DOE/MC/28065--1
RECEIVED
AUG 11 1998
OSTI

Molten Carbonate Fuel Cell Product Development Test

Final Report
September 30, 1992 - March 31, 1997

Work Performed Under Contract No.: DE-FC21-92MC28065

For
U.S. Department of Energy
Office of Fossil Energy
Federal Energy Technology Center
P.O. Box 880
Morgantown, West Virginia 26507-0880

By
M-C Power Corporation
8040 South Madison Street
Burr Ridge, Illinois 60521

DISTRIBUTION OF THIS DOCUMENT IS UNLIMITED

MASTER

Disclaimer

This report was prepared as an account of work sponsored by an agency of the United States Government. Neither the United States Government nor any agency thereof, nor any of their employees, makes any warranty, express or implied, or assumes any legal liability or responsibility for the accuracy, completeness, or usefulness of any information, apparatus, product, or process disclosed, or represents that its use would not infringe privately owned rights. Reference herein to any specific commercial product, process, or service by trade name, trademark, manufacturer, or otherwise does not necessarily constitute or imply its endorsement, recommendation, or favoring by the United States Government or any agency thereof. The views and opinions of authors expressed herein do not necessarily state or reflect those of the United States Government or any agency thereof.

DISCLAIMER

**Portions of this document may be illegible
electronic image products. Images are
produced from the best available original
document.**

TABLE OF CONTENTS

1.0 EXECUTIVE SUMMARY	1
2.0 INTRODUCTION.....	3
3.0 TECHNICAL RESULTS	4
Task 1 Project Management and Permitting	4
Task 2 250 kW MCFC Demonstration Design	4
Task 3 250 kW MCFC Stack Manufacturing and Assembly	8
Task 4 BOP Fabrication	29
Task 5 Plant Site Activities	30
Task 6 Demonstration Plant Testing	61
Task 7 ARPA Enhancements	90
4.0 CONCLUSIONS AND RECOMMENDATIONS	90

List of Tables

Page

Table 1	Summary of Full-Area Mock-Ups	13
Table 2	PDT-1 Initial Stack Height Measurements	16
Table 3	Summary of Leak Tests	27
Table 4	ATF Cell Performance Summary	29
Table 5	Precommissioning Tests/Checkouts	41
Table 6	Reducing Safegas Compositions and Estimated Quantities	47
Table 7	Regression of Cathode Pressure Drop Data for PDT-1 Stack at NAS Miramar	70
Table 8	Example of Estimating Anode Inlet Flow and Composition from Reformer Operation	76
Table 9	Example of Calculating Reformer Product Flow Rate and Composition	78
Table 10	Example of PDT-1 Stack Material Balance for NAS Miramar Operation	79
Table 11	Summary of Material Balance for PDT-1 Stack at NAS Miramar	82
Table 12	Overall Mass Balance for NAS Miramar Plant at 1200 Amps Operation	85
Table 13	Summary of Overall NAS Miramar Plant Energy Balance during 1200 Amps Operation	87

List of Drawings

Drawing B-101	Process Flow Diagram	32
Drawing B-102	Process Flow Diagram Balance of Plant (Instrument Air and Inert Gas Systems)	37

List of Figures

Page

Figure ES-1	2
Figure ES-2	2
Figure 1 Schematic of the MCFC Stack	6
Figure 2 Picture of the Acceptance Test Facility	7
Figure 3 Schematic of Separator Plate	9
Figure 4 Assembly Area Layout	15
Figure 5 ATF Voltage Lead Routing	18
Figure 6 ATF Enclosure Furnace	19
Figure 7 Open Circuit Voltage	21
Figure 8 Stack Voltage During Performance Testing	22
Figure 9 Current Density During Performance Testing	22
Figure 10 Power Output During Performance Testing	23
Figure 11 Energy Produced During Performance Testing	23
Figure 12 Start-Up Block Flow Diagram	44
Figure 13 History of Stack Temperature at NAS Miramar	63
Figure 14 History of Stack DC Current at NAS Miramar	63
Figure 15 History of Stack DC Voltage at NAS Miramar	64
Figure 16 History of Stack DC Power at NAS Miramar	64
Figure 17 Polarization Curve for PDT-1 Stack at NAS Miramar	65
Figure 18 Voltage Distributions for PDT-1 Stack at Miramar	65
Figure 19 History of Cathode Temperatures at NAS Miramar	66
Figure 20 History of Cathode Pressure Drop for PDT-1 Stack at NAS Miramar	68
Figure 21 Cathode Pressure Drop Correlation	68
Figure 22 Cathode Pressure Drop vs. Flow for PDT-1 Stack at Miramar	69
Figure 23 Cathode Pressure Drop vs. Time for PDT-1 Stack at NAS Miramar	71
Figure 24 PDT-1 Stack Design vs. Bench-Scale Cell Voltages	72
Figure 25 Predicted vs. Actual Average PDT-1 Stack Cell Voltages at NAS Miramar	74
Figure 26 Predicted vs. Actual Average PDT-1 Stack Cell Voltages at ATF	74
Figure 27 Schematic for PDT-1 Stack Material Balance at NAS Miramar	75
Figure 28 History of Material Balances for PDT-1 Stack at NAS Miramar	83
Figure 29 NAS Miramar Plant Material Balance	84

1.0 EXECUTIVE SUMMARY

This report summarizes the work performed by M-C Power Corporation, under the U.S. Department of Energy (DOE) Cooperative Agreement No. DE-FC21-92MC28065, for manufacturing and demonstrating the performance of its 250-kW molten carbonate fuel cell (MCFC) stack in an integrated system at the Naval Air Station Miramar (NAS Miramar) located in San Diego, California. The program was jointly funded by DOE and the Department of Defense (through the Advanced Projects Research Agency). M-C Power's cost sharing obligation was partly funded through a Gas Research Institute (GRI) contract, and subcontract cost sharing from San Diego Gas & Electric (SDG&E) and Bechtel Corporation.

