximum DC bus voltage equivalent to 70% of the device blocking voltages, i.e. an over-voltage margin of $(1.0/0.7) = 1.42$ the optimum number of cell is shown in Figure 83 up to a device blocking voltage of 1700V. The over-voltage margin is needed to prevent a device breakdown during short circuits and is influenced by the stray inductances in the converter set-up, the dc bus and intrinsic device capacitances.

For the optimum voltage level selection, the calculation of the specific losses, i.e. the losses per unit power converted, are analyzed for various voltage levels. The source current is kept constant at rated current density of the fuel cells. Roughly the same device current margin, a ratio between the nominal device current and the rms output current, of $I_C \sim 1.3 I_{rms}$ is installed for all sample device voltage classes. The modulation indices and displacement power factors are equal assuming the transformer winding ratios are adjusted adequately. Figure 84 shows the specific converter losses as a function of the switching frequency and device blocking voltage capabilities, which in turn is set by the number of cells. For switching frequencies between 2 and 4 kHz, i.e. at typical PWM converter switching frequencies, the optimum device blocking voltage is 1200V. At low switching frequencies typical for the multi-pulse concept a higher DC bus voltage is preferable. Power conversion efficiency can be increased by ~0.5% if the stack is extended from 400 cells to 800 cells allowing a transition from 600V to 1200V devices. The effective commutation voltage at rated current is only ~49% of that of the device blocking voltage capability. This is given by the over-voltage margin needed for no-load operation plus the intrinsic cell voltage drop at rated current.

Individual semiconductor devices for IGBTs are manufactured up to a nominal current rating of ~ 50A. Larger current ratings are achieved by paralleling these individual chips in semiconductor modules. Semiconductor losses are unaffected by the current of the fuel cell stack

Figure 83 Number of cells per SOFC stack as a function of device voltage with over-voltage margin of 1.42

Figure 84 Specific semiconductor losses for various converter switching frequencies ($V_{dc} = 0.5 V_{CE}$)

Device Type	Device Voltage Class	Device Name	Device Current Rating at 75/80C	# parallel per switch position
MOSFET	75	VWM 350-0075P	250	6
MOSFET	100	VMM 650-01F	500	3
MOSFET	200	VMM 580-02F	430	3
MOSFET	300	VMM 300-03F	220	6
IGBT	600	BSM150GB60DLC	150	9
IGBT	600	CM200DU12F	200	7
IGBT	1200	FS450R12KE3	450	3
IGBT	1200	CM200DU24F	200	7
IGBT	1700	FS450R17KE3	450	3

Table 32 Device technologies

7.2.3.4.3 Converter Design

The following section describes a single stage power conversion design avoiding the high stack input voltage requirements, eliminating conventional pulse width modulation, providing the minimum number of components in the conduction path, retaining full AC rms voltage controllability, and providing a modular structure for high volume production.

SOFC stacks change their internal impedance during their operational life. In order to avoid circulating currents in differently aged stacks each stack must be connected to an individual converter. A central DC bus configuration cannot be implemented without a second power conversion stage.

A single-stage directly coupled dc-to-ac power conversion derived from the multi-pulse GE Battery Energy Storage Systems (BESS) design is proposed. The new system is based on a separation of the DC bus into multiple DC bus channels such that independent DC sources can be connected, an adjustable DC bus voltage for optimum conversion efficiency and system availability, a unique transformer design that can be used for all conversion channels, and state-of-the-art trench gate IGBTs. At least six channels per multi-pulse system, resulting in a 36-pulse system, are needed to comply with existing utility power quality requirements. A higher number of converter channels forms a higher order multi-pulse system improving the power quality. Several multi-pulse systems can be connected in parallel.

The electric diagram of a six channel, independently sourced multi-channel systems is shown in Figure 85. The primary windings of the polygon transformers are connected to individual converter channels. The secondary windings of the polygon transformers are connected in series. There is no additional AC filter needed.

The control over the line harmonics is achieved by the phase-shifted transformers in combination with an adequately phase-shifted square wave converter voltage featuring dedicated phase delays for each converter channel. Control over the output voltage is achieved by (i) controlling the dc bus voltage or (ii) by introducing a notch in the modulation function reducing the effective volts-seconds. A control over the phase angle of the entire system is achieved by shifting the entire phase reference system on which the individual channels derive the phase information.

The controls described above are sufficient to adjust real and reactive power flow (magnitude and direction) as well as the harmonic content of the voltage at the medium voltage side of the transformer.

The details of the converter and the fuel cell interconnection are shown in Figure 86. The fuel cells may be linked to the DC bus converter via a protective circuitry. Details of the over-current and reverse voltage protection circuits are yet to be determined. The selection of the dc bus voltage is based on efficiency and availability criteria. The transformer ratios of the polygon transformers can be adjusted for any desired transformation ratio. Galvanic isolation is intrinsically provided.

Figure 85 Six-channel, independently sourced multi-pulse converter

Figure 86 Primary conversion stage associated with each SOFC stack

7.2.3.4.4 Plant level layout

The integration of the multi-pulse converters into the 25 MW SOFC hybrid architecture is shown in Figure 87. The system is highly modular, can be installed in phases, and can easily be expanded to higher power levels if grid capacity is available.

At the predicted conversion efficiency, each multi-pulse converter associated with one fuel cell module is designed to provide 1.8 MW of real power (with 6 stacks per module). The maximum apparent converter power capability is 2.1 MVA.

On the turbine side, a gear-less high-speed turbine generator provides the approximately 5 MW. The high-speed generator requires a dedicated DC/AC power converter capable of providing 5 MW power at 300 to 800 Hz fundamental frequency. Partially bi-directional power flow is needed for start-up purposes. Various design options exist for the power converter architecture, the discussion of which are beyond the scope this report. Figure 87 depicts one power conversion system using a similar multi-pulse converter system for the grid connection and a parallel configuration of rectifiers on the generator side. The converter rating is set to 6MVA to support capacitive or inductive power needs.

Each multi-pulse converter system associated with one SOFC module and the high-speed turbine are separately connected to the medium voltage grid. Each generation unit can be disconnected for maintenance. Medium voltage grid voltage and frequency can freely be chosen based on the local needs (10kV to 34.5kV, 50Hz/60Hz). The entire system can only handle base load operation. The ramp up times for both the fuel cell and the gas turbine are too large to follow quick load changes.

Figure 87 Power conversion arrangement of one sub-module using multi-pulse converter

7.2.3.5 25 MW plant stack building block selection

Summarizing the results of stack cost optimization and power electronics subsystem loss studies, the optimal SOFC stack configuration for a 25 MW hybrid plant has 400 cells 45.7 cm (18 inches) in diameter. Assuming cell power density of 0.5 W/cm^2 , this SOFC stack size translates into a 320 kW stack power. Given the 20 MW total SOFC stack subsystem power requirement, a minimum of 62 stacks will be required to construct the 25 MW hybrid plant. Assuming that the SOFC stack subsystem will be divided into modules of 8 stacks each, a minimum of 8 modules will be required for the 25 MW hybrid plant.

Note that the stack cost optimization study showed that the stack cost is relatively flat in the 30-60 cm cell diameter range. This translates into a 135-550 kW SOFC stack building block power range and a 5-19 range for the number of the eight-stack modules.

7.2.4 25 MW plant reliability projections

As a base-loaded, multi-megawatt power generator, the hybrid SOFC power plant has high reliability and availability requirements. The plant specifications call for 98% reliability and 97% availability. This requires careful system design with reliability in mind. The plant is divided into multiple power generating blocks. The multiplicity provides challenges for control system design, but it could be advantageous from the reliability viewpoint. Redundancy could be added in the plant, providing slightly most cost but enormously more benefit in terms of reliability, availability and maintainability.

7.2.4.1 Operability scenarios

Given the multi-stack nature of the power plant, one obvious way to design for reliability is to add redundant stacks in the plant. The stacks are pressurized, and hence located inside pressure vessels. For cost reduction, multiple stacks are placed inside one insulated and pressurized container. A number of stacks inside a single pressure vessel is called a module. In the present designs, the air entering the module and exiting the module is provided with on-off valves, which could be used to isolate a module hermetically from the rest of the plant. Once inside the module, the air is distributed in and out of the stacks by large diameter manifolds. No additional valves are provided in the air path, in order to reduce costs. The air could be used to provide a source for pressurization and heating for the stack. The fuel flow, on the other hand, is individually controlled in this design. Thus, the amount of fuel that flows in and out of each stack is controlled. So when a stack does not produce power, little or no fuel could be flown through it. This flexibility provides enough opportunity to draw different redundancy scenarios with the extra stacks.

Three practical redundant configurations along with maintenance schemes are described below. These three schemes were considered for this analysis. To make comparisons meaningful, each of the following power plant configurations consists of the same number of fuel cell stacks; i.e., 72 fuel cell stacks, distributed equally in a pressure vessel (fuel cell module).

Other redundant designs can be formed with combinations of these 3 schemes.

7.2.4.1.1 Scheme A – Standby Redundancy

The configuration is shown in the following Figure 88.

Figure 88 Standby redundancy (Scheme A)

Eight fuel cell modules (each of which is made up of 8 fuel cell stacks) operate fully and 1 fuel cell module is powered off or in standby in the normal operation mode. An outage of any stack in any active module will lead to switching off one entire working module and starting/switching in the standby module. The failed module will be under restoration immediately.

This scheme is said to be operating in an N+I redundancy mode if there are N active modules and I stand-by modules.

7.2.4.1.2 Scheme B – Parallel Redundancy

This design is shown in the following Figure 89.

Figure 89 Parallel redundancy (Scheme B)

All 9 modules are powered up, but operate in a reduced-load mode, during normal operations. Whenever a fuel cell stack fails, the entire module will be switched off and the remaining 8 modules will be fully loaded to compensate for the deficit from the failed module. The failed stacked is under repair immediately and put back once it's fixed. After that, normal reduced load operation is resumed.

This scheme is said to be operating in an N+I redundancy mode if the required number of modules is N but the number of active modules in the normal operating mode is N+I.

7.2.4.1.3 Scheme C – Enhanced Module

Instead of having a redundant module as shown in the designs above, an extra fuel cell stack is put into each module. System configuration is shown in the following Figure 90.

Figure 90 Enhanced redundancy (Scheme C)

All stacks normally operate in the reduced load mode. The failure of a stack leads to the remaining stacks in that degraded module working at full load, and failed stack will be replaced upon next maintenance interval. After being fixed, it will be put back into service.

This scheme is said to be operating in an N+I redundancy mode if the required number of stacks per module is N but the number of active stacks per module in the normal operating mode is N+I.

7.2.4.1.4 Reliability Comparison

The power plant reliability of 3 schemes above have been modeled precisely and analyzed, using the commercially available software BlockSim™. To compare the three scenarios, only the assembly of fuel cell stacks are considered – the balance of plant or the bottoming cycle is not considered at this stage. The assumptions for the analysis are given as follows.

- A fuel cell stack has a MTTF of
 - 40,000 hours at full load,
 - $MTTF_{stack} \propto load^{-1}$ or 45,000 hours at 8/9 full load
- MTBF is not a function of anything other than the stack power level
- Average repair time is 72 hours
- Scheduled maintenance interval is 6 months
- Stacks within a module operate independently
- Modules are independent of each other. There are no common failure modes across the modules.
- Both “time to fail” and “time to repair” are exponentially distributed

7.2.4.1.5 N+1 redundancy results

The MTBF of 3 scenarios are listed below in Table 33. It should be kept in mind that this is the MTBF of the assembly of the stacks, arranged in modules. To that extent, these numbers are artificial, but they do help distinguish between the scenarios.

	Power Plant MTBF (hours)
Scheme A - Standby Redundancy	4,647
Scheme B - Parallel Redundancy	6,054
Scheme C - Enhanced Module	3,151

Table 33 Mean time between failure (MTBF) for N+1 redundancy case

From the results it is seen that

- Scheme B is the most effective way to introduce redundancy with maintenance schedules among these 3 configurations.
- The absolute MTBF is rather small for all the three scenarios. More degrees of redundancy are needed to achieve the desirable reliabilities.

Since the Scheme B is the most efficient way of redundant arrangement, the rest of report will focus on this scheme only.

7.2.4.1.6 N+2 redundancy results

To further improve the plant reliability, N+2 level redundancy has been investigated. Table 34 shows the MTBF at different redundancy levels, using the previous assumptions. Only Scheme B is reported.

	Power Plant MTBF (hours)
N+0 (no redundancy)	625
N+1 redundancy	6,054
N+2 redundancy	105,927

Table 34 Redundancy trade study

The N+2 level of redundancy shows a significant jump in plant reliability from the N+1 level. With 2 extra modules in place and a repair made immediately after a failure, a power plant maintains a redundant configuration almost all the time. The plant is down only if at least 3 modules fail within a repair-time window, which has a very low probability unless there is a common failure cause. In short, the Scheme B with a N+2 level of redundancy may be able to provide a SOFC power plant with a satisfactory level of reliability.

7.2.4.2 Reliability optimization and cost effects

Having selected Scheme B with N+2 redundancy as the best candidate for a 25 MW class hybrid SOFC power plant, several trade-off studies could be performed.

7.2.4.2.1 Number of modules in plant

A 25 MW plant would need several building blocks, defined as stacks of a certain nominal power output. These stacks must be divided in modules to lower capital costs. Based on reliability considerations, the optimum number of stacks per modules, and hence modules per plant, could be determined. A schematic of this grouping procedure is shown in Figure 91 below.

Figure 91 Building block grouping scenarios

Thus, a 64-stack plant could be divided in 4 modules with 16 stacks each or 16 modules with 4 stacks each, amongst other possibilities.

Results for stack assembly MTBF for different sized modules are given in Figure 92 below. These results correspond to Scheme B, using the same data as the previous section. Both N+1 and N+2 redundancy situations are shown.

It appears that there is an arrangement that maximizes the MTBF for the stack assembly. This is intuitively clear, as seen in the figure above. In Scheme B, failure of one stack requires the shut down of the whole module containing the failed stack, while raising the load in the operational stacks. In the left hand side scenario, this means a much-increased load for the remaining stacks, since a large number of stacks have to be shut down. On the right hand side scenario, however, there are too many modules, so the probability of all of them working at the same time is relatively smaller. Thus, there is an optimal arrangement.

Figure 92 MTBF analysis for N+1 redundancy cases

7.2.4.2.2 Cost penalty of redundancy

Increasing the number of redundant modules improves the system reliability but at the same time increases the cost as shown in Table 35, below. The choice of 2 redundant modules in the base case meets the cost requirements and at the same time provided satisfactory system reliability.

Number of redundant modules	Cost/kW	Total cost (M\$)
0	292.76	7.319
1	305.08	7.627
2	317.40	7.935
3	329.72	8.243

Table 35 Cost of redundancy

7.2.4.3 Cell reliability effects on optimal cell size selection

The plant MTBF strongly depends on the stack MTBF. However, stack MTBF could depend on cell size, since the number of defects in ceramic components is expected to

depend on the component surface area. Most, but not all, of the defects are caught during cell QC and are accounted for in the stacking yield. Thus the failure mechanisms are cell area dependent, causing the MTBF to decrease with the cell area. Better understanding of failure mechanisms may lead to stack designs that could diminish the dependence of cell MTBF on cell size.

No experimental data is currently available for proving how the reliability assumptions matter for the cell size selection. In order to investigate the impact of the cell size on the stack MTBF we assumed the following 4 scenarios:

- MTBF independent of the area
- MTBF increases 5% with the cell area
- MTBF decreases 15% with the cell area
- MTBF decreases 26% with the cell area

Figure 93 shows the assumed dependence of the stack MTBF as on the cell size for the above scenarios.

Figure 93 Stack mean time between failures as a function of cell diameter

Using the above assumptions, stack MTBF could be rolled up into plant MTBF. The impact of cell size on overall plant reliability for Scheme A (no redundancy) is shown in Figure 94.

Figure 94 Plant mean time between failure as function of cell area results (assuming a 25 MW plant with no redundancy using scheme A)

As cell size increases, the number of stacks needed in the plant to produce 20 MW decreases. This is the main reason why the plant MTBF increases with cell size. In fact, if the cell reliability is independent of the cell size then the cell size should be chosen as large as the system cost goal allows. However, the cell reliability may be a strong function of the cell size. Therefore, beyond a certain size, the stack MTBF could be low enough to have a negative impact on the plant MTBF, as shown in Figure 94.

Recall that the optimal cell size was determined solely by minimizing the SOFC stack cost. Other constraints however may limit the cell size. In particular, Figure 94 shows that there can be reliability constraints on the cell size. For example, if the cell MTBF falls by more than 26% from cell diameter of 30.5 cm (12 inches) to cell diameter of 60.96 cm (24 inches) then cell diameter of 45.7 cm (18 inches) is sub-optimal because it leads to sub-optimal plant MTBF in Scheme A. Since there is very limited data on cell reliability in the size range of our interest, it is impossible to factor the cell reliability constraints into the cell size selection process. However, it must be recognized that the size reliability effects must be taken into account when the cell reliability data is available.

A similar study was done on the down selected arrangement of Scheme B with N+1 redundancy, with MTTR=72 hrs, Figure 95.

Figure 95 Plant mean time between failure verses cell area results (assuming 25 MW plant with N+1 redundancy using scheme B)

It could be seen that even with a 60% decrease in stack MTBF, the plant MTBF does not decrease with cell size. This is a further testimony that Scheme B is more robust than Scheme A.

7.2.5 25 MW plant component selection

Table 36 presents a list of major components of the 25 MW plant design. The performance, cost and reliability studies of the 25 MW hybrid plant have yielded the optimal cell, stack and module sizes. The gas turbine and balance of plant components were sized using the design point analysis. Some additional considerations were taken into account as follows.

- One gas turbine per system is preferable because the gas turbine component efficiencies and costs both improve with increasing size.
- Pre-reformers have economies of scale that result in one pre-reformer per system being the preferable configuration. However, reliability Scheme B selected in the study calls for an isolation of the module with a failed stack and subsequent maintenance. The procedure requires that fuel flow to the failed stack be temporarily cut off. The most efficient way to isolate individual stacks is to use a cold-temperature valve at the inlet of the pre-reformer that feeds the failed stack, which requires one pre-reformer per stack. Even though there is a

cost penalty associated with building many small pre-reformers rather than a fewer number larger pre-reformers, the savings from placing cold-temperature valves on the natural gas side instead of high-temperature valves on the reformat side make up for it.

Part	Quantity	Comments
Gas Turbine & Auxiliaries	1	
Recuperator	1	
Pre-Reformer	80	One per stack
Blower	1	
Ejector	80	Part of pre-reformer
SOFC Stack	80	
Pressure Vessel	10	One per module
Burner	1	
DC-AC Inverter	80	One per stack
Small Transformer	10	One per module
AC-AC Inverter	1	
Cathode Isolation Valves	10	
Anode Isolation Valves	10	
Start up Valves	2	
Fuel Preheater	1	
Water Pump	1	
Water Tank	1	
Steam Generator	1	
Fuel Compressor	1	
Desulfurizer	1	
Main Fuel Shut-off Valve	1	
Switch Gear	10	One per module
Control Box	1	
Start up Auxiliaries		
Instrumentation		
Cold Piping		
Hot Piping		

Table 36 List of major components for 25 MW SOFC-gas turbine hybrid plant

7.2.6 25 MW plant manufacturing cost projections

Our analysis with baseline yield assumptions and a 45.7 cm (18") diameter cell results in a \$317/kW system manufactured cost, Figure 96, which is significantly below the \$400/kW goal. SOFC stacks are the main driver of the system cost reduction, as they comprise 36% of the system cost. Recall that a 0.5 W/cm^2 was assumed in cost analyses. In fact, the fuel cell power density can be as low as 0.305 W/cm^2 and still enable the system design to achieve the \$400/kW cost target. Figure 97 shows the system cost breakdown for the 0.305 W/cm^2 case that results in a \$400/kW system cost.

Figure 96 Plant cost breakdown assuming 25 MW hybrid plant with baseline cost assumptions and cell current density of 0.5 W/cm^2 , system specific cost = \$317/kW

Figure 97 Plant cost breakdown assuming 25 MW hybrid plant with baseline cost assumptions and cell current density of 0.305 W/cm²

7.2.7 Discussion of system size effects

The discussion has so far concentrated on the 25 MW plant design. The design methodology is likely applicable to a range of system sizes. Note that the product specification was adopted from the central power generation market application and is valid for a range of system sizes up to hundreds of megawatt. The next section will address the applicability of the SOFC building block design to the system power rating range 5-250 MW. System performance, size and reliability effects in this range will also be analyzed.

7.2.7.1 SOFC stack building block selection

The analysis in section 4.2.3 identified the optimal stack size for the 25 MW hybrid plant. The optimization process found the cell size that minimized the stack cost and the number of cells per stack that minimized the power conditioning losses while addressing stack reliability issues. These two optimization problems are independent of each other and, more importantly, independent of the system size. The stack cost will be minimized independent of whether the system power rating is 25 MW or 250 MW. Similarly, the power electronics loss per stack is minimized independent of the system size. However, a minimized stack cost and power electronics loss-per-stack do not necessarily mean that the system cost and system power electronics are minimized with

the same stack building block. The differences between the smaller and larger systems need to be examined.

For example, let us consider a 250 MW hybrid plant. If the same stack and module building block sizes as those in the 25 MW hybrid plant are assumed, then the 250 MW hybrid plant would have ten times the amount of hot piping. As Figure 96 and Figure 97 show, hot piping comprises only about 1% of total system cost, which is dominated by the SOFC stack subsystem cost. Therefore, the increase in hot piping length with system size increase should not significantly affect the optimal cell size that minimizes the SOFC stack subsystem cost and hence, the system cost. The stack block of 400 cells 45.7 cm (18 inch) in diameter is still the optimal stack building block for the 250 MW plant as well.

Similarly, a 5 MW hybrid plant can also use the same stack design as the building block. In fact, the smallest system size for which the module block of eight 320 kW stacks can be used as a building block is about 1.3 MW. The 1.3 MW plant however will only have one module and therefore, the operability Scheme B cannot be applied to this system. Reliability and operability considerations will determine the minimum system size for which the stack and module blocks can be used. Section 4.2.7.3 will examine system size effects with respect to reliability.

7.2.7.2 Performance size effects

Hybrid plant performance at the design point depends on component efficiencies, pressure drops, and heat losses. The component efficiencies are generally functions of component size. The pressure heat losses usually decrease with system size because the system component surface area increases more slowly than the system component volume.

The hybrid plant SOFC efficiency does not change with the system size because the stack and module building blocks do not vary with the system size. The gas turbine component efficiencies on the other hand do depend on the system size. In general, the compressor and turbine efficiencies increase with size. Similarly, natural gas compressor efficiencies also increase with component size. Using GE's projected components efficiencies, it can be shown that the efficiency of a 250 MW hybrid plant is 67.7%. This efficiency is about 1.7 percentage point higher than the 25 MW plant efficiency due to higher turbomachinery component efficiencies and a lower heat loss.

Similarly, the system efficiency of a 5 MW plant will be lower than that of the 25 MW plant. A system performance analysis shows that the 5 MW plant efficiency is about 65.1%, about one percentage point lower than that of the 25 MW plant due to lower component efficiencies and a higher heat loss.

7.2.7.3 Reliability projections

The proposed layout of 6 or 8 stacks per module and 12 or 8 modules in the plant is specific to the 20 MW plant size, which needs about 64 standard sized (320 kW) stacks, plus 16 redundant stacks. For a smaller plant, the building blocks or the stacks will stay the same in size. This is because the considerations that led to the selection of this specific stack size (power electronics and cost considerations) are not linked to the

plant size. However, the layout for a smaller plant will be different. Given that N+2 redundancy is needed to meet the availability criteria, smaller plants would tend to have fewer stacks per module.

Let us consider a 5 MW class hybrid power plant, where about 3.5 MW of power is generated by the SOFC stacks. This plant would require about 12 standard sized stacks in full load operations. If there are 4 stacks per module, the plant would have a total of 20 stacks, including redundancy. On the other hand, for 2 stacks per module, the plant would have only 16 stacks, including redundancy. Although the cost of the pressure vessels and pipes would be larger in the later arrangement, the savings in the stacks, which tend to be the biggest cost item in the plant, would largely offset this additional cost. For very small systems (1 MW or less), if the availability criterion is relaxed, it is conceivable that the whole set of stacks would be contained in a single pressure vessel to keep cost to a minimum.

Similar reasoning could be applied for larger systems. The building blocks stay the same in size. The stack count in the module would tend to go up for optimal cost. However, a limit might be reached in terms of power conditioning capability. In the power electronics layout proposed in this study, there is a single transformer per module. For a large number of stacks per module (10 plus), this transformer would be required to handle very large currents, which would lead to high costs. Correspondingly, the cost of the insulated pressure vessel, which scales with the vessel volume, would keep going up. Thus, beyond about 10-12 stacks per module, the cost benefits from stack arrangement would be offset by rising BOP costs.

7.2.7.4 Cost projections

The SOFC stack subsystem of larger hybrid systems consists of the same stack and module building blocks as the 25 MW hybrid system's SOFC stack subsystem. A hybrid system with the power rating larger than 25 MW is built by adding more SOFC modules. Since the cost of SOFC stack modules dominates the SOFC stack subsystem cost, the SOFC stack subsystem shows no economies of scale, i.e. its cost per kW stays virtually constant with varying system power. The same cannot be said however about the gas turbine and balance-of-plant components, which generally show strong economies of scale. For example, the system cost analysis of a 250 MW shows that the system specific cost declines to \$260/kW (Figure 98).

Figure 98 System cost breakdown for a 250 MW hybrid plant, plant manufactured cost = \$260/kW

Similarly, the specific cost of a hybrid system with the power rating of lower than 25 MW is higher than that of the 25 MW hybrid system. For example, Figure 99 shows that the system specific cost of a 5 MW hybrid system is \$512/kW.

Figure 99 System cost breakdown for a 5 MW hybrid plant, plant manufactured cost = \$512/kW

7.2.8 25 MW hybrid system operation on coal gas

The 25 MW hybrid system design permits operation on coal gas with some modifications to the plant. The coal gas may be supplied from an outside coal gasification system. A typical coal gas composition is shown in Table 37. Since the coal gas contains methane in concentrations acceptable to SOFC, there is no need for additional fuel reforming. Therefore, pre-reformers and the associated piping and valves can be removed from the plant and replaced by coal gas piping and interfaces. The rest of the hybrid system design, including the SOFC stack subsystem and the gas turbine, requires no modification. The amount of coal gas flow to the hybrid plant is limited by the cooling capacity of the air supplied by the gas turbine compressors. In the analysis presented in this section, the coal gas flow to the system was adjusted to meet but not exceed the SOFC stack subsystem air temperature rise and temperature out constraints.

The result of a performance analysis of such a plant is shown in Table 38. The analysis shows that the power produced in the SOFC stack subsystem is lower than that of the natural-gas-fueled 25 MW hybrid plant, 11.2 MW vs. 20 MW. The drop-off in the SOFC power can be attributed to the following two factors. (1) The SOFC single-cell voltage drops below the design assumption of 0.7 due to lower concentration of usable fuel in coal gas. Our calculations showed that the single-cell voltage for the coal gas

composition shown in Table 37 declines to 0.6V. The lower single-cell voltage results in a lower SOFC power. (2) The coal gas methane concentration at the SOFC stack inlet is only 3.1% compared to 16.1% for the natural-gas-fueled system. The internal reforming reaction of methane in the SOFC stacks serves as an additional cooling mechanism and reduces the cooling requirement imposed on the cathode air. For a fixed gas turbine size and therefore cathode airflow, the smaller amounts of methane in the fuel inlet translates into a lower amount of the total fuel flow that the SOFC stack subsystem can convert to electricity without violating the SOFC stack subsystem air temperature rise and temperature out constraints. Hence, the system operating on coal gas has a lower capacity to convert useful LHV in the fuel stream into electricity than the natural-gas-fueled system due to a lower stack-inlet methane composition, resulting in a lower SOFC stack power.

Table 38 shows that the gas turbine power in the coal-gas-fueled hybrid is 9.5 MW compared to 7.8 MW in the natural-gas-fueled hybrid plant. The reason for this increase is a higher turbine mass flow in the coal-gas-fueled hybrid plant compared to the natural-gas-fueled hybrid plant due to a higher concentration of CO₂ and water in the fuel composition, resulting in a higher turbine mass flow.

The coal-gas-fueled plant efficiency is about 69%. Note that this efficiency does not include the losses for converting the fuel (coal) to the SOFC stack subsystem fuel feed. Coal gasifiers typically have efficiencies in the 70-80% range, where the gasifier efficiency is defined as (LHV of gasifier output)/(HHV of coal feed). Factoring in the gasifier efficiency, the total coal-fueled plant efficiency of the plant is in the 48-55% range.

Stream Component	Molar Fraction
H ₂	0.0348
H ₂ O	0.293
N ₂	0.0224
CO ₂	0.445
CH ₄	0.0314
CO	0.173

Table 37 Coal gas composition (BGL gasifier output on Pittsburgh 8 coal)

Component	POWER PRODUCED (CONSUMED), KW
SOFC Stack Subsystem	11186
Gas Turbine	9353
Blower	(1549)
Fuel Compressor	N/A
Net Power	17977
System Efficiency	69%

Table 38 Performance analysis results of a 25 MW hybrid plant fueled by coal gas

8 TECHNOLOGY GAPS AND DEVELOPMENT STRATEGY

The hybrid plant design developed in this study requires cell size scale up, stack pressurization and operating plant components at high temperatures. It is not feasible, at present, to realize the plant concept presented in this report because significant technological gaps exist in certain areas, primarily in SOFC technology. Moreover, in other areas, existing products must be extensively re-engineered to achieve the desired objectives. This section lists the gaps and development requirements that must be closed to enable feasibility of the plant concept.

8.1 Technology gaps

8.1.1 SOFC Seals

One of the biggest risks embedded in the system design presented in this report is the availability of the SOFC seals. The seals are essential to enable the cathode and anode recycling that significantly increases the system efficiency and reduces the plant cost. The seals must operate at temperatures above 1400°F. Analyses show that seal leakage must be limited to less than 10% of fuel flow rate. No feasible seal solution exists today, a significant effort is required to develop seal materials and cell and stack designs that would satisfy the sealing requirement.

8.1.2 SOFC Power Density

With the current set of materials widely in use, YSZ for electrolyte, LSM-YSZ for cathode and Ni-YSZ for anode, the entitlement for reaching the power density target of 300 mW/cm² or higher at 0.7 V exists without a doubt. However, improved understanding of materials and microstructures is needed to meet or exceed the performance goals. For instance, optimizing the composition and microstructure near the interfaces for improved electrochemical activity can help reduce activation polarization at the electrode/electrolyte interfaces. Though the pressure helps in

improving the diffusion polarization, work needs to be done to engineer the microstructure for better gas access to the electrodes. Another area of significant importance that needs better understanding and improvement is the ohmic drop across various interfaces, especially the interconnect/electrode interfaces. Concentrated efforts need to be made to reduce the ohmic drop across these interfaces for improved performance as well as reliability. Possible studies include new interconnect materials with low degradation rates and improving the characteristics of current collection layers at the interconnect/electrode interfaces.

8.1.3 Cell Size Scale up Feasibility

Present study indicates that a cell size of 45.72 cm (18 inches) diameter is needed for optimized stack cost. Reported state-of-the art in cell size for planar stacks tested worldwide is as follows: Mitsubishi Heavy Industries, Japan (20 cm x 20 cm); Fuel Cell Energy, USA (15 cm or 5.9 inch diameter); Julich, Switzerland (20 cm x 20 cm); Ceramic Fuel Cells Limited, Australia (13 cm diameter); General Electric, USA (20.32 cm or 8 inch diameter); and Indec, Switzerland (15.6 cm or 6.13 inch diameter). The cell size needs to be scaled up more than 100% from the best one available today. As cell size scales up, the influence of cell area on mechanical properties and electrochemical properties need to be understood. General Electric has recently demonstrated cell size scale up to 30.48 cm (12 inches) diameter on anode-supported cells using the tape calendaring method but more effort is needed to understand the yield, strength and performance of these large area cells. Interconnect supported cells using deposition techniques have also been proposed but much gap remains to be filled regarding the cost and performance. Efforts need to be directed in electrochemical and stress and thermal modeling to define cell configurations capable of high electrochemical performance and strength. For electrochemical performance remaining the same, yield data needs to be collected as a function of cell size to validate/refute the cost model assumptions used in this study. With the help of modeling, process development efforts would be needed in the area of cell manufacturing to demonstrate the feasibility of manufacturing large size cells. To reduce labor costs, this study further assumes continuous manufacturing as opposed to batch manufacturing that is currently being used in the industry. Hence, routes to continuous manufacturing of SOFC cells need to be explored to achieve the cost targets.

8.1.4 Stacking Improvements

For this study, a stack design currently in use for the DOE SECA program was chosen for concept down selection. However, for a stack building block of 320 kW using large cells, many stack design issues, such as stresses, pressure drop, leakage and thermal gradients, need to be understood as a function of cell size. For instance, the interconnect design and/or cell-to-cell manifolds may need to be optimized to take care of differences in temperature gradients and gas flow rates at larger cell sizes. Design efforts are also needed to understand the stresses generated during assembly and operation of large cells. This information can then be used in cell manufacturing to design and fabricate cells with strength higher than that governed by the design stress.

Appropriate quality control methods will need to be developed to screen out defective cells and/or other defective parts not meeting the specifications. As cell size increases, the area over which seals are needed also increases. Seal materials that can withstand pressurized operation over large areas and for stack sizes in the range of few hundred kW are not proven at present and will merit further investigation.

8.1.5 Reliability

This study assumes different scenarios of system reliability as a function of cell size. However, to reduce variability of system study results, statistical data on the stack/system reliability as a function of cell size needs to be generated. For example, stack MTBF was assumed to be inversely proportional to stack power. Although most experts assume a dependence of MTBF on the power output, there is a strong need for experimental data to find the right relationship between the two. The same statement could be made about the effect of stacking (number of cells per stack) on the stack MTBF. The failure mechanisms that cause de-rating and outage in large stacks are also poorly understood, and must be investigated carefully.

8.1.6 Degradation

Degradation in fuel cell systems is universally accepted as a major issue. Long-term stack degradation is poorly understood, both in terms of the fundamental mechanisms as well as its repercussions for the system. Degradation affects all aspects of system design, performance, reliability, operability and cost. This study focused on the system scalability, thus no assumptions regarding degradation was made. The effects of stack degradation on a hybrid plant should be carefully studied.

8.2 Engineering development needs

The following engineering development needs have been identified in the course of the study. Unlike the technology barriers discussed in the previous section, these areas bear a much lower development risk and lower investment needs.

- Most of the plant layout surrounding the SOFC stack modules is novel and untested.
- No gas turbine presently exists with the particular combination of pressure ratio, flow rate, and low firing temperature called for by this design. Additionally, this design requires 100% of the compressor air to be piped to the fuel cell modules before returning to the combustor and turbine. Engineering development is needed to develop compressor and turbine plenums with low pressure losses.
- Large number of stacks, along with redundant stacks, operating at full load and part load need robust design of control systems. This is especially important given the high availability requirements of a hybrid SOFC plant. Integration of gas turbine and fuel cells, with very different transients, would also be a challenge in control system design.

- The power conditioning topology proposed in this study is untested. Given the critical role power conditioning plays in determining system efficiency, further modeling and simulation is needed. Effects of ripple, fault tolerance and similar issues need to be dealt with in detail. Similarly, the integration of a hybrid SOFC plant with the grid could pose design challenges that need to be studied.

Plant start-up and shut down was not covered in this study at all. This is an important subject, which has cost and reliability impacts on the plant. More focus is needed on this topic.

Conclusions

The study developed a system design that satisfies the product specification and a SOFC stack development strategy. The system design of a 25 MW plant showed that system efficiencies of over 65%, system cost of under \$400/kW, and system reliability of over 98% are feasible.

The plant SOFC subsystem is constructed from a 400-cell stack building block. A wide range of the cell diameter is acceptable: the system cost does not vary significantly in the cell diameter range 30-60 cm (12-24 inches), with the corresponding stack building block power in the 140-570 kW range. Reliability considerations favor large cell diameters and a smaller cell count, however the cell reliability may decline rapidly with the cell diameter. Cell reliability dependence on cell size can put additional constraints on the cell size and stack building block selection and must be studied for the 30-60 cm (12-24 inches) cell diameter range. Due to the lack of available reliability data on large solid oxide fuel cells operating in a stack, the cell size constraints were not quantified, and instead the cell size of 45.7 cm (18 inches) in the middle of the cost-minimizing cell range was chosen, which translates into a stack building block power of 320 kW.