The stack constructed for the demonstration test at the NAS Miramar consisted of 250 cells. It was manufactured using M-C Power's patented Internally Manifolder Heat Exchanger (IMHEX®) stack design. Each cell in this stack has an active area of 10,800 cm², the maximum active area per cell ever operated in MCFC stacks.

The demonstration test at NAS Miramar was designed to operate the 250-kW MCFC stack in a cogeneration mode. This test represented the first attempt to thermally integrate an MCFC stack in a cogeneration system. The test was started on January 10, 1997, and voluntarily terminated on May 12, 1997, after 2,350 hours of operation at temperatures above 1,100°F and at a pressure of three atmospheres. It produced 160 MWh of d.c. power (Figure ES-1) and 346,000 lbs of 110 psig steam for export during 1,566 hours of on-load operations. The test demonstrated a d.c. power output of 206 kW.

The test also demonstrated excellent uniformity of voltage among the 250 cells. The average cell voltage for the test was 720 mV/cell (Figure ES-2) and the standard deviation of the cell voltages was about 20 mV. The uniformity of voltages for the top and bottom cells indicates the absence of undesirable carbonate pumping in the system. The fuel cell stack also demonstrated its ability to transition smoothly from an on-load condition to standby mode and vice versa. The test generated good quality data, with energy and material balances of 95% and 96%, respectively.

Most of the balance of the plant (BOP) equipment operated satisfactorily. However, the off-the-shelf automotive turbocharger used for supplying air to the plant failed on numerous occasions and the hot gas blower developed seal leakage problems which impacted continuous plant operations.

Overall the demonstration test at NAS Miramar was successful in demonstrating many critical features of the IMHEX® technology. Lessons learned from this test will be very useful for improving designs and operations for future MCFC power plants.

Figure ES-1

250-kW Stack at NAS Miramar
Cumulative Stack DC Power History

Figure ES-2

NAS Miramar 250 kW Stack Performance
Conditions: 1140 ADC, 206 kW

2.0 INTRODUCTION

This report summarizes the work performed by M-C Power Corporation, under the U.S. Department of Energy (DOE) Cooperative Agreement No. DE-FC21-92MC28065, for manufacturing and demonstrating the performance of its 250-kW molten carbonate fuel cell (MCFC) stack in an integrated system at the Naval Air Station Miramar (NAS Miramar) located in San Diego, California. The program was jointly funded by DOE and the Department of Defense (through the Advanced Projects Research Agency). M-C Power's cost sharing obligation was partly funded through a Gas Research Institute (GRI) contract, and subcontract cost sharing from San Diego Gas & Electric (SDG&E) and Bechtel Corporation.

M-C Power manufactured the MCFC stack and managed the project. Bechtel was responsible for the design, detailed engineering, construction management and plant start-up. Stewart & Stevenson Services, Inc. (S&S) fabricated the balance of plant (BOP) in their shop on a skid and tested the performance prior to shipping it to the site. SDG&E performed site construction work with its own staff which was supplemented by craft labor under subcontracts with them. M-C Power operated the plant with its own funding and utility support. The plant was operated 2½ months beyond the March 31, 1997 Cooperative Agreement end date.

The fuel cell demonstration project site was originally planned to be at the Kaiser Permanente Medical Center in San Diego, California. Due to stringent siting requirements, and escalating site construction cost estimates, the Kaiser Medical Center location was deemed prohibitive. M-C Power and SDG&E relocated the fuel cell demonstration project to an alternate site at the Naval Air Station Miramar.

The MCFC demonstration plant operated at NAS Miramar represents the first truly integrated MCFC power plant in a cogeneration application. The power plant was fueled by natural gas and produced electricity to the SDG&E electrical distribution grid. It also supplied thermal energy in the form of saturated steam to the Naval base. Miramar was chosen as the site for the product development test (PDT) to gain operating experience with an MCFC power plant in an application which is similar to a commercial environment.

GRI sponsored the balance of plant (BOP) design and procurement efforts. SDG&E and The Electric Power Research Institute (EPRI) participated in site related and operational activities. The roles of GRI, EPRI and SDG&E were particularly important as their emphasis was on addressing design issues which would impact acceptance by the gas and electric utility industries.

The project was divided into the following tasks:

1. Project Management and Permitting
2. 250 kW MCFC Demonstration Design
3. 250 kW Stack Manufacturing and Assembly

4. BOP Fabrication
5. Plant Site Activities
6. Demonstration Plant Testing
7. Site Revision Modifications

This report presents the work conducted under each of the above tasks.

3.0 TECHNICAL RESULTS

Task 1 Project Management and Permitting

Objective: Development of a project management system for planning, budgeting, executing, and controlling the costs and schedules associated with the work. To obtain permits required for stack and BOP fabrication/testing and site related activities.

Subtask 1.1 Management Plan

Discussion: This subtask was completed on November 11, 1992 when the management plan was submitted to DOE.

Subtask 1.2 Environmental Information

Discussion: This subtask was completed on September 29, 1995 when a Topical Report on the subject was delivered to DOE. This environmental document, containing the information necessary to allow DOE to prepare documentation to comply with National Environmental Protection Act (NEPA) requirements, was prepared to reflect the new project site.

Subtask 1.3 Local Permits

Discussion: This subtask was completed on February 9, 1995 when all local permits and regulatory approvals were obtained.