The 25 MW hybrid plant design developed in the study can be scaled to both higher and lower power levels using the same stack building block for wide range of system sizes, 5-250 MW or larger.

High-temperature stack exhaust seals are the highest system and stack development risk for the recommended system design. The stack and system risk mitigation strategy should first address this risk. Cell scalability and yield and reliability improvement efforts at large cell sizes must be addressed parallel to the high-temperature seal development to prove the feasibility of the hybrid system design.

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Appendix B

Coal Integrated Gasification Fuel Cell System Study

Final Report

April 2003 to January 2004

Chellappa Balan

Debashis Dey

Sukru-Alper Eker

Max Peter

Pavel Sokolov

Greg Wotzak

January 2004

Performed under DOE/NETL Cooperative Agreement
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Hybrid Power Generations Systems, LLC
19310 Pacific Gateway Drive
Torrance, CA 90502

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ABSTRACT

This study analyzes the performance and economics of power generation systems based on Solid Oxide Fuel Cell (SOFC) technology and fueled by gasified coal. System concepts that integrate a coal gasifier with a SOFC, a gas turbine, and a steam turbine were developed and analyzed for plant sizes in excess of 200 MW. Two alternative integration configurations were selected with projected system efficiency of over 53% on a HHV basis, or about 10 percentage points higher than that of the state-of-the-art Integrated Gasification Combined Cycle (IGCC) systems. The initial cost of both selected configurations was found to be comparable with the IGCC system costs at approximately \$1700/kW. An absorption-based CO₂ isolation scheme was developed, and its penalty on the system performance and cost was estimated to be less approximately 2.7% and \$370/kW. Technology gaps and required engineering development efforts were identified and evaluated.

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12 1 INTRODUCTION

Coal is the most abundant of all fossil fuels, and power generation in the United States relies heavily on coal. As seen in Figure 100, 51% of the total power generated in 2001 came from coal. In the next 20 years, coal is expected to remain the primary fuel source for power generation, although its share of total generation declines as natural gas increases its share (U.S. DOE/EIA, “Annual Energy Outlook 2000”).

Figure 100. 2001 U.S. Electric Generation by Fuel Type

Source: U.S. DOE/EIA “Annual Energy Review 2001”

As concern about the environment generates interest in clean energy, fuel cell power plants can respond to the challenge. Fuel cells convert hydrocarbon fuels to electricity at efficiencies exceeding conventional heat engine technologies, while generating lower emissions. Emissions of SO_x and NO_x are expected to be well below current and anticipated future standards. Nitrogen oxides products of combustion are expected to be extremely low in this power plant because power is produced electrochemically rather than by combustion. Fuel cell power plants also produce less carbon dioxide. Fuel cells in combination with coal gasification are an efficient and environmentally

acceptable means of using the abundant coal reserves both in the United States and around the world (Seinfeld et al.).

1.1 Background

A fundamental part of gasification-based systems is the coal gasifier. A gasifier converts hydrocarbon feedstock into gaseous components by applying heat under pressure in the presence of steam, which is used to produce electricity as depicted in Figure 2.

Figure 101. Gasification-based System Concept

Source: U.S. DOE Office of Fossil Energy

The amount of air or oxygen available inside the gasifier is carefully controlled so that only a relatively small portion of the fuel burns completely. This "partial oxidation" process provides the energy for the chemical reactions to take place. Rather than burning, most of the carbon-containing feedstock is chemically broken apart by the gasifier's heat and pressure. Chemical reactions taking place in the gasifier produce so-called syngas, which consists primarily of hydrogen, carbon monoxide, and other gaseous constituents, the proportions of which vary depending on the conditions in the gasifier and the type of feedstock. The gasification reactions are typically endothermic, with the required heat supplied by the combustion process of the remaining carbon-containing feedstock.

Sulfur impurities in the feedstock form hydrogen sulfide, from which sulfur is easily extracted, typically as elemental sulfur or sulfuric acid, both valuable by-products. Nitrogen oxides and other potential pollutants, are not formed in the oxygen-deficient

(reducing) environment of the gasifier. Instead, ammonia is created by nitrogen-hydrogen reactions. The ammonia can easily be stripped out of the gas stream.

In Integrated Gasification Combined Cycle (IGCC) systems, the syngas is cleaned of its hydrogen sulfide, ammonia, and particulate matter and burned as fuel in a combustion turbine. The combustion turbine drives an electric generator. Hot air from the combustion turbine is channeled back to the air separation unit (ASU), while exhaust heat from the combustion turbine is recovered and used to create steam in a Heat Recovery Steam Generator (HRSG) for a steam turbine-generator.

The use of these two types of turbines—a combustion turbine and a steam turbine—in combination, known as a combined cycle, is one reason why gasification-based power systems can achieve relatively high power generation efficiencies. Higher efficiencies or lower heat rates mean that less fuel is used to generate the rated power, resulting in better economics (which can mean lower costs to ratepayers) and the formation of less greenhouse gas per unit of produced power. A 60%-efficient gasification power plant could cut the formation of carbon dioxide by 40% compared to a typical coal combustion plant.

All or part of the clean syngas can also be used in other ways:

- As chemical “building blocks” to produce a broad range of liquid or gaseous fuels and chemicals (using processes well established in today's chemical industry);
- As a source of hydrogen that can be separated from the gas stream and used as a fuel or as a feedstock for refineries (which use the hydrogen to upgrade petroleum products);
- As a fuel producer for highly efficient fuel cells that operate on the hydrogen generated in a gasifier or, in the future, fuel cell-turbine hybrid systems of the kind that is the subject of this program.

Another advantage of gasification-based energy systems is that when oxygen is used in the gasifier (rather than air), the carbon dioxide produced by the process is in a concentrated gas stream, making it much easier and less expensive to separate and capture. Once the carbon dioxide is captured, it can be sequestered—that is, prevented from escaping to the atmosphere and potentially contributing to the “greenhouse effect.”

1.2 Program Objectives

Integrated Gasification Fuel Cell (IGFC) systems using top-level parametric assessment combined with technical judgment on what was achievable with technology advancement. Various component conceptual designs and cycle configurations were addressed in selecting the systems for analysis. Two down-selected configurations were assessed for impact on system complexity, performance, and estimated costs. The best configuration were selected and, for the selected system:

- The impact of carbon dioxide (CO₂) segregation (but not including eventual sequestration) was assessed.
- The impact of fuel cell fuel gas recycle was assessed.
- A rough order of magnitude (ROM) capital cost assessment was made for the identified configurations.

1.3 Prior and related work

Past studies indicated that using conventional gasification and clean-up technologies at a 200-MW scale, can achieve 43.6 % HHV efficiency with IGCC using British Gasification Lurgi (BGL) gasification and low temperature clean-up (Farooque et al., 1990; Sander et al., 1992; and Sandler and Meyers, 1992). Later studies indicated that higher efficiencies can be achieved with higher methane producing gasifiers and by using hot gas clean-up (EPRI 1990; Meyers 1990; and Seinfeld and Willson 1993). Most recently, studies of hybrid fuel cell/turbine systems have shown that LHV efficiencies of 70% can be achieved on natural gas (Ghezel-Ayagh, Sanderson, and Leo, 1999). In another program by Fuel Cell Energy Inc. (FCE), the goal is to build and test a fuel cell power plant operating on coal-derived gas as a part of the Clean Coal IGCC project.

All of these projects and studies have focused on carbonate fuel cells operating on coal-derived gas. The goal of the project reported here is to identify efficient plant system configurations of coal gasification combined with a planar solid oxide fuel cell (SOFC) and a bottoming cycle.

13 2 EXECUTIVE SUMMARY

The Solid Oxide Fuel Cell (SOFC) is regarded as one of the most promising power generation technologies for the future. With unmatched thermodynamic efficiencies and the ability to be combined with turbines to form hybrid power plants, SOFC power plants could be of various sizes, ranging in output from a few kilowatts to several hundred megawatts. Traditionally, natural gas is used as the fuel for SOFCs. For large power plants, the use of coal as the fuel becomes a possibility, since the scale permit gasifiers to be integrated in a SOFC-based hybrid power plant. This project studies the integration opportunities between a gasifier and a SOFC-based power plant.

Initial analysis was carried out on a reference Integrated Gasification Combined Cycle (IGCC) plant. This plant includes a BGL oxygen blown gasifier that operates on Pittsburgh No.8 coal feed, two gas turbines, and a steam turbine. This reference plant is estimated to operate with an overall efficiency of 40.8% coal on a HHV basis. In subsequent analyses, one of the two gas turbines in the reference power plant was replaced by SOFC modules. This configuration, referred to as the pre-baseline system, is expected to have a system efficiency of 43%. The primary reasons for the limited performance improvements were identified as the use of status technology fuel cell modules along with a significant amount of fuel being consumed in the combustion gas turbine rather than the more efficient SOFC.

After the initial analysis, a number of system concepts with efficient fuel cell stack thermal management and pressurized SOFC, were proposed and evaluated. Two concepts having the most promising efficiency, cost, and reliability entitlement were downselected from the many configurations evaluated.

After the downselection process, a detailed analysis was carried out on both systems to optimize the overall plant performance. The baseline system was found to be 53.4% efficient at a stack operating pressure of 10 bars, while the alternate configuration had a system efficiency of 53.8%. That is about 10 percentage points more than the efficiency of a state-of-the-art technology IGCC plant.

Two different concepts for CO₂ isolation were evaluated, one based on a Selexol system and the other based on combustion of the spent fuel using pure oxygen from an air separation unit. The Selexol-based CO₂ separation resulted in a performance penalty of about 2.6%, while the oxygen-based combustion resulted in a performance penalty of about 2.2%, for both baseline and alternate systems.

The effect of adding methane to the syngas was found to improve plant performance significantly. This result implies that higher plant efficiencies could be realized if the gasification process parameters are optimized to produce syngas with significant methane content.

This baseline IGFC plant is expected to be cost competitive on a per kW basis with today's IGCC plants, provided the capital cost of the SOFC portion of the plant is consistent with the DOE SECA cost targets. The initial costs of the baseline and

alternate concepts were estimated to be \$1654/kW and \$1700/kW, respectively, compared to \$1588/kW for the state-of-the-art IGCC plant. Even with CO₂ isolation, these plant concepts were found to be cost competitive on an initial capital cost basis.

To realize this plant concept, several technology goals must be met, primarily in the SOFC stack. In addition, the turbomachinery and balance of plant (BOP) components will require significant re-engineering.

14 3 EXPERIMENTAL

No experimental work has been performed as part of this study.

15 4 RESULTS AND DISCUSSION

4.1 Pre-Baseline Model and Analysis

4.1.1 Approach

The conceptual analysis of the pre-baseline configuration involved the integration of SOFC modules with commercially feasible IGCC elements and syngas treatment. A structured approach was used to analyze each of the necessary subsystems in order to yield a set of optimum elements, which would result in an efficient IGFC system.

The overall system requirements were identified and flowed down to the subsystem level by utilizing a Six Sigma tool, Quality Function Deployment (QFD) analysis.

These candidate subsystems were evaluated against individual sets of subsystem requirements, and the highest-rated subsystem options were integrated into an overall pre-baseline IGFC Design.

4.1.2 System Description

The IGFC system is an evolved system, which combines the current advantages of the IGCC system for the conversion of coal energy into electric power with the highly efficient SOFC technology. For the pre-baseline case, a conventional (available technology) IGCC system with a gasifier/gas cooling/cleanup system, with two GE 6FA+e gas turbines and a bottoming cycle was modified so that one of the gas turbines was replaced by a solid-oxide fuel cell system. This IGFC System included the following subsystems:

- Gasifier subsystem
- Gas cooling and cleanup subsystem
- Air separation subsystem
- Gas turbine subsystem
- Bottoming cycle subsystem
- Fuel cell subsystem
- Fuel cell stack subsystem

The candidates for each of these subsystems were analyzed by means of global and subsystem requirements in order to yield a credible pre-baseline IGFC system.

4.1.3 Major System Development Drivers

Development and integration of the pre-baseline IGFC case were driven by the following concepts:

- Use of near-term available commercial technology whenever possible.
- Use of projected large (MW Size) planar SOFC technology.
- Use of QFD techniques to provide criteria for down-selection of appropriate subsystems.
- System optimization aiming for achievement of 50% short-term and 60% long-term overall system efficiency.

4.1.4 Quality Function Deployment for Subsystems

In order to select a suitable subsystem, an analysis tool, the Quality Function Deployment (QFD) analysis, was used with the following overall DOE expectations and requirements:

DOE Expectations and Requirements

- Low Cost of Electricity (COE)
- Plant capacity (250 MW or greater)
- High efficiency (60% HHV ultimate system efficiency)
- Applicable to use of fuel cells
- “Near Zero” emissions (of traditional pollutants)
- CO₂ reduction or capture
- 2010-2030 Technology feasibility
- Co-product capability (transportation fuels, H₂, etc.)
- Coal flexibility (plus other optional fuels)
- Minimum water usage
- Minimum hazardous waste

These expectations were cross-correlated in a matrix with the following IGFC System Requirements.

IGFC System Requirements

- Plant cost
- System power output
- System efficiency

- System availability
- NO_x Emissions
- CO Emissions
- CO₂ Emissions
- SO_x Emissions
- Maintainability
- Water usage
- Subsystem R&D required
- Technology choice

This yielded the relative ranking of the key system parameters to be used in determining the suitability of any Subsystem for the final IGFC System, as shown in Figure 102. These relative rankings were used as weighting factors for individual Subsystem Requirements. The available commercial offerings for each subsystem were evaluated using this weighted ranking, as shown in Appendix A.

Figure 102. Relative Ranking of Overall System Parameters

4.1.5 Analysis of Pre-Baseline System

The preceding subsystem optimizations must be considered within the framework of existing studies for production of syngas for IGCC systems. A previous IGCC internal study provides the best example for use as a syngas source for the SOFC Fuel Cell Stack. The original BGL IGCC System was modified for integration with the SOFC stack

Syngas from the BGL system is sent to the fuel cell subsystem, which includes a water-gas shift of the syngas to a fuel more suitable for the SOFC fuel cell stack. Fuel from the fuel integration section is then sent to the fuel cell stack. The heat and mass balance is shown in Appendix C and a summary of results is provided in Table 1.

Table 39 Fuel Cell Stack Overview

Material Balance:

Stream No.		FC 1	FC 2	FC 3	FC 4	FC 5
Stream Name		Fuel to FC Stack	Hot Air to FC Stack	Cool Air to FC Stack	Spent Fuel from FC Stack	Vitiated Air from FC Stack
CO	- lbmol/h	1423	0	0	436	0
H2	- lbmol/h	3980	0	0	963	0
CH4	- lbmol/h	406	0	0	0	0
CnHm	- lbmol/h	46	0	0	0	0
H2S+COS	- lbmol/h	0	0	0	0	0
CO2	- lbmol/h	2335	7	15	3648	148
H2O(v)	- lbmol/h	3754	241	507	7418	1006
N2+Ar	- lbmol/h	185	18706	39317	160	58029
SO2	- lbmol/h	0	0	0	0	0
O2	- lbmol/h	0	4959	10424	0	12476
Total Gas:	- lbmol/h	12129	23914	50263	12625	71659
Gas MW	- lb/lbmol	19.09	28.85	28.85	24.82	28.71
Total Gas Mass F	- lb/h	231518	690013	1450287	313390	2057612
Total Flow	- lb/h	231518	690013	1450287	313390	2057612
HHV/Enthalpy	- Btu/lb	3656	0	0	547	0
Energy Flow	- MMBtu/h	846	0	0	171	0
Pressure	- psia	44	44	43	41	41
Temperature	- deg-F	1300	1285	920	1450	1452

Comp	Fuel	Hot Air	Cool Air	Spent Fuel	Vitiated Air
CO	11.7349%			3.4572%	
H2	32.8154%			7.6257%	
CH4	3.3496%			0.0002%	
CnHm	0.3769%			0.0000%	
H2S+COS					
CO2	19.2511%	0.0298%	0.0298%	28.8984%	0.2059%
H2O(v)	30.9506%	1.0090%	1.0090%	58.7551%	1.4035%
N2+Ar	1.5215%	78.2226%	78.2226%	1.2634%	80.9801%
SO2					
O2		20.7386%	20.7386%	0.0000%	17.4105%
Total:	100.0000%	100.0000%	100.0000%	100.0000%	100.0000%

Analysis of the complete heat and mass balances indicates that mass balances for the complete IGFC system are consistent within 0.002 % and energy balances are found to be consistent within 0.03 %.

Results for the overall net power output and efficiency for the pre-baseline system are given in Table 40.

Table 40. Pre-Baseline Performance Summary

Pre-Baseline System		
Gross Power Gen.		
-Gas Turbines	-kW	70217
-Net Fuel Cell System	-kW	90034
-Steam Turbine	-kW	65207
Sub-Total:	-kW	225458
In-Plant Power Cons.		
-Gasification	-kW	2810
-Air Separation	-kW	13158
-Combined Cycle	-kW	2149
-Cooling Water CC	-kW	365
-Cooling Water PP	-kW	782
-BOP+Misc	-kW	1283
Sub-Total:	-kW	20548
Net Power To Grid	-kW	204910
Heat Input, HHV		
Heat Input, HHV	-MMBtu/h	1625.2
Net Heat Rate, HHV	-Btu/kWh	7931.4
Net Efficiency, HHV	-%	43.0

4.1.6 Summary of Pre-Baseline system trade-offs

This pre-baseline IGFC system is presented as a commercially feasible integration of an advanced SOFC with current gas turbine technology. The overall efficiency of 43.0% presents a 2.2% improvement in system efficiency over the equivalent IGCC system.

Most of the system trade-offs in the pre-baseline system involve the integration of the fuel cell stack with a fixed gasifier system and a sub-optimized fuel integration system. Within the bounds of the available gasifier system information, and the medium-fidelity HYSYS simulation of the fuel integration system, the pre-baseline system results provide a realistic benchmark case for further optimization in order to yield an acceptable baseline system.

The CO level in the fuel gas for the fuel cell stack system could be kept within the recommended 15% maximum CO level (a nominal 12% CO level) by use of a single WGS reactor in the fuel cell subsystem since the syngas from the gasification system was highly saturated with H₂O. Effects of optimization of the shift level for the syngas from the gasifier system would be the subject of post-baseline studies.