Subtask 1.4 Management and Reporting

The management plan, budget, and schedule were revised regularly to reflect changes in the project. M-C Power prepared and submitted to DOE monthly and topical reports which provided general project updates and in-depth technical reports on selected topics.

Task 2 250 kW MCFC Demonstration Design

Objective: Prepare designs for the stack, balance of plant, and site that are integrated and consistent with the 250 kW product development test (PDT) demonstration plant performance objectives.

Subtask 2.1 Stack Design

Discussion: The original plan was for the stack design to be a duplicate of that used in the Unocal demonstration performed under contract No. DE-AC21-90MC27394. However, based upon experience gained on the Unocal project, a number of design changes were included in the PDT stack. The changes relate to a more simplified clamping device, thinner electrodes, new matrix formulation and an improved separator plate design. A number of trade-off studies were performed regarding stack sealing, box plenum design and the pressure vessel design. Specifications were developed for all of the key components and parameters affecting the overall stack. A schematic of the stack is shown in Figure 1.

Intermediate Plates

Based upon experience gained at the Unocal demonstration and stack component fit-up studies in M-C Power's factory, a decision was made to incorporate intermediate plates in the PDT stack.

The PDT intermediate plates were designed to be a barrier to the stack-up of tolerances of the cell assemblies. The Unocal stack exhibited several gaps which were a direct result of the stack-up of bi-polar plate and active area component thicknesses. The thickest parts of the components overlap on the two non-manifold sides of the stack while thinnest parts all fall in the middle of the active area. In an effort to alleviate the stack-up of tolerances, nine intermediate plates were used in the stack. The benefits of intermediate plates were proven in several mechanical verification tests. The number and placement of these intermediate plates was determined by a structural analysis model of the 250-cell stack that was developed specifically for the M-C Power 250 kW MCFC stack.

Guard Heaters

Design changes were dictated by the Unocal demonstration experience. A design for the anode bus support in the pressure vessel that comes off the guard heater frame was completed and incorporated in the final guard heater design.

Acceptance Test Facility (ATF)

M-C Power completed several activities in the Acceptance Test Facility (ATF) shown in Figure 2. The ATF is used to support the conditioning and testing of 250 kW stacks prior to leaving the factory. Testing in the ATF took place over the period May 23, 1996 through July 3, 1996. During conditioning, the stack height shrinks due to binder removal. Stack shrinkage measurement equipment was installed and used during the conditioning and testing of the stack to ensure that stack tilting did not take place.

Figure 1 Schematic of the MCFC Stack

Figure 2 Picture of the Acceptance Test Facility

Subtask 2.2 BOP Design

Discussion: The NAS Miramar balance of plant design represents a significant improvement over the Unocal BOP design. The PDT plant design is more compact covering approximately 40% of the Unocal demonstration site area. Bechtel performed the design work associated with the specification of BOP equipment and prepared progress and instrument diagrams for assembly. Stewart & Stevenson assembled the BOP equipment on a skid that was truck transported to NAS Miramar in February, 1996. S&S personnel were on site to assist with the mechanical and electrical interconnections of the BOP equipment skid to the other power plant equipment and to verify installation in accordance with design specifications. They also played a key role in design reviews related to the recycle gas blower shaft seal leakage problem. A final BOP Design report was submitted on October 23, 1996.

Subtask 2.3 Site Interface Design

This task was completed and a topical report on the subject was submitted to DOE on October 4, 1995. The site interface design was used as the basis for preparation of bid packages for field construction work.

Task 3 250 kW MCFC Stack Manufacturing and Assembly

Objective: Fabricate stack repeat and non-repeat components, assemble and test a 250 kW stack.

Subtask 3.1 Component Manufacturing

Discussion: Active components were manufactured at M-C Power. Essentially all of the non-repeat components were manufactured by vendors who were pre-screened for competence prior to receiving a bid solicitation.

Separator Plate

A new plate design and manufacturing procedure was developed for improved sealing compared to the stack fabricated for the Unocal demonstration was developed and used. A new automated aluminization procedure was developed. The new procedure was developed to reduce the laborious task of manually masking the separator plates before aluminization. A schematic of the separator plate is shown in Figure 3.

Intermediate Plate Manufacturing

The manufacturing of the intermediate plates consisted of several steps. After the base material was received it was straightened to meet the flatness and parallelism specifications. The manifold holes were then cut out of the base plates using a water jet and half cells were welded to the base plates.

Figure 3 Schematic of Separator Plate

"PROCESS GAS"	"MANIFOLD"
"ANODE INLET"	A1, A2, A3
"ANODE OUTLET"	A4, A5
"CATHODE INLET"	C1, C2, C3
"CATHODE OUTLET"	C4, C5

"CORNER MANIFOLDS"	A1, C1, A3, C-3
"DOCHOUSE MANIFOLDS"	A2, C2, A4, C4, A5, C5

ALUMINIZE THIS
"SHADED" SURFACE AREA
"WET SEAL AREA"

Water Jet Cutting Process

The manifold holes were cut out of the base plates using a water jet process. This eliminated base plate warpage which would have been experienced if an alternative method such as laser cutting or wire EM had been used. All of the base plates had to be positioned three times during the water jet cutting process. The initial positioning allowed for cutting of two corner manifolds and two doghouse manifolds on one side of the plate. The second position allowed for cutting of two corner manifolds and two of the doghouse manifolds on the other side of the plate. The final position allowed for cutting of the center doghouse manifolds. The first three plates which were cut using the water-jet process differed from the cut out shims by 1/16 inch. This was due to a slight variance in the cutting programs and was corrected for the remaining seven base plates.