Integration of fuel cell heat energy has not been fully optimized in the pre-baseline case because of the minimal heat energy integration with the gasifier and fuel integration systems. More extensive optimization of heat exchanger approaches, pinches and

configurations would be possible in future studies of IGFC systems with more complete integration of the fuel cell system and the gasifier system, as well as alternate cooling strategies for the fuel cell stack system.

This pre-baseline system configuration was set up with initial specifications of pressure drops, cell temperature rises, fuel utilization, and cell losses. The fuel cell stack operating pressure was limited to 3 bars consistent with status technology.

4.2 Design Concept Development

A baseline hybrid SOFC system was designed around the BGL gasifier. Since the gasifier system (gasifier, gas cleanup, and ASU) was fixed, efforts were centered around the design of the fuel cell modules and bottoming cycle and their integration with the gasifier system.

4.2.1 Approach and Analysis Basis

The following considerations were taken into account for proceeding from pre-baseline to baseline.

- Air management. Air is needed in the SOFC stack to supply oxygen as well as to cool the stack. Normally, about 5 to 10 times the stoichiometric amount of air is needed on the cathode side for heat removal. This air needs to be fed to the cell at an elevated temperature to ensure proper stack operation. The issue of supplying heat to raise the air to the operating temperature is one of the challenging problems. Once the air has been heated to the operating temperature, it could be passed through multiple stacks.
- Pressure drop in the piping. The piping to and from the stack is extensive, as air and fuel need to be supplied to every single cell. As a result the system experiences a substantial pressure drop. The piping must be made large enough to slow the gases down but not so large as to cause excessive cost penalty. As a general rule of thumb, all piping external to the fuel cell stacks was designed to keep fluid velocities below 30 m/s (about 100 ft/s). This solution proved quite satisfactory in terms of pressure drop and size.
- Heat loss from the piping. Loss of enthalpy along the length of piping could have a substantial effect on the plant performance. This problem can be solved easily and inexpensively by using a reasonable quantity of insulation, and by designing the plant so that the hot-side piping lengths are kept to a minimum.
- Use of large-sized turbomachinery components. For typical large industrial gas turbines, compressor efficiencies are in the high 80s and turbine efficiencies are in the low 90s. These are advanced, high performance components, and their use adds very large performance gains to the plant. The plant is designed to use one or two of these large-scale gas turbines. The alternative approach would be

to use microturbines for each fuel cell module. While this solution simplifies the piping issues, the degradation in performance is substantial because of the relatively poor component efficiencies associated with small-sized turbo-machinery.

- Use of SOFC as a topping cycle. Fuel cells are more efficient than combustion-based processes. In order to maximize cycle efficiencies, the entire topping cycle should consist of fuel cells; while the excess heat generated by the fuel cells is recovered in the bottoming cycle.

4.2.2 Plant description

Nominal plant layout

The IGFC plant is sized for a nominal power output of 300 MW. The unit is connected to a BGL gasifier unit, whose syngas output is scaled to 60 kg/s (132 lbm/s, 475200 lbm/hr). The syngas has a molecular weight of 19.1 and HHV of 8503.8 kJ/kg (3656 Btu/lb). The syngas composition fed to the fuel cell is shown in Table 41 (mass fractions are given).

Table 41. Fuel Cell Syngas Composition

	Mass Fraction
H ₂	0.03
CO	0.17
CH ₄	0.03
H ₂ O	0.29
CO ₂	0.44
N ₂	0.01
AR	0.01

The syngas output of the AGR system is distributed uniformly among the forty fuel cell modules in the plant. The air and fuel flow rates inside each module are shown in Figure 103.

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Overall mass flows for a 10-stage SOFC module

Figure 103. Overall Mass Flow

The plant consists of the following major components:

- the gasifier system (gasifier, ASU, clean-up system)
- forty SOFC modules
- one or two gas turbines
- one HRSG
- one steam turbine
- generators for the gas turbine and the steam turbine

Fuel Cell Modules

A SOFC module is a refractory-lined pressure vessel that includes the following components

- five to ten SOFC stages
- heat exchangers, mixers, or similar units in between SOFC stages
- current collector bus bars
- manifolds for fuel distribution

In this context a stage is viewed as a SOFC stack, or several stacks placed side by side. The baseline design is viewed as three stacks side by side. Thermal insulation with a thickness of 5 cm is placed inside the pressure vessel to preserve strength in the steel casing material and to reduce heat loss from the pressure vessel. The dimensions are in meters.

The stack characteristics are as follows.

- Cell area: 1500 cm²
- Current density: 0.7 A/cm² (power density 0.49 W/cm² at cell operating voltage of 0.7 Volts)
- Cell power 735 W; stack power 154.35 kW; stage power 463.05 kW

The above performance parameters are comparable to SECA goals of .35 W/cm² at 0.75 volts and 80% utilization. The chosen operating voltage is lower than SECA goals, while the power density is higher. It should also be noted that the SECA goals are for ambient pressure operation, while the IGFC stacks are expected to operate at significantly higher pressures.

Since a 3-stack stage makes 0.46 MW DC power, a pressure vessel with 10 stages would be able to produce 5 MW nominally. The pressure vessel would be 2 m in OD and about 14 m long.

The space between the stacks is large enough to bring the fuel manifolds in and out of the stack. The space between the stacks and the refractory liner would also house the fuel lines and electrical bus bars.

Plant layout

Since large gas turbines have higher component efficiencies, it is more efficient to compress the air going to the SOFC modules with one or two large compressors, and similarly expand the SOFC output flow with one or two large turbines. This approach necessitates a means to distribute the flow uniformly among all 40 of the modules and collect it to burn and expand in the turbine. Various concepts to do this were investigated, and it was decided that the simplest and the most effective arrangement was to have plenum chambers do the air distribution and collection.

The fuel cell modules were arranged in two rows (one on top of the other), adjacent to the gas turbine with two plenums on either side of the fuel cell modules. The plenum further away from the gas turbine is the cold side plenum, while the one closest is the hot side. The compressor outlet flow is collected in the cold side plenum (where the temperature is about 300°C) and distributed among the fuel cell modules. The large-diameter pipes providing passage to and from the plenums make sure that the pressure drop is minimal. The outlet flow from the modules is collected in the hot side plenum, after which it is mixed with the spent fuel from the modules and burned in the burner before passing into the turbine.

Since the length of the piping is different for each module, the piping must contain auxiliary pressure drop devices to ensure equal pressure differential in all the pipes.

Material Choices for the Fuel Cell Modules:

The primary assumption to meet the requirements of the SOFC module are given below:

- Hot section piping mean temperature is 800°C and the cold section piping mean temperature is 300°C.
- All the piping and the cold section plenum chambers are externally insulated. The hot section plenums and the fuel cell modules are refractory lined on the inside. Fiber wool with thermal conductivity 0.05 W/mK and 5 cm thickness is used for external insulation.
- The fuel cell module and the hot plenum are refractory lined.
- The only part of the piping system that needs stainless steel is the connection between the fuel cell module output and the hot plenum. Because the pipes are relatively small (30 cm ID), 304 stainless is used for this section of the piping. All the remaining piping system is of carbon steel.
- The following pipe dimensions were arrived at in compliance with ASME pressure vessel codes.
 - Cold side pipes 30 cm ID 0.5 cm wall 214/226
 - Duct 150 cm ID 2.75 cm wall 214/226

○ Plenum	500 cm ID	9.0 cm wall	214/226
○ Module	200 cm ID	3.75 cm wall	214/226
○ Hot side pipe	30 cm ID	4.25 cm wall	304

4.2.3 Pinch Analysis

In this analysis, heat integration in only one of the fuel cell pressure modules described in the plant layout section is analyzed since all other pressure modules will replicate the design. In order to understand the overall heat integration concept of the process streams in the pressure module, a simplified pinch analysis was performed. This analysis quickly revealed that, for number of cells, n , greater than three, the process hot streams have more heat content than needed to heat the cold streams.

4.2.4 Candidate Baseline Configurations

A number of candidate baseline configuration concepts were developed, and a baseline system was down-selected for further detailed analysis and optimization.

4.2.4.6 Down-Selection to Baseline concept

A design trade-off analysis was performed using standard Design for Six Sigma (DFSS) tools to down-select from the five proposed concepts. This down-selection was made according to the following three critical aspects of the proposed systems

- Achievable maximum efficiency
- Initial capital cost
- System reliability

Preliminary performance evaluation of all of the concepts was performed. The concepts were ranked according to the above-mentioned criteria. The outcome of the analysis is depicted in Figure 104. Concept 2 was selected as the “baseline” for its simplicity, which leads to a lower cost and higher reliability. Concept 4 was also selected for further study as the “alternate”. Although the alternate has the risk of higher cost and lower reliability, it has potential for higher efficiency.

Configuration	Rank	Efficiency	Stages	Recycle	Complexity	Cost Potential
Concept 1	3	52.5	10	Yes	Medium	Low
Concept 2	1	51.8	10	yes	Low	Low
Concept 3	5	51.0	10	No	High	Medium
Concept 4	2	52.5	5	No	High	Medium
Concept 5	4	52.2	10	Yes	High	High
<i>All systems evaluated at Pressure Ratio of 8.0</i>						

Figure 104. Down-Selection Trade-off Matrix

4.3 Baseline Design and Modeling

4.3.1 Description of the Baseline and Alternate

In this section the detailed layout of the baseline and alternate concepts are presented. The discussion centers on the SOFC and gas turbine part of the plant. The gasifier unit (including clean-up and ASU) and the bottoming cycle stay the same in both cases.

4.3.1.1 Baseline concept

The plant consists of several fuel cell modules. Gases are supplied to the modules by one or two large gas turbines and a syngas expander. Several modules are served by the same gas turbine system. Compressed air is fed to a plenum and distributed equally to the modules. On the fuel side, high-pressure syngas is expanded down to the fuel cell operating pressure and distributed to the various modules and cells inside the module in a similar way.

The design point number of stages per module is ten. The fuel cell stages have built-in fuel pre-heaters. The spent fuel from all the stages is collected and sent to a burner to burn with the cathode exhaust.

4.3.1.2 Alternate concept

In the alternate concept the compressed air from station is preheated to the fuel cell operating temperature through a set of parallel heat exchangers.

4.3.2 Major assumptions in the modeling

The pressure drops and the heat losses at the plant design point are given in Table 42 and are summarized graphically in Figure 105. The actual fuel cell stage pressure losses were varied from the nominal value as a function of the local flow properties at each stage. All components inside the pressure vessel stack modules have been sized for low airflow velocities.

Table 42. Design Point Air Pressure and Heat Losses

Design Point Air Pressure Losses			
Component	Units	Baseline	Alternate
Compressor Inlet	inch H ₂ O	3	3
Compressor Extraction	$\Delta P/P$ (%)	2	2
Cold Piping & Plenum	$\Delta P/P$ (%)	0.13	0.13
Recycle Mixer	$\Delta P/P$ (%)	0.1	0
Air Preheat HEX cold	$\Delta P/P$ (%)	0	0.1
Module Inlet	$\Delta P/P$ (%)	0.1	0.1
Fuel Cell Stage (nominal)	$\Delta P/P$ (%) each	0.1	0.1
Fuel Preheat HEX hot	$\Delta P/P$ (%) each	0.05	0.05
Air Preheat HEX hot	$\Delta P/P$ (%) each	0	0.1
Air Injection Mixer	$\Delta P/P$ (%) each	0.1	0.1
Module Outlet	$\Delta P/P$ (%)	0.1	0.1
Hot Piping & Plenum	$\Delta P/P$ (%)	0.2	0.2
Turbine Plenum	$\Delta P/P$ (%)	2	2
Burner	$\Delta P/P$ (%)	3	3
Back Pressure	inch H ₂ O	15	15
TOTAL (compressor exit to turbine inlet)	$\Delta P/P$ (%)	11.24	10.32

Design Point Heat Losses		
Component	Units	Loss
Cold Piping & Plenum	BTU/s	116.1
Pressure Module + Plenum + Hot Piping	BTU/s	161.4

Figure 105. Summary of Baseline Design Point Pressure Drops ($\Delta P/P$)

A maximum fuel cell exit temperature of 775°C was assumed. The largest allowable temperature gradient across the fuel cell stage was assumed to be 125°C, meaning that the inlet temperatures can be no less than 650°C.

A two-pressure steam cycle with reheat was used for the bottoming cycle. In order to use a two-pressure steam system, the minimum allowable temperature for the HRSG inlet was taken to be 1000°F (538°C). The maximum HRSG inlet temperature was 1280°F (693°C), set by material temperature limits. For the CO₂ isolation study, the lower temperature limit on the HRSG inlet was relaxed slightly to 900°F (482°C), although the bottoming cycle efficiency was correspondingly reduced.

4.4 Final Conceptual Design Performance Summary

The performance results for the baseline and alternate designs are discussed below.

4.4.1 Baseline Performance

Figure 106. Overall Baseline Performance vs. Pressure

Figure 106 shows the overall plant performance for the baseline design versus fuel cell operating pressure. The HRSG inlet temperatures are also shown in red. The plant efficiency increases with operating pressure for the same compressor inlet mass flow rate and fixed fuel utilization. At operating pressures above 10 bar, the HRSG inlet temperature drops below 1000°F. Below this temperature, the bottoming cycle performance begins to suffer as it becomes difficult to drive a two-pressure reheat steam turbine. This HRSG limit prevents operation at very high pressures and higher efficiencies. At low pressures (below 6 bar), the HRSG inlet temperature exceeds the assumed maximum (1280°F). The design point is shown as a star. At the design point pressure of 10 bar, the net plant efficiency is 53.4% (Coal HHV basis).

4.4.1.1 Sensitivity Analysis

The sensitivity of the baseline plant efficiency to various parameters was examined.

Figure 107. Effect of SOFC Fuel Utilization
10 stacks, P=10 bar

Figure 107 shows the effect of SOFC fuel utilization on plant efficiency and HRSG inlet temperature. Higher fuel utilization at a fixed stage exhaust temperature raises the plant efficiency, since the SOFC topping cycle is the most efficient use of fuel. However, as more fuel is consumed in the fuel cell, there is less fuel in the end of the stage exhaust available for combustion. The HRSG inlet temperature drops, limiting an upper limit of about 82% fuel utilization for a two-pressure reheat bottoming cycle. The design point was chosen to be 80% fuel utilization.

Figure 108 shows the effect of the number of stages in series per module on plant efficiency. The efficiency decreases slightly as additional stages are added in series. The cathode oxygen concentration decreases slightly as air passes through more stages in series. The design point was chosen at ten stages per module. Slightly higher efficiency could be achieved with fewer stages, but at an increased capital cost due to additional number of modules needed to achieve the same net plant power.

Figure 109. Effect of Fuel Cell Power Density

Figure 109 illustrates the effect of fuel cell power density at a fixed Area Specific Resistance (ASR). For a fixed technology cell, fuel cell efficiency decreases at higher power density. However, capital cost could be reduced as the total number of fuel cell modules will be reduced in proportion to the power density increase. Conversely, if efficiency was the sole criteria, the fuel cells could be operated at lower power densities and increased performance.

Figure 110. Effect of Area Specific Resistance

The sensitivity of the net plant efficiency to the fuel cell area specific resistance (ASR) is shown in Figure 110. At a fixed power density and fuel utilization, if fuel cell technology is improved and ASR is lowered, the plant efficiency increases as shown in this figure.

Figure 111. Effect of Fuel Cell Leak

The sensitivity of plant efficiency to fuel cell leak is shown in Figure 111. Total fuel leak (mass basis) is shown on the x-axis. Half the leak is assumed to be at the inlet of the cell, the other half is at the exit of the cell. Since the anode side is slightly pressurized versus the cathode, the fuel leaks into the cathode air. The design point was chosen at 2% fuel leak. Significant improvement in the fuel cell plant performance can be achieved if the fuel leak can be reduced. Such a reduction may be feasible with large area cells, as the total sealing length is reduced considerably.

Figure 112. Effect of Methane Addition

The effect of methane addition to the syngas is shown in Figure 112. This was done to simulate a gasifier that produces a higher methane content syngas. Methane addition raises the energy content of the fuel and also provides some additional cooling of the cells as the methane is internally reformed, improving the plant efficiency. In order to achieve a net plant efficiency of 60% (HHV basis), over 35% of the fuel energy must come from methane.

4.4.2 Alternate Design Performance

The performance of the alternate design is compared with the baseline design. Both systems were modeled with equal fidelity.

Figure 113. Efficiency Comparison of Baseline and Alternate

Figure 113 compares the performance of the baseline design and the alternate at a fixed operating pressure (10 bar). The alternate design has slightly higher efficiency by about 0.4%, regardless of the number of stages. Although the alternate system is capable of slightly higher efficiency, it may have higher capital costs and lower reliability than the baseline system, as discussed later.

4.4.3 Performance Summary

The power generation breakdowns for the baseline and alternate systems are given in Table 43. The reference IGCC case and the pre-baseline case have also been included for comparison. In the pre-baseline case, the efficiency was improved by 2.2% over the IGCC case simply by sending half the syngas to a fuel cell power island, with minimal heat and pressure integration. This concept also suffered from the fact that the stack operating pressure was only 3 bars, consistent with status technology. This system had only four stages in series and the pressure losses were assumed to be 2% per stage. The baseline system pressure module loss is only about 4% for ten stacks in flow series. In the baseline and alternate systems, the efficiency has been further improved

by sending all of the syngas to the highly-efficient fuel cell topping cycle and integrating the fuel cell island with the gas turbines and HRSG. All of these factors significantly increase the systems performance relative to the pre-baseline configuration. A review of the baseline and alternate system performance presented in Table 7 indicate that the primary difference in net performance is the recycle blower power consumption.

Table 43. Baseline Performance Summary

Configuration		IGCC	Pre-Baseline	Baseline	Alternate
Gross Power Gen.					
Gas Turbines	kW	180000	70217	96059	95563
Syngas Expander	kW	0	0	10162	10162
Net Fuel Cell System	kW	0	90034	181038	182912
Steam Turbine	kW	65200	65207	25518	24374
Sub-Total:	kW	245200	225458	312777	313011
In-Plant Power Cons.					
Gasification	kW	3211	2810	3196	3196
Air Separation	kW	15034	13158	14966	14966
Combined Cycle	kW	2190	2149	1333	1312
Cooling Water CC	kW	340	365	226	223
Cooling Water PP	kW	894	782	485	477
Recycle Blowers	kW	0	0	1554	0
BOP+Misc	kW	1379	1283	1459	1459
Sub-Total:	kW	23047	20548	23220	21633
Net Power To Grid	kW	222153	204910	289557	291378
Heat Input, HHV	MMBtu/h	1856.9	1625.2	1848.5	1848.5
Net Heat Rate, HHV	Btu/kWh	8358.6	7931.4	6383.9	6344.0
Net Efficiency, HHV	%	40.8	43.0	53.4	53.8

4.5 CO₂ Isolation

4.5.1 Description of CO₂ Isolation Concepts

Two methods for isolating CO₂ from the exhaust were considered. The first was a Selexol-based physical absorption system with anode recycle. The second method combusted the spent fuel with pure O₂ from the ASU rather than with the cathode exhaust. Ideally, this creates a fuel exhaust that is almost completely composed of CO₂ and H₂O. The water can be easily condensed out, isolating the CO₂.