Half Cells

The half cells were welded to the base plates using the bi-polar plate welding fixture. The welding parameters were based on numerous welding trials, which were performed on various sample half cells and stainless steel plate samples. All of the welding trial samples were inspected and approved before the welding of half cells to base plates began. The procedure used for welding half cells to the base plates was similar to that used for welding the shims to the base plates. The half cells were clamped and tack welded in place prior to edge welding. Inspection of the welded half cells indicated an acceptable fusion weld. However, during leak testing it was discovered that there were several areas with no metal-to-metal contact between the half cell and shim plate. This occurred most often in the manifold areas at the corner radii and along the outside perimeter. All of the areas with leaks were rewelded.

Intermediate Plate Measurements

Eight measurements were taken on the intermediate plates to insure compliance with specifications. These were dimensional compliance, base plate flatness and parallelism, wet seal flatness, modified total wet seal thickness, total wet seal thickness, step height, dimple amplitude, and rib amplitude.

Electrodes

The general specifications for the PDT anodes and cathodes define the raw materials, processing conditions, and post-manufactured specifications utilized for manufacturing the stack electrodes. The major improvements and changes incorporated into the PDT stack are discussed below, although many of the specifications remained virtually unchanged from previous commercial-size and subscale stack tests.

Decreased Component Thicknesses

Decreasing the thickness specification of the PDT electrodes was made in response to the need to utilize the inventoried current collectors which had been specified prior to the change to a negative fit up. Manufacturing, casting and sintering trials were completed prior to full-area production to demonstrate the manufacturing ability within specification.

As a basis for lowering the component thickness specifications, thinner components were tested in a full-area 20 kW stack and in a 100-cm² cell test to demonstrate the baseline performance of the thinner electrodes in the cell package. The cell test operated for 1850 hours under systems conditions pressurized to 3 atm with a performance of 776 mV at 160 mA/cm² and 60% $U_{t_{fuel}}$. The cell showed steady performance for the first 1600 hours of operation followed by a rapid decrease in the performance. Post-test analysis determined that the cause of the performance loss was accelerated corrosion of the flame-sprayed aluminized anode wet seal.

Milled Chromium Powder

Analysis of the degree of alloy formation in the Unocal demonstration anodes indicated that incomplete dissolution of the chromium into the nickel particles occurred. Depending upon the belt speed and high heat temperature in the sintering furnace, the percentage of chromium in nickel varied between 1.2% and 9.5% for the Unocal anodes. The remainder of the chromium remained exterior to the nickel particles as large chunks of chromium which do little to strengthen the anode and consume more of lithium carbonate than a properly alloyed anode.

In addition, the poor dispersion of the chromium powder resulted in areas of the nickel which did not have chromium to inhibit sintering. As a result, the anode porosity after sintering dropped from 60% to less than 50% for the Unocal anodes. In response, the sintering temperature was lowered and the belt speed increased to meet the porosity specification of 60%.

A particle size analysis of the starting chromium powder indicated that 50% of the as-received chromium powder particles exceeded the size of the starting nickel particles. The size of the chromium powder yielded a poor distribution of the chromium resulting in areas of nickel which are not stabilized against sintering.

Prior to manufacturing the anodes, a procedure was developed for reducing the average chromium particle size. By lowering the average chromium particle size prior to casting and sintering, an improved dispersion of the chromium resulted in higher porosities at the temperatures required for complete dissolution of the chromium into nickel. Although currently there is not a room temperature method for analyzing the strength of anodes, based on die cutting experience, the PDT anodes are much stronger than the Unocal anodes used in the Unocal demonstration project.

Anode Composition: Ni+6% Cr

Prior to the PDT stack, all of M-C Power's stack anodes were Ni-10% Cr, which is double the chromium content of the standard IGT anodes. The primary difference between the MCP Ni-10% Cr anodes and the IGT Ni-5%Cr was the particle size of the chromium powder. As discussed above, the chromium powder particle size was lowered prior to manufacturing the PDT anodes to sizes compatible with decreasing the chromium levels to 5 weight percent. The smaller particle size chromium aids in dispersing the chromium prior to heat treating to obtain a uniform alloy content throughout the nickel particles.

After manufacturing the PDT anodes, analysis of the degree of Cr alloying was completed and the uniformity of the dispersion verified.

Die Cutting Electrodes

In order to reduce costs and cycle time and to improve quality, the process of cutting the electrode plan form was successfully changed from laser cutting to die cutting/stamping. The PDT-1 stack was the first stack to use die cut electrodes. A total cost savings of \$5,768 (\$9.61/ea.) was realized in changing from laser cutting to die cutting. It took approximately 10 working days to pre-trim, package, die cut and ship the PDT electrodes compared with 15 working days to package, laser cut and ship Unocal electrodes. The post cutting yields were 94% for anodes and 96% for cathodes.

Matrix

The matrix manufacturing procedure remained virtually unchanged from that utilized for the Unocal stack, except that the total thickness was decreased. The total matrix thickness now used is the IGT standard, and matrices with this thickness were tested in the full-area Stack MCP-7 and in the 100-cm² cell test M0195001. Carbonate loadings were established based on electrode and matrix thicknesses.

Subtask 3.2 Stack Assembly

Discussion: Stack assembly was completed on April 26, 1996. As a result of the lessons learned from the Unocal demonstration stack this PDT stack included intermediate plates.

Final Assembly Verification Testing

A mock-up is a stack model assembled with actual stack metal components (separator plates, electrodes, current collectors) and actual or surrogate electrolyte matrix packages.

The main purpose of the mock-ups was to determine contact pressure distribution on wet seals and on the active area. Sealability was determined along with internal resistance. In parallel, a cell package mechanical analysis was conducted. Matrix, carbonate, electrode, and plate compressive strength characteristics were determined as a function of pressure, temperature, and package layers using Instron testing equipment. An engineering consultant hired by M-C Power integrated basic materials properties into a first approach cell model. The model has been used to assess, retrospectively, the two dimensional stacking of multilayer (tall) stacks - force distribution and component thickness pattern/tolerance effects.

Eight mechanical verification tests (mock-ups #1 thru #8) were conducted to help insure the successful operation of the stack which incorporated the changes made. Table 1 summarizes the mock-up components and conclusions from testing.

Table 1 Summary of Full-Area Mock-Ups

Mock-up No.	Number of cells	Components	Conclusions
1	250	<ul style="list-style-type: none"> * Bipolar plates * Current collectors * Electrodes * Simulated matrices * No intermediate plates 	<ul style="list-style-type: none"> * Unacceptable tolerance accumulation
2	50	<ul style="list-style-type: none"> * Bipolar plates * Current collectors * Electrodes * Matrices * No intermediate plates 	<ul style="list-style-type: none"> * Intermediate plates are desirable within 50 cells
3	250	<ul style="list-style-type: none"> * Bipolar plates * Current collectors * Electrodes * Simulated matrices * Five simulated intermediate plates 	<ul style="list-style-type: none"> * External sealing improved * Internal sealing inadequate
4	125	<ul style="list-style-type: none"> * Bipolar plates * Current collectors * Electrodes * Simulated matrices * Two intermediate plates 	<ul style="list-style-type: none"> * External sealing improved * Internal sealing inadequate
5, 6, 7	125	<ul style="list-style-type: none"> * Bipolar plates * Current collectors * Electrodes * Simulated matrices * Two intermediate plates * Various internal sealing techniques 	<ul style="list-style-type: none"> * External sealing within specification * Internal sealing improvement demonstrated
8	250	<ul style="list-style-type: none"> * Bipolar plates * Current collectors * Electrodes * Matrices * Nine intermediate plates * Internal seals 	<ul style="list-style-type: none"> * External sealing verified * Internal sealing verified

Methods to Measure Contact Pressure

Initially, to evaluate contact pressure distribution between the components in a mock-up, only pressure measuring film was used. A two-sheet type film, comprised of an A-film, featuring a layer of micro encapsulated color-forming material, and a C-film, featuring a layer of color-developing material. The pressure grade of the film was chosen for the range of 29 psi to 87 psi. When pressure was applied, the micro capsules on the A-film burst, and the clear color-forming material is released and absorbed by the color-developing material of the C-film. The micro capsules of color-forming material were adjusted to break at different pressure levels, thus resulting in color densities, which varied by the amount of pressure used. The film only indicated the maximum pressure obtained.

The film was a useful tool at the beginning of these studies and provided quick practical information on sealing pressures. However, the film had a number of disadvantages such as high cost, a limited range and accuracy of measured pressure, the ability to "memorize" only the maximum local pressure, and the inability to provide real-time information. An alternative procedure was evaluated and implemented.

The alternative system is an advanced pressure distribution measurement system free of the disadvantages described above. Using paper-thin reusable sensors, the system can obtain pressure profiles and present the information for real time color-coded display. A record of this information is stored for later reference and comparison. The heart of the system is a high resolution, matrix based tactile sensor capable of measuring pressures from 0.3 iwc to > 1000 psi.

The sensors are characterized by a grid of rows and columns. Each intersection of a row and a column forms an individual sensing element. The system's pressure resolution is 255 levels (8 bits) over the sensor's operating range. The system electronics scans the sensor matrix yielding a resistance measurement at each intersection of a row and a column. The application of a mechanical force to a sensing element results in a monotonic change in measured resistance. The software uses a calibration algorithm to relate the digital output to the applied pressure. The software then displays this information in two and three dimensional colored displays. This thin film strain gauge system was used for all stack mock-ups starting with Mock-up # 5.

Conclusions From Mock-Up Tests

1. Based on the mock-up results, stacking flatness, and sealing tests, the PDT stack met the assembly acceptance criteria and was determined to be ready for conditioning in the ATF. The stack showed considerable improvement over the Unocal demonstration stack in terms of flatness, shape, and sealing.
2. The stack model was quantitatively verified by the mock-up results with regard to load deflection and qualitatively with regard to load distribution. Absolute values of load distribution can be recalculated by taking into account the actual local component thicknesses.

3. Negative fit-up, the intermediate plates, and component shape improvement proved to be beneficial.
4. The strain gauge system proved to be a valuable tool for the stack sealing analysis.

Assembly Plan/Layout

Final assembly included stacking the electrolyte/matrix packages on top of the already subassembled (electrodes and current collectors installed) bi-polar plates, installation of the clamping device, doing pre-test clamping pressure adjustments, taking pre-test measurements and performing a leak test. Stack final assembly occurred in a humidity controlled dry room.. Component sorting and stacking were performed concurrently. The layout of the assembly area is shown in Figure 4.

During stack assembly, all of the component identification numbers which were visible were recorded. This was done to confirm use of proper cell packages during assembly. Both the total matrix thickness and carbonate provided to each cell were tracked during assembly to verify that correct cell compositions were used. After assembly, the cell packages were matched back to bi-polar plates and cathodes using the anode identification numbers.

Figure 4 Assembly Area Layout

Stack Height

The initial stack height with a clamping force of 40 psi (on the top cell) was measured at six locations around the stack. Three measurements were taken on the anode inlet side, one each at A1, A2 and A3. Three measurements were taken on the cathode inlet side, one each at C1, C2, and C3. Table 2 includes all of the stack height measurements. The average stack height was 2275 mm. The average stack height measured on the anode inlet side was 0.3 mm higher than the average cathode inlet side stack height. In comparison, this difference was 13.3 mm on the Unocal demonstration stack.