4.5.2 CO₂ Isolation Results

The plant performance with CO₂ isolation and sensitivity to various factors are presented here.

Figure 114. Plant Performance with CO₂ Isolation by Selexol

Figure 114 shows the performance of the plant with 75% of the spent fuel being recycled to the CO₂ capture loop. CO₂ isolation via Selexol imposes a work penalty, thereby lowering the system efficiency compared to the baseline plant. CO₂ isolation via fuel recycle also limits the operating pressure to about 7 bar because less spent fuel is available for combustion, and the corresponding HRSG inlet temperature is lower. The

minimum HRSG inlet temperature required for a two-pressure reheat system has been lowered to 900°F to allow operation at reasonable pressure, even though the steam bottoming cycle performance suffers. Higher operating pressure and corresponding higher plant efficiency are possible if less spent fuel is recycled, but the amount of CO₂ captured will decrease. These trends are shown in Figure 115.

Figure 115. Effect of Fuel Recycle on HRSG Temperature

Figure 115 shows the effect of fuel recycle (mass fraction) on the HRSG inlet temperature. As additional fuel is recycled, less is available for combustion and the

turbine exhaust temperature decreases. To maintain a minimum HRSG inlet temperature of 900°F at an operating pressure of 7 bar, no more than 75% of the anode exhaust can be recycled. To maintain an HRSG inlet temperature over 1000°F at an operating pressure of 7 bar, no more than 50% of the anode exhaust can be recycled.

Figure 116. Effect of Fuel Recycle on CO₂ Isolation

Figure 116 shows the effect of fuel recycle on plant performance and CO₂ isolation. As more fuel is recycled and less is combusted, the fraction of CO₂ that is captured rises. At 75% fuel recycle, over 80% CO₂ capture is feasible. The net efficiency shown in red does not significantly increase because the operating pressure has been held constant at 7 bar. At lower fuel recycle fractions, the pressure ratio could be increased and efficiency would increase as described above.

Figure 117. Effect of Fuel Utilization on CO₂ Isolation

The effect of fuel use on plant performance and CO₂ isolation is shown in Figure 117. Because 75% of the spent fuel is recycled, fuel utilization has only a minimal effect. With anode recycle, the fuel cells could be operated at lower fuel utilization at the same plant efficiency, relaxing some of the technology requirements of the fuel cells. This is in contrast to the baseline concept, which has no anode recycle. As described in the Sensitivity Analysis section, for the baseline plant with no anode recycle, the efficiency decreases as the fuel utilization is lowered.

The performance summary for CO₂ isolation concepts is given in Table 44. The CO₂ isolation concepts have been modeled with both the baseline air recycle and alternate air preheat. For reference, the standard baseline and alternate without CO₂ isolation from Table 43 are also included.

Table 44. Performance Summary of CO₂ Isolation Concepts

Configuration		Baseline	Alternate	Selexol, Baseline	Selexol, Alternate	Pure O ₂ , Baseline	Pure O ₂ , Alternate
Gross Power Gen.							
Gas Turbines	kW	96059	95563	42699	42862	92206	91685
Syngas Expander	kW	10162	10162	18959	18973	10162	10162
Net Fuel Cell System	kW	181038	182912	226666	228040	181038	182912
Steam Turbine	kW	25518	24374	13320	11673	26067	24991
Sub-Total:	kW	312777	313011	301644	301548	309473	309750
In-Plant Power Cons.							
Gasification	kW	3196	3196	3196	3196	3196	3196
Air Separation	kW	14966	14966	14966	14966	23225	23225
Combined Cycle	kW	1333	1312	1164	1133	1369	1348
Cooling Water CC	kW	226	223	198	192	233	229
Cooling Water PP	kW	485	477	424	412	498	490
Selexol Plant	kW	0	0	2558	2558	0	0
Condensers	kW	0	0	299	299	0	0
Recycle Blowers	kW	1554	0	2558	0	1554	0
BOP+Misc	kW	1459	1459	1459	1459	1459	1459
Sub-Total:	kW	23220	21633	26822	24216	31534	29947
Net Power To Grid	kW	289557	291378	274822	277332	277939	279803
Fuel Utilization = 80%							
Heat Input, HHV	MMBtu/h	1848.5	1848.5	1848.5	1848.5	1848.5	1848.5
Net Heat Rate	MMBtu/kWh	6344.0	6344.0	6726.2	6665.3	6650.8	6606.5
Net Efficiency	%	53.4	53.8	50.7	51.2	51.3	51.6

Figure 118 summarizes the efficiency of the baseline, alternate, and various CO₂ isolation systems versus operating pressure. The design point for each system is shown with a star. The alternate system has the highest net efficiency at 10 bar of 53.8% on a coal HHV basis. The baseline system has a slightly lower efficiency of 53.4% at 10 bar. Although the alternate system is capable of slightly higher efficiency, it may have higher capital costs and lower reliability than the baseline system, as will be discussed later. Both systems improve with increasing operating pressure.

The Selexol-based CO₂ isolation system results in an efficiency of 50.7% for the baseline and 51.2% for the alternate. Both systems optimize at an operating pressure of about 7 bar. The pure-oxygen combustion CO₂ isolation system results in an efficiency of 51.3% for the baseline and 51.6% for the alternate. Both systems operate at 10 bar.

4.6 Cost Analysis

Estimates of the capital cost were made for the IGCC (reference) case along with the baseline and alternate cases. There were two variations for CO₂ separation, one based on Seloxol process and the other based on pure O₂ combustion. Table 45 summarizes the findings.

The following assumptions were made in this estimate.

- The SOFC costs are consistent with the cost targets laid out by the DOE SECA program on a \$/kW basis.
- Gasification costs are specific to the gasifier chosen for this study (BGL) and are estimated from internal GE numbers
- Other component costs were estimated based on scaling of existing components in a typical power plant.

The \$/kW capital cost for the baseline and alternate IGFC cases are smaller than the corresponding figure for the reference IGCC configuration. However, the total plant capital cost is higher for the IGFC configurations. Thus the relatively high efficiency of the IGFC plant plays a vital role to keep the \$/kW capital costs down.

The capital costs of the baseline and alternate concepts were estimated to be \$1654/kW and \$1700/kW respectively (Table 45). The difference in costs between the concepts is attributed to the differential cost of replacing the recycle blowers in the baseline configuration with the air heat exchangers. The analysis indicates that the air heat exchangers cost is higher than the recycle blower cost by \$45/kW on average. The difference is positive even after estimation errors are included. Hence, the baseline configuration will have a lower capital cost on a per-kW basis than the alternate configuration.

The baseline configuration is the preferred configuration despite its relatively lower efficiency compared to the alternate. The baseline system has a relatively lower initial capital cost on a per-kW basis than the alternate. Additionally, one of the most significant factors in the cost of electricity analysis of power plants is the plant's reliability. The baseline system is more reliable than the alternate system, as failure of one of the high temperature heat exchangers within the pressure vessel would be a reliability issue for the alternate. Standard practice for the baseline case would have "spare" recycle blowers available for quick changeover. In addition, the baseline configuration has the potential to be more cost effective by further optimization of the plant layout.

Similarly, the cost of the CO₂ isolation by direct O₂ combustion of the spent fuel stream is comparable within the ROM cost estimate to the Selexol plant. However, the Selexol

plant has the potential for lower cost and better performance by integration of the Sulfur removal.

The capital cost is one of several inputs to the cost of electricity (COE). It should be noted that SOFC stacks are projected to have a replacement period of five years compared to 10 years or more for gasifiers and gas and steam turbines. Therefore, the overhaul and maintenance costs over the life of the plant are higher for the IGFC plant than those for the reference IGCC plant due to the high SOFC replacement costs. These aspects should be taken into consideration in arriving at the merits of the IGFC plant.

Table 45. ROM Initial Capital Cost Summary

	IGFC ROM Capital Cost Overview All in (1000's \$, 2003)							
	IGCC Reference	IGCC Base	IGFC Baseline Air Recycle	IGFC Alternate Air Preheat	IGFC Baseline CO2 by Soloxol	IGFC Alternate CO2 by Soloxol	IGFC Baseline CO2 by O2 Burn	IGFC Alternate CO2 by O2 Burn
Basis Clean Syngas to AGR (lbs/hr)	542,691	254,190	253,045	253,045	253,045	253,045	253,045	253,045
Output (kW)	470,767	222,153	289,557	291,378	274,822	277,332	277,939	279,803
HHV Heat Rate (BTU/kWH)	8,421	8,359	6,384	6,344	6,726	6,665	6,651	6,606
Gasification process Capital Cost (K\$)	534,917	292,248	290,970	290,970	290,970	290,970	299,970	299,970
Gasification process Capital Cost (\$/kW)	1,136	1,316	1,005	999	1,059	1,049	1,079	1,072
GT+ST+HRSG Capital Cost (k\$)	168,500	88,158	75,564	75,564	75,564	75,564	75,564	75,564
CC Contingency & Owner's Cost (k\$)	44,230	23,141	19,835	19,835	19,835	19,835	19,835	19,835
Fuel Cell stack Capital Cost (k\$)	n/a	n/a	40,159	40,603	50,987	51,296	40,168	40,584
Fuel Cell Integration System Capital Cost (k\$)	n/a	n/a	26,301	38,859	32,134	47,478	21,531	31,812
Selexol System (K\$)					13,742	13,742	0	0
Fuel loop Integration(k\$)					28,082	28,082	58,798	58,798
FC Contingency & Owner's Cost (k\$)	n/a	n/a	26,168	29,581	43,707	47,815	38,084	40,891
Total Power Island Capital Cost (k\$)	212,730	111,299	188,027	204,442	264,051	283,813	253,980	267,484
Total Power Island Capital Cost (\$/kW)	452	501	649	702	961	1,023	914	956
Total Capital Cost (k\$)	747,647	403,547	478,997	495,412	555,021	574,783	553,950	567,454
Total Capital Cost (\$/kW)	1,588	1,817	1,654	1,700	2,020	2,073	1,993	2,028

4.7 Technology Gaps

It is not feasible, at present, to realize the plant concept presented in this report. Significant gaps exist in technology in certain areas, while in other areas, existing products must be extensively re-engineered to achieve the desired objectives. This section lists the gaps and the development requirements and concludes with recommendations for future work.

4.7.1 Gaps

Fuel Cells

- Size. The plant concept assumes reasonably large-sized cells (1500 cm^2). Large cells decrease the cell count in the plant, thereby improving reliability. It also mitigates effects like seal leakage, thereby improving overall performance. Cells currently produced are limited in size. Manufacturing large-sized planar cells could be a challenge.
- Fuel cell performance. The nominal design point for this study assumes a SOFC operating with a power density of 0.5 W/cm^2 at 80% fuel utilization and 0.7 Volts. This performance is also under typical shifted syngas composition. Practical cells meeting these objectives need development. No such planar cell presently exists with an area of 1500 cm^2 . The plant concept assumes the SOFC can operate under pressures up to 12 bars.
- Staged SOFC concept. The plant design relies on the staged fuel cell concept for air management and cooling. Conceptually, this design is sound, but it has not been proven by testing.

Overall Plant

- Controls: The plant controls technology under various load conditions for such a large fuel cell plant does not exist. The various requirements of the Gasifier, Gas Clean Up system and the fuel cell modules need to be integrated to ensure proper plant operation. The analysis performed in the evaluation is limited to plant operation under steady state design point conditions.
- Plant Start Up and Shut Down: Specific sequence of operation for both plant start up and shut down needs to be developed.

Turbomachinery

- Low-Btu combustor. The spent fuel sent to the combustion chamber in the gas turbine has a relatively small heating value ($\sim 500 \text{ Btu/lb LHV}$). This is because a high percentage of the fuel is utilized in the fuel cell and electrochemically

converted into steam. Combustor technology needs to be developed to burn this low-quality fuel.

4.7.2 Engineering developments

- Most of the plant layout surrounding the SOFC modules is novel and untested. Piping and plenums of similar size are found in other applications.
- No gas turbine presently exists with the particular combination of low pressure ratio, flow rate, and low firing temperature called for by this design. Additionally, this design requires 100% of the compressor air to be piped to the fuel cell modules before returning to the combustor and turbine. Engineering is needed to develop compressor and turbine plenums for 100% extraction with low pressure loss.
- The steam turbine in the bottoming cycle is relatively small (~20 MW). Commercial units of this class are available, but some engineering may be required to design a two-pressure reheat turbine of this size.
- No Selexol-based CO₂ capture plant has been built for this particular gas composition. The gas stream entering the Selexol plant is virtually free of sulfur and water, so it should be straightforward to engineer an optimized Selexol plant for this system.
- It is beyond the scope of this project to determine the final fate of the isolated CO₂. Technology that is being developed to inject CO₂ into oil wells, depleted natural gas wells, underground aquifers, and into oceans.
- The CO₂-lean fuel leaving the Selexol plant is almost completely dry. This is not a problem at the low temperature of the Selexol plant, but becomes a problem as the fuel is heated up to the fuel cell operating temperature. Steam must be added to the fuel to prevent carbon deposition in the recuperators and fuel preheaters. Technology to prevent carbon deposition has been developed and implemented in steam methane reformers.
- The spent fuel is sent to its own HRSG, separate from the HRSG for the exhaust air. The spent fuel is still pressurized (~65 psia) so the fuel HRSG must be contained in a pressurized shell. The product steam must be piped to the steam turbine, where it is combined with the steam generated from the air HRSG. Exhaust and re-heat steam must be bled off and returned to the fuel HRSG. Balancing the two HRSG cycles could be a difficult but feasible controls problem.
- If the alternate plant configuration outlined here is to be realized, then the high temperature, low pressure drop heat exchangers used inside the stack module should be engineered for size, cost, and reliability.

4.7.3 Recommendations for further work

- The power electronics design is outside the scope of this work. It is recognized that significant effort needs to go into designing the power electronics for these megawatt-sized SOFC modules.
- The overall plant startup can pose significant problems, and specific subsystems need to be engineered based on the startup strategy and controls requirement.
- The plant configuration outlined has significant volume and thermal inertia between the compressor and turbine, which is expected to have significant issues with plant controllability. This specific control technology needs to be developed.
- High temperature heat exchangers are among the most expensive components in the plant concept. Given their role in the realization of high efficiency SOFC power plants, ideas to make them more cost effective must be investigated.

17 5 CONCLUSION

The integration of a coal gasifier with a SOFC power generation system has been studied. The proposed plant concept includes an oxygen-blown gasifier system, a set of fuel cell modules, each containing several stacks, one or two large sized gas turbines, an HRSG, and a steam turbine. The possibility of CO₂ isolation from the exhaust products, either by an absorption method using Selexol or by pure O₂ combustion, has also been investigated. Rough Order of Magnitude (ROM) initial capital cost and net plant efficiencies were evaluated for all configurations.

Given the inherently high efficiency of SOFC modules, it is necessary to have the entire topping cycle consist of fuel cells. The spent fuel is then burned and the product expanded through large, high efficiency gas turbines. The gas turbine exhaust is then passed through the HRSG and a relatively small steam turbine for additional energy recovery. The CO₂ separation unit involves an additional HRSG with the spent fuel on the hot side and steam on the cold side. Most of the fuel cell exhaust is recycled and mixed with the shifted syngas before it is cooled and passed through the Selexol system for CO₂ absorption.

Various staging configurations were studied from a thermal management point of view. It was verified that maximum efficiency could be obtained from a staged, inter-cooled system of SOFC stacks. Various inter-cooling methods and associated cycle concepts were investigated. A baseline concept (with air recycle) and an alternate concept (air heat exchange) were downselected for further analysis. The alternate concept was found to have a slightly higher efficiency of 0.4% than the baseline. The ROM cost for the baseline configuration is lower by about \$45/kW compared to the alternate configuration. The baseline configuration is the preferred configuration despite the 0.4% lower plant efficiency compared to the alternate configuration. The baseline system is more reliable than the alternate system, as failure of one of the high temperature heat exchangers within the pressure vessel could be a reliability issue for the alternate configuration. In addition, the baseline configuration has the potential to be more cost effective by further optimization of the plant layout.

Similarly, the cost of the CO₂ isolation by direct O₂ combustion of the spent fuel stream is comparable with in the ROM cost estimate to the Selexol plant. However, the Selexol plant may offer more flexibility for integration of the sulfur removal and elimination of the AGR.

Several technology gaps are recognized. Several areas needing significant engineering development work are also identified.

It has been shown that plant efficiency of approximately 53% is possible for the proposed baseline plant layout. The efficiency penalty for CO₂ isolation is about 2.5%. This projected penalty would be about 4% if the CO₂ has to be delivered at 1000 psia for sequestration purposes. ROM initial capital costs for this plant have been estimated

to be on the order of \$1700/kW for a plant without CO₂ isolation and \$2000/kW with CO₂ isolation.

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Appendix C

Integrated Gasified Fuel Cell System Extension Study

IGFC System Efficiency Improvement Study

Summary

This work focused on evaluation of approaches for Integrated Gasified Fuel Cell (IGFC) system efficiency improvements. Using the two down-selected system concepts from the original IGFC study (Appendix B) as a starting point, opportunities to incorporate Ion-Transfer Membrane (ITM) and mid-range clean-up technology have been investigated and evaluated to determine their impact on system performance and cost. Incorporation of these two technologies is expected to yield 0.7% and 0.2% efficiency improvement, respectively. Incorporation of ITM technology may potential lead to a plant cost reduction of as much as 11% while the mid-range clean-up technology is not expected to have a significant impact on plant cost. Several system concepts have been identified that may have the potential of achieving the 60% system efficiency target. These concepts are currently being investigated.

IGFC SYSTEM EFFICIENCY IMPROVEMENT STUDY

1. Scope and Objective

The performance and economics of power generation systems based on the integrated gasified fuel cell concept were analyzed as discussed in Appendix B. Two alternative system configurations were selected in the task that integrated a coal gasifier with an SOFC, a gas turbine, and a steam turbine. The two down-selected systems were projected to have system efficiency of approximately 53% on an HHV basis or approximately 10 percentage points higher than that of state-of-the-art Integrated Gasification Combined Cycle (IGCC) systems. While the study showed great promise for coal-based fuel cell hybrid systems, it assumed currently available gasifier, gas turbine, and balance-of-plant technology. There are potentially substantial system efficiency gains that can be realized if advanced technologies, currently in the conceptual phase or under development, are employed.

Under this effort, the IGFC system design as had been created in the original study was used as a starting point to evaluate the potential areas for system efficiency improvements with the objective of achieving a feasible system with a 60% system efficiency (AC power/Coal HHV) including carbon dioxide isolation. In addition, two specific tasks included in this study were to identify opportunities to incorporate Ion-Transfer Membrane (ITM) and mid-range gas clean-up technology in the IGFC system design and evaluate their impact on system efficiency and cost.

2. Original IGFC Study

In the original IGFC study, the initial cost of both selected configurations was found to be comparable with the IGCC system costs, at approximately \$1700/kW. An

absorption-based CO₂ isolation scheme was developed, and its penalty on the system performance and cost was estimated to be approximately 2.7% and \$370/kW. In addition, an alternative CO₂ isolation scheme was developed that used pure oxygen to burn an anode exhaust stream; the burned exhaust gases, containing primarily CO₂, H₂O and N₂, is cooled to condense H₂O with subsequent isolation of CO₂.

For this study, the term “carbon dioxide isolation” refers to the separation of carbon dioxide from the exhaust stream without the need to pressurize the carbon dioxide stream. While it is recognized that to sequester CO₂ a high pipeline pressure may be required, the energy required to pressurize this stream was not factored into the study.

The concepts developed in the study were modeled to determine the effect of integrating an SOFC into a coal gasification power plant. As such, the system is considered to include the following islands:

- Coal Gasification, Cleanup and Air Separation Unit (ASU)
- SOFC and Gas Turbine
- Heat Recovery Steam Generator (HRSG) /Steam Turbine
- CO₂ isolation

Three concepts from the study are of particular interest:

Concept A: (without CO₂ Isolation): The gasification and ASU unit performance data were taken from an EPRI report (1). The gasification unit is based on a British Gas Lurgi (BGL) unit and the ASU includes a separate, conventional cryogenic plant. The cleanup of the synthesis gas is based on Lurgi’s low-temperature Purisol process. The acid gas is sent to a conventional Claus plant with tail gas recycled to the Purisol unit. The BGL unit includes a bitumen feed to bind coal fines in addition to tar and gas liquor treatment. All of the technology adopted in the gasification and ASU island is considered available technology if a commercial plant were to be built today.

The SOFC/Gas Turbine island includes the use of multiple, pressurized SOFC modules. Each module contains separate SOFC stages. The gas compressors and gas turbines are considered state-of-the-art; while a specific unit is not available, it could be designed with performances achieved as modeled.

The Heat Recovery Steam Generator/Steam Turbine (HRSG/ST) island was modeled using a simple relationship between the available enthalpy of the hot exhaust gases from the gas turbine. A multi-pressure steam system is required to provide the necessary steam levels to the gasification/cleanup processes; while a detailed modeling of the steam levels/turbines/extractions, etc. was not performed in this study, the overall efficiency conversion of hot-gas to the HRSG to steam-cycle AC-power is judged achievable with existing technology.

Concept B: (having CO₂ Isolation): This system uses the same basic process configuration as described above with the addition of a Selexol process to remove CO₂ from the syngas.

Concept C: (having CO₂ Isolation): This system also uses a very similar process configuration as described above except for the CO₂ isolation. A feature of the planar SOFC technology is the separation of anode and cathode exit streams. By burning the anode exit streams with a pure O₂ stream from the ASU the CO₂ remains separated from the bulk of the process air. Hence, it is kept separate and expanded in a separate turbine from the cathode exhaust. After heat recovery in the HRSG, the CO₂ rich stream is cooled to condense water.

A summary of the efficiency projections from the original IGFC study, as reported in 2004, is plotted as a function of operating pressure in Figure 119. As shown, the system efficiency is 53.5% without CO₂ isolation. With CO₂ isolation, the efficiency is reduced by 2.2% to 51.3%. The cases shown as “alternate” included the use of internal heat exchangers to the SOFC module; while showing slightly higher efficiency, it was judged less preferable as it added more technical risk (high temperature heat exchangers) and would be more costly.

Figure 119 Summary of system efficiency projects from original IGFC study (1)

For the three cases described above, a summary and comparison of the separate power generation and parasitic contributions is provided in Table 46.

Table 46 Power Generation and Parasitic Summaries (Jan 2004 Report)

IGFC Configuration in Jan 2004 Report		Cathode Recycle	Cathode Recycle, CO2 Isolation via Selexol	Cathode Recycle, CO2 Isolation via Pure-O2 Anode Ex Combustion
POWER ISLAND SUMMARIES				
GASIFIER /AIR SEPARATION UNIT ISLAND				
Coal + Bitumen feed (HHV)	kW	541,606	541,606	541,606
Net Steam Energy from HRSG/ST Island (Note 1)	MW	95	95	95
Clean Syngas to SOFC/GT Island (LHV)	kW	433,396	433,396	433,396
Clean Syngas to SOFC/GT Island (HHV)	kW	487,548	487,548	487,548
Syngas Generation Efficiency (LHV/HHV)	%	80%	80%	80%
Syngas Generation Efficiency (HHV/HHV)	%	90%	90%	90%
Gasification & Clean-up	kW	3,196	3,196	3,196
Air Separation	kW	14,996	14,996	23,225
Sub-total Gasifier/ASU Island Power Use	kW	18,192	18,192	26,421
SOFC/GT ISLAND				
SOFC				
SOFC dc Power	kW	188581	236110	188581
Inverter loss	kW	7543	9444	7543
SOFC AC Power	kW	181038	226666	181038
Syngas Expander	kW	10162	18959	10162
Cathode Hot Air Recycle Compressor	kW	1554	2558	1554
Net SOFC	MW	190	243	190
Gas Turbines				
Combusted Gas Turbine	kW	161149	105967	68419
Cathode Exhaust Turbine	kW	0	0	98520
Air Compressor	kW	65090	63268	74733
Auxiliary	kW	0	0	0
Net Gas Turbine	MW	96	43	92
Sub-total SOFC/GT Island	MW	286	286	282
HRSG/ STEAM TURBINE ISLAND				
Steam Turbines	kW	25518	13320	26067
Net Steam Energy to Gas/ASU island (Note 1)	MW	95	95	95
Auxiliary	kW	3504	3245	3559
Sub-total HRSG/Steam Turbine	MW	22014	10075	22508
CO2 ISOLATION				
Selexol Plant	kW		2558.00	
Condensers	kW		299.00	228.00
Sub-Total:	kW		2857.00	228.00
PLANT POWER SUMMARY				
TOTAL POWER GENERATED	MW	313	302	309
POWER CONSUMED	MW	23	27	32
NET POWER TO GRID	MW	290	275	278
NET EFFICIENCY, HHV		53.5%	50.7%	51.3%

NOTES

1. Sensible enthalpy 77F basis, no H2O condensed

3. Baseline System for IGFC Extension Study

From the cases described above, Concept C was chosen as the baseline (BL) system against which to compare other systems in the current task because it includes carbon

dioxide isolation, over 90% isolation is possible, and an increase in its system pressure would result in the improvement of system performance, in comparison to that of concept B.

3.1 Baseline-1 (BL-1) Concept Description

For the Baseline-1 System (BL-1), the exact same system configuration as concept C is used. However, two operational parameters were modified as follows:

- The SOFC operational pressure was increased from 10 bars to 12 bars. As noted above, and shown in Figure 119, an increase in pressure results in an increase in system efficiency.
- The number of stages in the SOFC module was reduced.

For the analysis of this system, the gasification and ASU unit performance data were taken from an EPRI report (3). The island consumes coal, bitumen, and limestone to produce a synthesis gas (syngas) acceptable for feed to the SOFC/Gas Turbine island. Major parts of the Gasification/ASU/Cleanup island include the following:

- Coal Handling & Preparation
- Coal gasification (BGL)
- Acid Gas Removal (Purisol)
- Sulfur Recovery (Claus & Tail Gas Treatment)
- Gas Liquor Treatment
- Air Separation (cryogenic, non-integrated with Gas Turbine)
- Fuel Gas Saturation
- CO Shift

The SOFC/GT island consumes the clean Syngas produced from the gasification island to produce AC power from the SOFC modules and the gas turbine generator.

Major parts of the island include

- SOFC modules
- SOFC Inverter
- Syngas expansion turbine
- Air Compressors
- Anode exhaust gas, pure oxygen combustor
- Gas turbine
- Gas turbine generator

- Heat exchangers

The HRSG/ST island recovers heat from the hot turbine exhaust gas to generate steam for the gasification/cleanup island and to produce power from steam turbines. Major parts of the island include

- Heat exchangers to recover heat from the vitiated air exhaust gases
- Heat exchangers to recover heat from the combusted anode exhaust gases
- Multi-steam level generation including several extractions for the gasification/cleanup island
- Steam turbines and generators
- Steam cycle balance-of-plant (BOP)

The CO₂ isolation island takes the anode combusted gases after being cooled in the HRSG/ST island and condenses most of the water to isolate the CO₂. Major parts of the island include

- Air-cooled Condenser, including a blower for cooling air
- Condensate collection

Performance Assumptions

Gasification/ASU/Cleanup:

The coal, coal ash, limestone, and bitumen compositions are given in the Appendix, Table 55 thru Table 58 and are taken from EPRI TR-100376 (3).

Basic performance parameters are given in Table 47; the synthesis gas produced in shifted as shown in Table 48.

Table 47 Baseline Gasification/Cleanup/ASU Performance Assumptions

Gasification/ASU/Cleanup		
Coal as received Feed	lbs/lb coal as recvd	1
Bitumen	lbs/lb coal as recvd	0.02
Limestone	lbs/lb coal as recvd	0.04
Gasification Steam	lbs/lb coal as recvd	0.35
Oxygen (pure)	lbs/lb coal as recvd	0.63
Fuel gas saturation steam	lbs/lb coal as recvd	1.05
Power Use	MW	3.2
Steam Net Thermal Energy Consumed	MW	95
Air Separation		
Total power	MW	23.2
Pure O ₂ to gasifier	1000 lb/hr	82.2
Pure O ₂ to Anode gas	1000lb/hr	47.1

Table 48 Baseline Synthesis Gas Compositions

Mole Fractions:	Saturated Synthesis Gas	Shifted Synthesis Gas
CO	0.2934	0.1173
H2	0.1521	0.3281
CH4	0.0335	0.0373
CnHm	0.0038	0.0000
H2S+COS	<40 ppmw	<40 ppmw
CO2	0.0165	0.1925
H2O(v)	0.4855	0.3095
N2+Ar	0.0152	0.0152
Total Gas, mph	24141	24141
Pressure, psia	352	305
Temperature, F	572	905

System Efficiency Sensitivity Analysis

Parameters included in the analyses on the baseline system and prior study concepts are shown in Table 49. While not all were applied to the new baseline system, the effects of each parameter identified in the prior study are judged to be relevant to the baseline with the exception of the fuel recycle for the concept that used selexol technology for CO₂ isolation. The parameters are further described below.

Table 49 Parameters Included in Sensitivity Analyses on Baseline and Prior Study Concepts

BASELINE & PRIOR STUDY CONCEPTS	GASIFIER	SOFC						CO2 ISOLATION
	Increases in CH4 Yield from Gasifier	Operating Pressure	Fuel utilization	# of stages in each module	Cell Power Density (@ constant ASR)	ASR (@ constant power density)	Fuel Leak into Cathode	% Fuel Recycle
Prior Study Concept A: IGFC Using Cathode Air Recycle, no CO2 Isolation	x	x	x	x	x	x	x	NA
Prior Study Concept B: IGFC Using Cathode Air Recycle, CO2 Isolation Using Anode Recycle with Selexol		x	x					x
BASLINE (=Prior Study Concept C: IGFC Using Cathode Air Recycle, CO2 Isolation Using Pure-O2 Anode Effluent Combustion)		x						NA

Increase in methane yield from the gasifier: Figure 120 shows the effect of methane addition to the synthesis gas to simulate a gasifier that produces a higher methane content syngas. Methane addition raises the energy content of the fuel and also

provides some additional cooling of the cells as the methane is internally reformed, improving the plant efficiency. To affect a 1% increase in system efficiency, the increase in CH₄ yield as a fraction of the baseline yield is 6%.

Figure 120 Baseline Sensitivity Analyses: Effect of Increased Methane Yield from Gasifier

SOFC operating pressure: The effect of pressure on the baseline is shown in Figure 119. For the baseline system, the plot indicates that increasing the pressure beyond the 12 atm would increase the system efficiency: further analyses indicated a maximum efficiency could be reached at about 18 atm. As the pressure increases, the relative power from the gas turbine increases while the power from the steam turbine decreases. This is caused by a decrease in the turbine exhaust temperature as the system pressure increases, thereby allowing less energy to be recovered by the HRSG/ST. To affect a 1% increase in system efficiency in the range from 6 to 12 atm, an increase in pressure of 4% is required. To affect a 1% increase in system efficiency in the range from 8 to 18 atm, a 7% increase in pressure is required.

SOFC fuel utilization: The effect of fuel utilization is shown in Figure 121. As the utilization increases, the SOFC power increases linearly while the energy to the gas and steam turbines decreases. To affect a 1% increase in system efficiency in the range from 70 to 90%, a 17% increase in fuel utilization is required.

Figure 121 Baseline Sensitivity Analyses: Effect of SOFC Fuel Utilization

SOFC power density: In the prior study, an overall stack power density of 500 mw/cm² was assumed. Hence, if the fuel utilization, cell Area Specific Resistance (ASR) as represented in a voltage-current polarization curve of the cell, and operating conditions are kept constant, increasing the power density beyond 500 mW/cm² result in a lower cell voltage which decreases the system efficiency.

SOFC voltage: The effect of SOFC voltage is shown in Figure 122. As with fuel utilization, increasing the cell voltage increases the amount of energy taken from the fuel cell while decreasing the energy available to the gas and steam turbines. However, increasing cell voltage also decreases the amount of cooling air required to the SOFC. To affect a 1% increase in system efficiency in the range from 0.7 to 0.77 volts, an increase of 0.04 volts is required.

Figure 122 Baseline Sensitivity Analyses: Effect of SOFC Voltage

SOFC Area Specific Resistance (ASR): The effect of the fuel cell performance expressed as ASR is shown in Figure 123; this analysis is similar to changing the cell voltage as a power density of 0.49 w/cm^2 was maintained in the analysis. To affect a 1% increase in system efficiency in the range from 0.2 to 0.1 ASR, a decrease of 0.07 ASR is required.

Figure 123 Baseline Sensitivity Analyses: Effect of SOFC ASR

SOFC anode fuel leak into cathode: The effect of leaking anode gas in the fuel cell performance is shown in Figure 124. The anode is maintained at a slightly higher pressure than the cathode flow in the SOFC so as to ensure any leakage would occur from the anode into the cathode chamber. Such a leak not only causes a loss of fuel utilized in the SOFC, but also affects the cathode temperature as it is assumed to combust entirely. Hence, the cooling effect of the air in the SOFC is negatively affected. To affect a 1% increase in system efficiency in the range from 0.2 to 0.1 ASR, a decrease of 0.033 % leakage is required.

Figure 124 Baseline Sensitivity Analyses: Effect of SOFC Anode Leakage into Cathode

4. Study Subtasks

Specific subtasks to be completed in the current study include:

- Impact of Ion-Transfer Membrane (ITM) on the System Efficiency and Cost: Opportunities to incorporate an ITM into the IGFC system design are to be identified, including options in the gasifier, the fuel cell and the gas turbine subsystems. System analyses are to be performed to quantify effects of introducing an ITM into the baseline IGFC on the system efficiency and cost.
- Impact of Mid-Range Gas Cleanup on the System Efficiency and Cost: Opportunities to incorporate a mid-range gas cleanup subsystem into the IGFC system design are to be identified. The effects on the system efficiency and cost are to be estimated.
- Analyses of System Efficiency Improvement Opportunities: Brainstorming is to be performed to identify ideas to improve the system efficiency. Both cycle-level and component level opportunities shall be included. Analyses to quantify these improvements shall be conducted. Recommendations for any necessary future technology improvement efforts shall be made. In addition, impacts on the system cost shall be analyzed.

- Impact of Carbon Dioxide Separation: As the systems to be identified will include CO₂ isolation, the impact of the CO₂ separation on the system efficiency on the down-selected systems will be evaluated.

4.1 Impact of Ion Transport Membrane (ITM) on System Efficiency and Cost

ITM performance and costs estimates are based on public literature and telephone conversations with the developer, Air Products (4-9).

4.1.1 ITM Technology Characteristics

Technology Description: ITM technology is based on ceramic oxide materials that lose oxygen from their crystal structure at temperatures of 700°C and above, and where the loss of oxygen creates holes or vacancies in the crystal lattice. Under an applied partial pressure oxygen gradient, oxygen ions move from vacancy to vacancy within the lattice. Electrons move counter to the oxygen ions to maintain the charge balance in the material, displaying the so-called “mixed conductor” effect, as shown in Figure 125. Figure 126 shows how the devices could be stacked in a planar supported membrane device that includes laminated planar supported membrane wafers, spacer rings between each wafer, and an oxygen withdrawal tube(4-7).

Figure 125 ITM Mixed Conductor Effect

Figure 126 ITM Planar-Supported Membrane Stacking Principle

The performance of the ITM is described by the following set of equations (8):

$$R = \frac{F_{perm}}{X_{feed} F} \quad (1)$$

$$R_T = 1 - \frac{(1 - X_{feed})}{X_{feed}} \frac{P_{perm}}{(P - P_{perm})} \quad (2)$$

$$PX_{feed} \cong 7P_{perm} \quad (3)$$

$$X_{np} = \frac{X_{feed}(1 - R)}{(1 - RX_{feed})} \quad (4)$$

Where

R is the O₂ recovery fraction;

X_{feed} is the mole fraction of O₂ in the feed;

F is the feed molar flow rate;

P is the total pressure of the feed;

P_{perm} is the pressure of O₂ permeate; and

R_T is the maximum theoretical recovery.