Table 2 PDT-1 Initial Stack Height Measurements

Measurement Location	A1	A2	A3	C1	C2	C3
Stack Height (mm)	2277	2275	2274	2276	2275	2274

Clamping Device/ Clamping Pressure

The clamping device consisted of a load plate, six bellows, a top clamp plate and four tie rods. Once in the ATF, the tie rods are connected to four screw jacks and a 480 volt a.c. motor. During pre-conditioning, the bellows were maintained at a "safe" height which was within the manufacturer's limits of compression and extension. Mechanical stops were used to prevent the bellows from going beyond these limits. The bellows' height was monitored by four linear variable differential transducers (LVDTs). Limits for expansion and compression of the bellows which were within the "safe" range were programmed into the control system. When the bellows would reach the limit for expansion, the motor would engage and the screw jacks would pull the top clamp plate down using the tie rods. When the bellows height reaches the limit for compression, the motor engages in the reverse direction and raises the top clamp plate using the tie rods.

The six bellows were divided into two groups of three. Each group of three bellows had independent controls. The three bellows on the anode inlet side of the stack were in bellows bank "A" and the three bellows on the cathode inlet side of the stack were in bellows bank "B". This provided a tool to compensate for stack tilt during conditioning. The gas pressure in both banks was made equal for all of stack conditioning.

After stack assembly, the clamping pressure was increased and then released back to dead weight only. The clamping pressure was then slowly reapplied until the final stack clamping pressure was achieved. The application and subsequent release of the clamping pressure was done to complete initial plate flattening, plate deflection and matrix compaction to aid in the stack sealing when the clamping pressure was reapplied.

Stack Instrumentation

Based upon instrumentation failures observed in the Unocal demonstration stack assembly, all instrument specifications, hardware designs, and construction materials, were reviewed before assembly of the stack. Based upon this review, there were significant changes in instrumentation fittings and thermocouple assembly specifications.

New construction materials utilized in revised, or new instrument designs, were subjected to simulated fuel cell environmental testing prior to use in the stack assembly. Instrument hardware located inside the enclosure furnace of the ATF was also redesigned to prevent electrical shorting and carbonate losses during stack conditioning. These efforts were successful as demonstrated by the absence of stack instrument failures during the conditioning and acceptance testing of the stack assembly.

Temperature Instrumentation

The thermometry installed in the PDT stack was similar to the 250 cell stack shipped to the Unocal demonstration test site although greatly reduced in number.

Three thermocouples each were installed in middle locations of the cathode active area gas rib field of various cells. An additional interior cell thermocouple was installed in a middle location of cell five on the anode active area gas rib field. Thermocouples were also installed in select stack manifolds at the anode and cathode end plates, inside deep thermocouple wells in both end plates, and at the bellows location at the top of the stack assembly.

All internal cell thermocouples except for that installed in cell five were cut upon removal of the stack assembly from the ATF after conditioning.

Data sheets detailing the temperature data from the internal cell and gas manifold thermocouples during stack conditioning and operation were prepared. These data sheets were used to provide a tool for the estimation of internal stack temperature based upon the manifold temperature readings. This data was used to estimate internal stack temperature during the initial start up of the stack at the NAS Miramar site.

Pressure Instrumentation

Stack gas manifold differential pressures were monitored manually from selected sensing ports on both the anode and cathode end plates. The anode end plate had pressure sensing ports for each gas manifold. The cathode end plate had pressure sensing ports installed in one fuel inlet manifold (A2) and both fuel outlet manifolds (A4 and A5).

Voltage Instrumentation

Numerous voltage leads were installed onto the stack assembly for monitoring stack voltage. During ATF stack operation the stack voltage of cells with voltage leads were measured along with a cell package of 25 cells and eleven packages of 20 cells. A total of 19 stack voltages were

monitored to provide data as to the operating performance of the stack. Figure 5 shows the routing of the voltage leads in the ATF.

Figure 5 ATF Voltage Lead Routing

Subtask 3.3 Conditioning, Pretesting, and Shipping

Discussion: The 250 kW stack assembly was placed in the ATF and all of the appropriate connections were made. The stack test was started on May 24, 1996 and was completed on July 3, 1996. Testing in the ATF yielded the best results M-C Power had obtained to date for a 250 kW stack.

Conditioning

Conditioning of the 250 kW stack included binder removal, electrolyte melting, cathode oxidation, and heat-up to operating temperature. Conditioning of the stack began May 24, 1996 and lasted a total of 593 hours.

Binder Removal

The purpose of binder removal from a new molten carbonate fuel cell stack is to prepare the stack for carbonate melting and stack operation. As the temperature of the fuel cell stack is elevated above ambient, the organic binder used in active component manufacturing begins to decompose. The decomposing binder products are carried from the stack in the carrier gas. The flow rate of the carrier gas must be above the required threshold value to transport the gaseous products effectively. The gas flows and heat up schedules for the stack were based on previous stack testing experience and the results of binder removal studies performed in several bench scale cells.

During this phase of conditioning, the binder removal system was utilized. The system included two carbon beds in series on each side which were located directly after the coolers. Hydrocarbon analyzers were used to monitor the hydrocarbon levels at the inlet of the first carbon bed, the outlet of the first carbon bed and the outlet of the second carbon bed. In addition, the pressure drop across each carbon bed was monitored.

Electrolyte Melting

Before a fuel cell stack is operable, the electrolyte must be melted and distributed among the active cell components. For a 250 kW stack, the technique of controlled electrolyte melting is a crucial issue. Non-uniformity of melting could result in the development of a tilt which could in turn lead to shifting of the cells. In addition, the carbonate around the manifolds melts before the remainder of the active area resulting in a possible loss of sealing. If air enters the stack at this time, the nickel electrodes oxidize. Since nickel oxidation is an exothermic reaction, this would create a considerable amount of heat which would then cause a loss of control over the electrolyte melting process. If the anode electrodes oxidize, they will no longer be viable for fuel cell operation.

The majority of electrolyte melting for the stack was controlled using only the top three heating zones of the enclosure furnace, zones 6, 7 and 8. Figure 6 illustrates the enclosure furnace zone configuration. Initially, inlet gas temperatures and the bottom furnace enclosure zones were kept below the melting point and the top three enclosure zones were used to provide the heat for melting. As melting progressed from the top of the stack downward, the lower enclosure furnace zones were then used to help complete melting. Throughout the majority of the electrolyte melting process, the anode and cathode inlet gas temperatures remained below the melting point. Anode and cathode inlet gas flows were also kept equal to prevent uneven heat transfer.

Figure 6 ATF Enclosure Furnace

Throughout melting, the stack tilt was monitored with four Linear Variable Displacement Transmitters (LVDTs). The goal was to maintain the stack as close to level as possible. The tools used were: the enclosure purge gas flows, and the temperature differential between the feed gas on the cathode inlet and anode inlet sides.

Nitrogen was used to purge the enclosure furnace. Previous stack testing indicated that the stack temperatures near the purge gas inlet were lower than the rest of the stack. Additional purge lines were run into the enclosure to try to decrease the temperature differentials within the enclosure furnace. Purge lines were located near the C1 corner of the stack, above the cathode inlet pipe, above the cathode outlet pipe and below the anode outlet pipe. As the stack tilt began to increase, the purge gas flow rates would be reduced near the highest corner of the stack and increased near the lowest corner of the stack.

The tilt from the anode inlet side to the cathode inlet side could also be affected by the inlet gas temperatures. The differential between the inlet gas temperatures was increased or decreased as needed based on the stack tilt.

Internal stack temperatures were monitored during electrolyte melting to ensure a controlled melt. Internal manifold pressures were also closely monitored during melting to ensure no loss of external sealing. During melting, manifold pressures remained positive indicating that sealing was maintained throughout conditioning.

Cathode Oxidation

The cathodes of a new molten carbonate fuel cell stack are manufactured from porous nickel. To provide the required function in the individual cell packages, lithiated nickel oxide will need to be formed. This is achieved by small additions of air to the oxidant at an elevated temperature. Because of the extremely exothermic nature of nickel oxidation, the introduction of air to the oxidant gas mixture must be carried out very slowly and in a controlled manner. The process of oxidation is monitored by internal stack temperatures, cell group voltages and gas chromatographic analysis. During cathode oxidation, care must be taken to prevent anode oxidation due to gas crossover.

Cathode oxidation of the stack was initiated at a lower temperature than previous stacks at M-C Power. There were two reasons for initiating cathode oxidation early. The lower stack temperature would result in a smaller exotherm making it easier to control stack temperatures. Oxidizing the stack earlier was also thought to improve the cathode microstructure. Cathode oxidation conditions were as follows:

- * The anode inlet manifold pressure was to be kept 2.8 iwc above the cathode outlet manifold pressure to avoid any possibility of oxidizing the anodes.
- * Temperature ramps were maintained at 1 °C/hr for the enclosure furnace and 4.5° C/hr for the gas temperatures up to 580 °C.

- * The differential between the maximum stack temperature and the maximum furnace enclosure temperature was kept below 20 °C.
- * The maximum temperature differential across a cell was 30 °C.
- * Gas chromatographic analysis was used to verify between 0.2 and 0.5% O₂ in the cathode inlet.
- * The clamping pressure was reduced to 30 psi in accordance with the schedule.

Heat up to Operating Temperature

At the completion of cathode oxidation, heat up to operating temperature was initiated.

Acceptance Testing

Open Circuit Voltage

Open circuit voltage (OCV) was measured twice during acceptance testing. The gas compositions were slightly different for the two OCV measurements but the stack temperature was the same. The first measurement was taken following introduction of process gases. The initial OCV was 268 V for the entire stack or 1.072 V for an average cell. The second measurement was taken at the end of Acceptance Testing, 212 hours later. The final OCV was 267 V for the entire stack or 1.068 V for the average cell. The predicted OCV for this gas composition and temperature was 1.073 V. Figure 7 illustrates the average OCV of each cell group measured as well as the individual cell voltages measured at the beginning and end of operation.

Figure 7 Open Circuit Voltage

Performance Testing

The total time that the stack spent on load was 210 hours. The maximum load reached was 546 A and the maximum power output was 107 kW. Figure 8 illustrates the stack voltage throughout performance testing.

Figure 8 Stack Voltage During Performance Testing

Figure 9 shows the current density throughout performance testing and Figure 10 shows the power output during performance testing.

Figure 9 Current Density During Performance Testing

Figure 10 Power Output During Performance Testing

Figure 11 illustrates the energy generated during performance testing.

Figure 11 Energy Produced During Performance Testing

The stack operated at a steady state load of approximately 400 A for 164 hours with only minor interruptions due to a power outage and a few hours of operation at 546 A. During this time no loss in performance was seen. Throughout the steady state operation, the fuel flows were once through. At twenty hours of operation at 420 A the stack voltage was 212 V and at 163 hours of operation at 420 A the stack voltage was 215 V. The stack met the acceptance criteria of 207.5 V - 210 V at 40 mA/cm². The average power produced during the 400 A operation was 90 kW.