The fraction of theoretical recovery recommended is 0.75. The device operates isothermally with a large pressure differential between the inlet feed and permeate but a small pressure difference from the air feed to the vitiated air stream. A sample calculation using the above equations and recommendations is provided in Table 50.

Table 50 Example Calculation of ITM Oxygen Recovery

SAMPLE CALCULATION OF ITM OXYGEN RECOVERY		
Required O2	- lbmol/h	3952
Feed to ITM Conditions		
Xf, Mole fraction O2	- mole fraction	0.141
P, Total Feed Pressure	- psia	172
Pperm [=P*Xfee/7] (recommended)	- psia	3.5
Rt theoretical recovery [=1-(1-Xf)/Xf*(Pperm/(P-Pperm))]	- fraction	0.875
frRt, fraction of theoretical recommended	- fraction	0.800
Actual O2 recovery fraction [=Rt*frRt]	- fraction	0.700
Total Air Feed flow Required to ITM	- lbmol/h	40139

Testing: Small scale testing in 2000-2001 by Air Products included experiments conducted at the appropriate driving force for oxygen flux at approximately the commercial operating conditions of 800 to 900°C, 200 to 300 psig feed gas pressure, and typically <1 atmosphere oxygen withdrawal pressure. In addition, a 0.1 tons O₂ per day technology development unit has been operated.

Costs: The ITM has been under development by Air Products for the Department of Energy with a cost goal described as follows:

“To commercialize a tonnage-quantity ITM oxygen production process at one-third lower cost compared to competing, conventional technologies.”

Several references, such as those given above, provide a cost comparison of a cryogenic vis-à-vis ITM in a coal Integrated Gasification Combined Cycle (IGCC) power plant. The costs are given at the unit level. In other words, individual component costs are not provided. For example, the costs are given on an oxygen short-Ton Per Day (TPD) basis as follows:

	<u>\$/s TPD</u>
• ITM ASU	\$13,000
• Cryogenic ASU	\$20,132

4.1.2 ITM IGFC Concept Configurations

Four system configurations incorporating the ITM technology were reviewed. Configurations A1 and A2 are similar in that compressed air from the gas turbine air compressor is fed to the ITM directly after being preheated. The vitiated air from the ITM is fed to the cathode side of the SOFC. However, it is substantially reduced in oxygen, thereby resulting in either a low SOFC voltage or an increase in air to the system to compensate for the lower oxygen partial pressure. In A1, a permeate oxygen stream is sent to the anode exhaust burner, similar to the baseline system described above. In A2, permeate for both the anode gas burner and the gasifier are supplied. As the permeate is at low pressure, it must be compressed for the anode gas by a three-stage,

inter-cooled compressor since the combustor operates at about 11 atm. For the gasifier oxygen, an additional compressor stage is required as it operates at about 27 atm.

Configurations B1 and B2 are similar in that hot, vitiated cathode exhaust air acts as the feed to the ITM. As in A1 and A2, oxygen needs to be compressed for the anode gas burner (3-stages) and the gasifier (an additional stage).

Analyses indicate both configurations A1 and A2 would be less efficient than configurations B1 and B2, primarily due to the need for increased air flow.

4.1.3 ITM Impact on System Efficiency

Configurations B1 and B2 were further analyzed. For configuration B1 the ITM utilizes the cathode exit air stream to create a pure oxygen stream for the anode gas combustor. For this case, the amount of air required for the ITM to supply the anode O₂ is in excess. Hence, some air is by-passed around the ITM. The ITM permeate, at less than 4 psia, must be compressed in a 3-stage inter-cooled compressor to 173 psia.

For Configuration B2, the ITM utilizes the cathode exhaust air to create oxygen for the anode gas combustor and the combustor. The amount of air required to the ITM to supply the anode O₂ is still in excess; hence, air is by-passed around the ITM. As in Case B1, the ITM permeate must be compressed in a 3-stage, inter-cooled compressor to 173 psia for the anode gas combustor. For the gasifier, the oxygen must be additionally pressured to 465 psia by a single-stage compressor.

A comparison of the BL configuration, which uses a conventional cryogenic ASU, to Configurations B1 and B2 is given in Table 51. Each case uses the same coal feed. Configuration B2 shows a 0.7% advantage over the BL (53.5 % versus 52.8%). The efficiency improvement is due to the reduction in parasitic power associated with the ASU vis-à-vis the ITM: the net parasitic power for oxygen generation is reduced from 23 MW to 8 MW. Some similarities and differences in power required and/or generated between the systems are noted as follows:

- The coal feed for all three system concepts are identical. However, the syngas from the ITM cases is slightly more concentrated in fuel constituents as the oxygen from the ITM is purer (99.7v%) than the oxygen from a cryogenic (95v%) oxygen plant
- The power consumed in the ASU for the baseline case is 23 MW verses 8 MW for the ITM case. This represents a 65% efficiency reduction. It should be noted that the original cryogenic ASU baseline configuration is a “non-integrated” configuration (13). If the ITM were to be compared to an “integrated” configuration, where the air to the cryogenic plant is supplied using a bleed air stream from the gas turbine as air intake, the reduction in power attribute to the ITM compared to the cryogenic plant, could be less.
- The power attributed to the ITM is for compression of the permeate oxygen (at 3.5 psia) to the required pressure for gasification and anode gas burning.

- Power generated from the SOFC and expansion turbine are the same for all three cases. Likewise, power consumed from the inverter and recycle compressor are the same as the generated and consumed for all three cases
- Air flow for all three cases is the same. Therefore, the air compressor power is the same. Likewise, the un-utilized fuel from the SOFC is the same for all three cases, resulting in the same power from the combustion gas turbine. However, the vitiated air flow from the ITM is lower than the baseline, resulting in lower cathode exhaust turbine power.

Table 51 Power Generation and Parasitic Summary Comparisons: Baseline-1 & ITM Cases

SYSTEM / ISLAND		BASELINE-1 (BL-1)	ITM Case B1	ITM Case B2
GASIFIER /AIR SEPARATION UNIT ISLAND				
Coal + Bitumen feed (HHV)	MW	542	542	542
Clean Syngas to SOFC/GT Island (LHV)	MW	433	433	433
Clean Syngas to SOFC/GT Island (HHV)	MW	488	488	488
Syngas Generation Efficiency (LHV/HHV)	%	80%	80%	80%
Syngas Generation Efficiency (HHV/HHV)	%	90%	90%	90%
Gasification & Clean-up auxiliary	MW	3	3	3
Air Separation	MW	23	15	-
ITM	MW	-	3	8
Sub-total Gasifier/ASU Island Power Use	MW	26	21	12
SOFC/GT Island				
SOFC				
SOFC dc Power	MW	192	192	192
Inverter loss	MW	8	8	8
SOFC AC Power	MW	184	184	184
Syngas Expander	MW	7	7	7
Cathode Hot Air Recycle Compressor	MW	2	2	2
Net SOFC	MW	190	190	190
Gas Turbines				
Combusted Gas Turbine	MW	73	73	73
Cathode Exhaust Turbine	MW	107	104	98
Air Compressor	MW	82	82	82
Auxiliary	MW	0	0	0
Net Gas Turbine	MW	99	95	89
Sub-total SOFC/GT Island	MW	289	285	279
HRSG/ Steam Turbine Island				
Steam Turbines	MW	29	28	26
Auxiliary, ST island	MW	5	5	3
Sub-total HRSG/Steam Turbine	MW	24	23	23
CO2 ISOLATION				
Condensers	MW	0.3	0.3	0.3
Sub-Total CO2 ISOLATION	MW	0.3	0.3	0.3
TOTAL POWER GENERATED	MW	319	315	307
POWER CONSUMED	MW	34	28	17
Net Power To Grid	MW	286	287	290
System Efficiency, % of HHV Input	%	52.8	53.1	53.5

4.1.4 TM Impact on Cost

The cost impact, based on a ROM-type estimate, is provided in Table 52.

Table 52 Cost Comparison with ITM

	SUMMARY MM\$	NET POWER MW	COST \$/kW	COST REDUCTION %
BASELINE-1	\$488	285	\$ 1,710	--
ITM-B1	\$453	287	\$ 1,578	8%
ITM-B2	\$447	289	\$ 1,544	11%

4.2 Impact of Mid-Range Gas Cleanup on System Efficiency and Cost

Mid-range gas cleanup performance and cost estimates are based on public literature, telephone conversations with the developer, RTI, and GE engineering judgment (10-16). Cost data were not available in the public literature. Other data regarding systems required to remove trace compounds such as Hg, were also not available.

4.2.1 Mid-Range Gas Cleanup Technology Characteristics

Technology Description: Research Triangle Institute (RTI) and Kellogg Brown and Root (KBR) are jointly developing a pressurized, mid-range temperature desulfurization process capable of desulfurizing a sour and wet syngas. The process includes KBR's patented transport reactor technology and RTI's proprietary zinc oxide based sorbent for sulfur removal from gaseous streams, called RTI-3. KBR's unit consists of two coupled transport reactors, a sulfur absorber and a sorbent regenerator, as shown in Figure 127. RTI-3 zinc oxide sorbent removes H₂S and COS from syngas by circulating it continuously in the absorber and also transferring a portion of it from the absorber to the regenerator and vice versa. The following reactions occur:

- $\text{ZnO} + \text{H}_2\text{S} = \text{ZnS} + \text{H}_2\text{O}$ (absorber)
- $\text{ZnS} + 3/2 \text{O}_2 = \text{ZnO} + \text{SO}_2$ (regenerator)

Figure 127 Warm Gas Cleanup Process

Testing: Production of RTI-3 has been done up-to 200 to 1000 lb batches by Süd-Chemie, Inc. of Louisville, KY. Using a simulated syngas stream with 60 vol% steam and 0.7% (7,000 ppmv) H₂S and 0.035% (350 ppmv) COS, RTI-3 consistently reduced the effluent sulfur concentrations to non-detectable levels at 600°F and 130 psig. Regeneration of the sorbent was successfully demonstrated at temperatures above 1000°F with significant sulfur removal per pass. Post-test analysis of the RTI-3 confirmed earlier attrition test results with essentially no fines generation (< 20 micron) or change in the particle size distribution. In a 3 ton per day gasification unit at ChevronTexaco's Montebello Technology Center in South El Monte, CA, 17,200 scfh of warm syngas enters the absorber at 600 to 800°F. RTI-3 is continuously regenerated in the regeneration reactor. This reaction occurs at 1230 - 1300°F. As the reaction is highly exothermic, typically 6 to 10 volume % O₂ in N₂ is used as the oxidant gas. The oxidant enters the regenerator at about 1100°F and from the heat generated in the regenerator reaction, the temperature increases to 1250°F to 1300°F. Typically, sulfate formation is avoided by only partially regenerating the sorbent. The high temperature of the returning, regenerated sorbent provides the energy needed to boost the temperature of the incoming syngas from 500 °F to 600 – 800 °F.

As noted in the literature sources, RTI has done a significant amount of work to prove the capabilities and reliability of their system. Of particular interest is their collaborative work with other organizations including Chevron/Texaco and KBR. In addition, there has been recognition of the necessity to develop mercury and ammonia cleanup processes, which are critical to producing an acceptable gas for a power production application. They have also developed a Direct Sulfur Recovery Process (DRSP) that

would replace the CLAUS Plant in the sulfur removal process, with a slight possible increase in efficiency of sulfur removal.

In summary, RTI has developed a sulfur removal process from syngas with the following characteristics:

Operating Temperature Ranges:

- 500°F to 1200°F Overall Range
- 700°F to 800°F Optimum Range

Demonstrated Capabilities:

- Syngas with 5% H₂O content – H₂S and COS below 1 ppm
- Syngas with 20% H₂O content – H₂S and COS below 1 ppm
- Syngas with 50% H₂O content – H₂S and COS below 1 ppm
- Syngas with 63% H₂O content – H₂S and COS at single digit ppm level

Costs: Although their system is cost competitive (~\$60-\$75/KW) with other sulfur removal schemes such as Amine and SELEXOL plants, the main advantage of their system is obviating the need to cool the syngas down to the 100°F level necessary for traditional sulfur removal systems.

Sulfur Recovery Options and Chemistry: The process described above and utilized in this study includes the DSRP process for recovering sulfur after it is removed from the syngas. Three alternatives of many possibilities for conversion of SO₂ to a solid or liquid product from the Warm Gas Cleanup system are as follows:

1. Elemental sulfur from a Direct Sulfur Recovery Process: Figure 128 shows how the RTI regenerator off-gas (ROG) is processed in a Direct Sulfur Recovery Unit from the Regenerator Off-Gas. As shown in the sulfur chemistry given in the diagram, for each mole of sulfur, 2 moles of an equivalent hydrogen reducing gas (H₂ + CO) are required. In our baseline system, this represents over 1% of the synthesis gas produced;
2. Sulfuric acid production: Figure 129 shows the production of sulfur from the ROG. This requires no fuel gas slip stream;
3. Reduction of SO₂ with Anthracite: Figure 130 shows the ROG used a direct reduction process with anthracite (or coal) to convert the SO₂ to elemental sulfur.

Figure 128 Warm Gas Cleanup System: Sulfur Recovered as Elemental Sulfur

Figure 129 Warm Gas Cleanup System: Sulfur Recovered as Sulfuric Acid

Figure 130 Warm Gas Cleanup System: Sulfur Recovered as Sulfur from Anthracite Reduction

4.2.2 Mid-Range Gas Cleanup Concept Configurations

Baseline Purisol Process: In the baseline system, using the BGL gasification system, raw syngas at approximately 280 °F is cooled to approximately 100 °F and sent to the Purisol process for sulfur removal from the syngas, as shown in Figure 131. The Purisol system consists of a Purisol absorption unit, which absorbs sulfur from the syngas and yields a clean syngas and an acid gas stream. This acid gas stream is sent to a Claus unit, which converts the acid gas to elemental sulfur. The Purisol unit requires steam for the reboiler, and pumping power for the solvent circulation pumps. The Claus unit requires fuel gas for oxidation of the acid gas. An additional heat load in the Purisol system is the heat load resulting from the difference of heat duties between the cool-down of the raw syngas to the Purisol operating temperature.

Original Purisol System

P_{CD} = Cooldown Equivalent Power = 1610 KW

P_{RB} = Reboiler Equivalent Power = 610 KW

P_{PW} = Pumping Equivalent Power = 580 KW

P_{FG} = Fuel Gas Equivalent Power = 110 KW

**Total Required Power:
2900 KW**

Figure 131 Low-Temperature (Purisol) Gas Cleanup Process (original) in Baseline IGFC

Mid-Range Gas Cleanup in Baseline: Substituting the RTI system into the baseline, raw syngas at approximately 600 °F is sent to the RTI system for removal of sulfur from the syngas as shown in Figure 132. The RTI system consists of a warm gas absorption unit, which absorbs sulfur from the syngas and yields a clean syngas and an acid gas stream. This acid gas stream is sent to a Claus unit, which converts the acid gas to elemental sulfur. Additional mercury and ammonia absorption units must be included in order to provide a clean syngas suitable to the requirements of the fuel cell stack. However, the processes to remove these contaminants are not well defined in the literature cited above.

Figure 132 Mid Range Gas Cleanup Process (RTI) Inserted in Baseline IGFC

4.2.3 Mid-Range Gas Cleanup Impact on System Efficiency

Analysis Method: A differential analysis was performed to determine the effects of replacing the Purisol acid gas removal system, which is part of the BGL gasifier system, with an RTI warm gas sulfur removal system. All heat loads in the process systems were converted to an equivalent electrical power produced by the steam turbine. The total required power for each system was determined and the net difference is the power saving resulting from replacement of the Purisol system by the RTI system.

The resulting power requirements, and the total required power, for the Purisol system and substituted mid-range gas cleanup processes are shown in Figure 131 and Figure 132 to be 2,900 and 1,800 equivalent kilowatts of power. The net power savings estimate via this differential analysis procedure is 1,100 kW. This represents approximately a system efficiency performance improvement of only 0.2% based on a coal and bitumen feed of 541 MW.

4.2.4 Mid-Range Gas Cleanup Impact on Cost

There were no specific cost references in the literature cited above to indicate the costs of the mid-range gas cleanup system. However, in a telephone conversation, as noted above, costs were estimated to be in the order of \$60-\$75/kW of net system power produced. This is comparable with other typical gas cleanup processes. It is judged that the cost impact, assuming the performance estimated above, is within the fidelity of the ROM-type estimate.

4.3 Analysis of System Efficiency Improvements

As shown above, the baseline efficiency was improved to 53.5% by inserting the ITM for oxygen. Another 0.2% may be achieved by inserting a mid-range gas cleanup process. The baseline configuration with these performance improvement incorporated is still well below the 60% efficiency goal for this study. Table 53, shows the power production

and consumption for the baseline prior to the incorporation of the advanced technologies. Note that the total power generators before subtracting any of the power consumers only adds up to 56%, well below the target 60% efficiency. Hence, for this configuration to reach 60%, the efficiency of the generators would have to be improved significantly. Any improvements that the ITM would have, even if reduced to zero power for the ASU, would not result in a system efficiency of 60%. Likewise, if the power associated with gas cleanup were reduced to zero, the system would still not achieve 60% efficiency.

Table 53 BL-1 System Net Power Generators and Consumers

BASELINE-1 NET POWER GENERATORS & CONSUMERS	% of Feed HHV
Power Generators:	
SOFC AC	33.6%
Gas Turbines (Cathode Exhaust+CombGas -Air Comp)	18.3%
Syngas Expansion Turbine	1.4%
Net Steam Turbine	4.4%
Subtotal Generators	57.6%
Power Consumers:	
ASU (cryo)	4.3%
Gasifier & Clean-up	0.6%
CO2 Isolation	0.1%
Subtotal Power Consumers	4.9%
NET System	52.7%

Further analysis of the baseline power generation and consumer is illustrated in Figure 133. Since the SOFC is the largest generator, an improvement to its performance is a likely candidate to increase efficiency. This will be analyzed further below.

Figure 133 Baseline IGFC Power Generators and Consumer Comparisons for BL-1

4.3.1 Entitlement Analysis for Baseline Configuration with ITM and Mid-Range Gas Cleanup

The following “entitlement analysis” is provided as a means to estimate which selected processes may possibly be improved to affect an overall improvement in the system efficiency. “Entitlement” is a term that implies the maximum possible benefits from engineering and/or technical developments. The analysis presented here represents GE engineering and technical judgment.

BGL Gasifier Entitlement: A differential analysis was performed to determine the effects of an optimum BGL gasifier in the baseline system. In order to maximize the level of chemical energy (for conversion to power in the SOFC), the BGL gasifier output syngas was modified to contain no CO₂. The level of other constituents in the syngas was maintained at the original level. This resulted in decreased coal, limestone, oxygen and process steam requirements for the gasifier.

An energy balance for both the original BGL IGCC system, and the modified BGL IGCC system with maximum entitlement are given in Table 54.

Table 54 Gasifier Entitlement Analysis

Original BGL IGCC System

Performance Summary:		
Gross Power Gen.		
- Gas Turbines	- kW	384000
- Steam Turbine	- kW	135100
Sub-Total:	- kW	519100
In-Plant Power Cons.		
- Gasification	- kW	6855
- Air Separation	- kW	32097
- Combined Cycle	- kW	3781
- Cooling Water CC	- kW	773
- Cooling Water PP	- kW	1908
- BOP+Misc	- kW	2919
Sub-Total:	- kW	48333
Net Power To Grid	- kW	470767

Heat Input, HHV	- MMBtu/h	3964.4
Net Heat Rate, HHV	- Btu/kWh	8421.2
Net Efficiency, HHV	- %	40.53

**BGL IGCC System
With
Maximum Entitlement**

Performance Summary:		
Gross Power Gen.		
- Gas Turbines	- kW	383500
- Steam Turbine	- kW	119170
Sub-Total:	- kW	502670
In-Plant Power Cons.		
- Gasification	- kW	6821
- Air Separation	- kW	27926
- Combined Cycle	- kW	3525
- Cooling Water CC	- kW	721
- Cooling Water PP	- kW	1779
- BOP+Misc	- kW	2800
Sub-Total:	- kW	43570
Net Power To Grid	- kW	459099.9

Heat Input, HHV	- MMBtu/h	3788.3
Net Heat Rate, HHV	- Btu/kWh	8251.5
Net Efficiency, HHV	- %	41.36

Based on these estimates, the maximum BGL gasifier entitlement is of the order of ~ 0.8% for the IGCC system. Since it is improbable that the gasifier can operate under conditions corresponding to this maximum entitlement, it was modified by a 'realization factor' of 0.3 to yield an assigned entitlement increase of 0.3%.

Sulfur Removal Entitlement: Additional efficiency improvements for the IGFC system may be possible if CO₂ could be removed from the syngas process stream in the sulfur removal block of the gasification process island. The syngas exiting the reactor is saturated, and a sour gas shift replaces the previous shift reactor. This shift is followed by the RTI gas cleanup system for H₂S and COS sulfur removal, and 90% of the CO₂ in the syngas stream is removed by a Selexol CO₂ removal process. The resulting syngas could then be sent to the SOFC.

Energy requirements for the syngas saturation before the sour gas shift reactor appear to be similar to saturation requirements for the original BGL gasification system. Reasonably detailed calculations for the Selexol CO₂ removal process indicate that a net equivalent power requirement of 3.6 MW would be necessary for this system. This power requirement was balanced against the added power output of the SOFC, and a resulting efficiency (HHV basis) increase of 0.3 % could be realized from the sulfur removal system for the case where 90% of the carbon dioxide was removed prior to introduction of the syngas into the SOFC.

Oxygen Plant Entitlement Results: Various schemes to modify the SOFC system do not result in any significant change in the oxygen concentration of the vitiated air entering the ITM process. Therefore, there is no significant entitlement for the oxygen plant after inserting the ITM in the baseline IGFC configuration.

Combined Cycle Entitlement (Effects of Turbine/Compressor Efficiencies): For the baseline system SOFC/GT island section, compressor efficiencies of 88% and turbine efficiencies of 91% were chosen, based on internal GE data. Other turbomachinery components are likely to be less efficient, because of their smaller size or the way they are operated. The stand-alone expanders were given an 89.3% efficiency, while the recycle blower and other motor driven blowers had a net 80% efficiency. The electrical machinery is usually highly efficient; generators and motors with 99% efficiency are not uncommon. Most of these efficiencies given are above or close to, entitlement. Low-temperature firing, low- pressure to medium-pressure turbomachinery components with a relatively small stage count are not known to achieve these levels of efficiency. Figure 134 and Figure 135 show the sensitivity of the plant efficiency to air compressor and turbine efficiencies respectively. In these plots, all the compressors and turbines in the plant are given the same efficiency, regardless of size. Heavy frame, advanced turbomachinery components are also shown. It could be seen that the sensitivity to turbomachinery components is relatively small. The total gas turbine island entitlement gain is estimated to be 0.3%, albeit at perhaps an increase in cost.

Figure 134 Entitlement Analysis: Effect of Compressor Efficiency on System Efficiency

Figure 135 Entitlement Analysis: Effect of Turbine Efficiency on System Efficiency

SOFC power conversion entitlement: SOFC generators produce DC power, which must to be converted to high voltage, 3-phase AC power before it is connected to the grid. Studies done in the “SOFC Scale-up” project has shown that power conversions efficiencies of 96% are theoretically possible, by eliminating or reducing losses from filtering in the inverter components and by connecting transformers in a novel manner to multiple inverter units. With only switching losses in the inverters, and minimal ohmic losses in the transformers and conductors, this is likely the entitlement in power conditioning. The entitlement gain from SOFC inverter power is assumed to be negligible and is assumed a value of 0%.

Combined Cycle Entitlement (Effects of Fuel Cell Operating Voltage): Fuel cell operating voltage is a measure of fuel cell efficiency; hence, increasing the voltage from the nominal 0.7 V to a higher value should be helpful. However, as the operating voltage increases, other effects are observed:

- (1) Since the fuel cell produces less excess heat with increasing operating voltage, the air flow in the cycle goes down as shown in Figure 136. Although the power produced from the fuel cell system goes up as shown in Figure 137, the gas turbine and steam turbine power decreases.
- (2) With the baseline configuration where an ITM is inserted to provide all the oxygen requirements, restrictions can be imposed on the operating voltage. The O_2 content of the fuel cell exhaust decreases with operating voltage, e.g. as the voltage is increased from 0.694 to 0.773, the O_2 mass % goes from 13.9% to

11.2%. The ITM works with a fixed O_2 mass requirement, so as the voltage goes up, the ITM demands more and more of the exhaust stream. At around 0.775 V, the entire exhaust stream is consumed by the ITM. This provided an upper limit of operation of the fuel cell with ITM under these assumed baseline conditions.

Figure 136 Entitlement Analysis: Effect of Average Cell Voltage on Air Flow to the SOFC in the Baseline System Configuration

Figure 137 Entitlement Analysis: Effect of Average Cell Voltage on SOFC, GT and ST in the Baseline System Configuration

In the baseline system configuration with an ITM inserted, increasing the voltage from 0.7 to 0.77 increases the efficiency by 1.7%. Since the ITM system cannot go beyond 0.775 V, this could be considered the efficiency entitlement of increasing voltage.

Final efficiency improvement: From the above analysis, we can project possible increases in the baseline system results, as follows:

Baseline Efficiency, %		52.8
ITM technology insertion	0.7	
Mid-T Cleanup insertion	0.2	
Subtotal		0.9

Improved System Efficiency, %		53.7

In addition to Tasks 1 and 2, we have determined Upper-Limit Entitlements for the IGFC System:

Improved System Efficiency, %		53.7
Upper-Limit Process Island Entitlement	0.6	
Upper-Limit Combined Cycle Entitlement	0.3	
Upper-Limit Fuel Cell Voltage Entitlement	1.7	
Total Upper-Limit Entitlement		2.6

Final System Efficiency with Upper-Limit Entitlement		56.3

Note that the final realistic system efficiency in the year 2020 time frame is expected to be less than this Upper-Limit Final System Efficiency

4.3.2 Brainstorm

Brainstorm was initiated with the objective of generating ideas that would lead to systems with 60% system efficiency within the technologies' entitlement and feasibility. The problem statement for this brainstorming was as follows:

- Identify technologies which, if developed to their entitlement, can be implemented in an IGFC system and can significantly (>1%) improve the overall system efficiency
- Identify an IGFC system configuration that, if all incorporated technologies are developed to their entitlements', could be shown by analysis to result in a 60% system efficiency

The concepts that resulted from this brainstorming session, summarized below are currently under investigation.

Baseline-2 System Configuration: The basic system configuration is as shown in BL and BL-1. As noted in the entitlement analysis, a system efficiency of 56% is judged possible. By changing the following operational parameters, the system efficiency could be improved further:

- SOFC pressure: preliminary analysis indicated system efficiency gains beyond 12 atm are possible
- SOFC operating temperature: by increasing the operating temperature of the SOFC, the temperature level of the waste heat from the SOFC to the GT and ST cycles would increase, resulting in higher conversion of waste heat to power
- Gasifier methane yield: if the gasifier can be operated to give higher methane yields, the internal reforming capability of the SOFC could result in lower air flows. In simple cycle and gas-fueled SOFC hybrid systems, this usually results in higher system efficiencies as the cooling air to the SOFC can be reduced. However, it is not known if the adopted BGL gasifier could operate to give higher methane yields at this time.

These parameters will be investigated further.

Dartmouth College Study: Figure 138 gives a schematic taken from the Ref. 17reference. Characteristics of this concept include the following:

Gasifier:

Type:	Fluidized bed, two-stage, CO ₂ Acceptor
Operating Pressure	10 Atm
Temperature	800 C (gasifier) 1027 C (regenerator)

Gas Cleanup	S captured primarily in regenerator S polishing (no power used)
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Air Separation	None
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SOFC

Operating Pressure	10 Atm
Cathode Air Temperature In	747 C
Cathode Air Temperature Ex	939 C
Air Temperature Rise	192 C
Oxygen utilization	~0.75
Fuel utilization	0.79
Cell voltage	0.7

Gas Turbine

Efficiency 0.86

Steam Turbines

High Pressure 40 atm

Low Pressure 10 atm

Figure 138 Dartmouth College IGFC Concept

Some unique features of this concept include the elimination of an oxygen air separation unit by using a two-stage gasifier. In addition, the gasifier regenerator stage is partially fueled by the anode exit gas and cathode exit gas. While the efficiency of 63% is significant, analysis as to the amount of air utilization and subsequent air temperature rise in the SOFC is relatively high. Also, the % sulfur capture in the dolomite is assumed to be essentially 100%; it is not known at this time whether this is a realistic assumption. This concept also does not capture CO₂.

2002 Fuel Cell Handbook Concept: Figure 139 gives a schematic taken from reference 18. Characteristics of this concept include the following:

Gasifier:

Type:	Destec entrained bed
Operating Pressure	19 Atm
Temperature	1043 C (gas exit)

Air Separation

Cryogenic

Gas Cleanup

Gas cooling	Radiant heat exchanger
Sulfur	Transport, high temperature
Temperature	593 C
Polishing	Zinc oxide

SOFC, High pressure

Pressure	18 atm
Cathode Air Temperature In	579 C
Air utilization	25%
Fuel utilization	90%
Cell voltage	0.69

SOFC, Low pressure

Pressure	4 atm
Cathode Air Temperature In	645 C
Air utilization	34%
Fuel utilization	90%
Cell voltage	0.69

Gas Turbines

HP pressure	15 atm
LP inlet temperature	979 C
LP pressure	3.3 atm
LP inlet temperature	982 C

Steam Turbines

High Pressure	100 atm
Reheat Pressure	26 atm
Low pressure	6 atm

Some unique features of this concept include the use of two SOFC stages and gas turbines. By expanding the high-pressure gas from the high pressure SOFC and extracting energy, the gas fed to the lower pressure SOFC is cooled. This turbine exhaust acts as the cathode feed to the low pressure SOFC as it has sufficient oxygen at 14 mole-%. However, it is uncertain at this time whether cathode materials can withstand moisture at these levels (~ 7 mole%). Another unique feature is the use of a high-temperature sulfur removal plus zinc-oxide polishing unit. Also, a unique feature to an IGFC system is the use of a recuperator with a large pressure differential between the hot gas (~ 1 atm) and the cold gas (15 atm). The system efficiency is listed in the reference at 59.7% which, for an entrained gasification system, with a cold gas efficiency almost 6percentage points lower than the BGL gasifier, is significant. An inverter conversion (dc to AC power) efficiency of 0.97 is relatively high. This concept also does not capture CO₂.

Gasifier:

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Pressure	15 atm
Cathode Air Temperature In	471 C
Cathode Air Temperature Ex	979 C
Air utilization	43%
Fuel utilization	85%
Cell voltage	0.69

Gas Turbine

Inlet pressure	15 atm
Inlet temperature	754 C

Figure 140 UCI Vision 21 IGFC Concept

Some unique features of this concept include

- The use an advanced transport reactor for the coal gasifier including the use of limestone to capture the majority of the sulfur and the chlorines form the coal

- A warm gas cleanup system to remove the balance of the sulfur and the chlorines
- An ITM for oxygen generation
- A Methanation reactor with an expansion turbine on the syngas prior to feeding the SOFC
- A high-temperature H₂ membrane and shift reactor to recycle anode gas and enable a high overall fuel utilization
- Pressurization of the CO₂ to pipeline pressure for sequestration

The system efficiency given in the reference is only 49.6%. However, the concept includes more than CO₂ isolation; the CO₂ is compressed to 140 bar for insertion into a pipeline and eventual sequestration.

Columbia University IGFC Concept: This concept has several unproven components that would need significant development work to test the feasibility, including a very high temperature SOFC. Figure 141 gives a schematic taken from References 22 and 23. Notable characteristics of this concept include the following:

Gasifier: This gasifier, while not novel in itself as a gasifier, is novel for an IGFC concept. Hydrogasification is a reaction of pulverized coal with hot, high-pressure hydrogen for a short period of time. The hydrogasification reactions are:

The optimal conditions for this process are:

- Temperature range of 1000°F (538°C) to 2000 °F (1076°C)
- Hydrogen partial pressure of 500 to 3000 psi (34 to 204 atm)
- Particle residence time of 0.05 to 10 seconds

Recent studies indicated that the reaction could reach 90% completion at 926°C and 69 atm.

Reformer/CO₂ Acceptor:

Hydrogen makeup for the hydrogasifier occurs in this reformer/CO₂ acceptor reactor. It is not known how well this unit will work as the CaO/CACO₃ is continuously cycled. The CO₂ acceptor reaction is:

Calcliner: Most carbon sequestration techniques involve a reaction of a concentrated carbon dioxide stream with an oxide to form carbonate. Calcium and magnesium oxides are the most studied acceptors of CO₂. A combination of the two such partially calcined dolomite (MgO*CaO) has been shown to be successful.

High temperature SOFC: the operational temperature (1200 °C) and a very high conversion efficiency (70% of reacted H₂ heat) to dc power was assumed. The high operating temperature allows waste heat from the SOFC to be utilized in the calciner which requires a temperature of 1000C.

With the very aggressive assumptions of the SOFC, this IGFC efficiency could approach 60% based on preliminary conceptual modeling.

Figure 141 Columbia University's Hydro-gasification Zero Emission Coal Power Plant Concept

Conventional State-of-the-Art Entrained Bed Gasifiers: Entrained-bed coal gasifiers that are considered ready for use in an integrated gasification combined cycle plant (IGCC) today would include the following:

- GE (formerly ChevronTexaco)
- E-Gas (or Destec, Conoco-Phillips)
- Shell

Studies by DOE (22) have indicated in an IGCC plant, the system efficiencies range in the low- to mid-forty-percent efficiency range (HHV coal to AC power) assuming cold-gas cleanup technology is used. A top-level analysis was performed replacing the BGL gasifier with a GE (Texaco) entrained bed gasifier, assuming similar baseline components downstream of the gasification/cleanup systems are employed. The analysis also assumed the inclusion of ITM technology to generate oxygen. The system efficiency is estimated to be slightly lower than 50%.

In addition to the above concepts, several technologies were identified that, if inserted into an IGFC system could have potential efficiency gains include:

- The use of GE's Unmixed Fuel Processor (UFP) for coal-based H₂ production in a IGFC configuration (19)
- The use of a mid-temperature membrane to separate CO₂ from a H₂ rich stream
- The use of a gasifier designed for higher CH₄ yield, such as a BI-GAS gasifier previously under development by the DOE (20)

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Appendix 1
Table 55 Coal Composition

Coal Characterisitcs	
Coal	Pittsburgh No. 8
Coal Rank	High volatile A bituminous
Mine	Blacksville No. 1
Proximate analysis (as received), wt%	
Volatile matter	36.5
Fixed carbon	46.0
Ash	11.5
Moisture	6.0
Total	100.0
Ultimate analysis (dry), w%	
Carbon	73.2
Hydrogen	4.9
Nitrogen	1.4
Oxygen	4.9
Sulfur	3.3
Ash	12.2
Chlorine	0.1
Total	100.0
Hardgrove grindability	57
Higher heating value (as received), Btu/lb	12350
Lower heating value (as received), Btu/lb	11851
Free swelling index	7.5
Size consist as received, wt%	
Between 2" and 1/4"	45
Less than 1/4"	55
Ash Fusion temperatures, reducing/oxidizing, deg F	
Initial deformation	2018/2319
Softening	2107/2340
Hemisphere	2179/2426
Fluid	2261/2518

Table 56 Coal Ash Composition

Coal Ash Composition and Sulfur Forms	
Major ash constituents, wt%	
SiO ₂	45.05
Al ₂ O ₃	18.45
TiO ₂	0.80
Fe ₂ O ₃	20.31
MgO	0.98
CaO	6.09
Na ₂ O	0.98 (soluble 0.22)
K ₂ O	1.55 (soluble 0.003)
P ₂ O ₅	0.42
SO ₃	5.25
Undetermined	0.12
TOTAL ASH	100.00
Coal sulfur forms, wt%	
Sulfate	0.06
Organic	1.60
Inorganic	1.64
TOTAL SULFUR	3.30

Table 57 Limestone Composition

Limestone Composition, wt%	
CaCO ₃	90.52
MgCO ₃	2.68
SiO ₂	5.08
Inert	1.72
TOTAL	100.00

Table 58 Bitumen Composition

Bitumen Composition, wt%	
Carbon	85.00
Hydrogen	9.70
Nitrogen	1.20
Oxygen	0.70
Sulfur	3.40
Chlorine	0.00
TOTAL	100.00