Wet Seal Efficiency

Wet seal efficiencies were calculated for both the anode and cathode sides on a dry basis. Two different methods were used to calculate wet seal efficiency. The first method is the same method that is used to calculate wet seal efficiency in subscale cells. The anode wet seal efficiency is the dry outlet gas compensated for the water gas shift reaction divided by the inlet dry gas. The cathode wet seal efficiency is the dry outlet gas divided by the inlet gas minus the gas utilized based upon the operating current. Since the cathode gas contained recycled gas, it was assumed that the amount of recycled gas was constant on the inlet and outlet.

The second method for calculating wet seal efficiency utilized the MCP-cell program to predict the dry outlet gas on both the anode and cathode sides. The inputs to the MCP-cell program were the cell operating conditions at the time, the anode inlet flow rates and gas composition based on these measured flow rates, and the cathode inlet flow rates and compositions from gas chromatographic analysis. Gas chromatographic analysis was not used for the anode gas because the program includes the water-gas shift reaction. The measured dry outlet flow rates were divided by the predicted outlet flow rates to obtain a wet seal efficiency. The disadvantage of both methods is that there is no compensation for any gas crossover. If any gas is crossing from one side to the other, this could indicate a wet seal efficiency which is higher than calculated because the additional gas flow is included in the measured outlet gas flow rate.

Both methods of calculating wet seal efficiency gave similar results. Wet seal efficiency calculations were performed for steady state operation at a load of 420 A. Using method #1 the calculated anode wet seal efficiencies ranged from 77.4% to 99.9% with an average of 86.7%. Using method #2 the calculated anode wet seal efficiencies ranged from 75.4% to 100.5% with an average of 86.2%. Using method #1, the calculated cathode wet seal efficiency ranged from 95.4% to 98.5% with an average of 97.6%. Using method #2, the calculated cathode wet seal efficiency ranged from 92.7% to 98.5% with an average of 97.5%. For comparison, a subscale cell, ATF-01, which was conditioned with the same schedule as the PDT stack had an anode wet seal efficiency of 92% and a cathode wet seal efficiency of 100%.

Cooldown

During cooldown, at a stack temperature of approximately 200 °C, a crossover test was performed. The test consisted of three separate steps. During the crossover test, the anode gas consisted of N₂ and the cathode gas consisted of CO₂. Both anode and cathode gas contained

recycle material. During most of the test the flows remained constant and only stack inlet and outlet pressures were varied. Gas chromatographic analysis was performed to confirm inlet and exhaust gas compositions.

To determine the amount of crossover, the total flows into and out of the stack were used. It was assumed that the flow from recycle was the same into and out of the stack. The amount of cathode gas crossing into the anode side was then calculated by subtracting the amount of CO_2 in the inlet from the amount of CO_2 in the outlet. The amount of anode gas crossing into the cathode side was then calculated by subtracting the amount of N_2 in the cathode inlet from the amount of N_2 in the cathode outlet. The amount of anode gas lost to the atmosphere was calculated by taking the amount of N_2 in the outlets and subtracting it from the amount of N_2 in the inlets. Likewise, the amount of CO_2 in the inlets minus the amount of CO_2 in the outlets was considered to be the cathode gas lost to atmosphere. Any negative crossover indicates that more gas is present than originally supplied.

Post Conditioning Observations and Measurements

Post Conditioning Leak Tests

Four leak tests were performed on the stack at ambient temperature. The first three tests were performed while the stack was still in the ATF and the supply and exhaust piping was still attached to the ATF. The fourth test was conducted after the ATF piping was removed. This test was similar to the pre-conditioning leak test that was conducted in the dry room. In this test, gas was supplied to the stack with the Leak Test Rig.

Using ATF and Gas Chromatograph Analysis

Three separate tests were conducted at ambient temperature after the stack was cooled down. Tests were performed while the stack was still in the ATF and the supply piping was still connected. In all three tests, gas was supplied to the stack from the ATF and exhausted through the ATF so that all inlet and outlet gas flows could be measured. Gas compositions were determined by gas chromatograph analysis.

Leak Test #1

In test #1, N_2 was fed to the anode side and CO_2 was fed to the cathode side. All flows were once through and both recycle blowers were turned off. Inlet gas flow rates were measured with rotameters and outlet gas flow rates were measured with an orifice plate. Gas compositions were determined by doing gas chromatographic analysis. Flows were set whereby the anode and cathode pressures were approximately equal.

The results would then imply that approximately 20% of the anode flow was crossing into the cathode side and approximately 8% of the anode flow was leaking out of the stack. Approximately 15% of the cathode flow was crossing into the anode side and approximately 1% of

the cathode flows was leaking to the atmosphere when the pressure differentials are all approximately equal.

Leak Test #2

In test #2, nitrogen was fed to the anode side and CO₂ was fed to the cathode side. The pressures were set so the cathode pressure was above the anode pressure by approximately 3 iwc. Inlet gas flow rates were measured with rotameters and outlet gas flow rates were measured with an orifice plate. Gas compositions were determined by gas chromatographic analysis.

A small amount of N₂ was detected in the cathode outlet even though the cathode pressure was maintained higher than the anode pressure. This was detected not only in the cathode outlet analysis, but also in the individual cathode outlet manifolds. One hundred seventy four slpm of N₂ were not accounted for in the outlets and 90 slpm CO₂ were not accounted for in the outlets.

The results of this test would indicate that approximately 4% of the anode gas was crossing into the cathode side and approximately 15% of the anode gas was leaking to the atmosphere. Approximately 45% of the cathode gas was leaking into the anode side and approximately 7% of the cathode gas was leaking to the atmosphere.

Leak Test #3

The third crossover test performed was carried out with no anode gas flows. CO₂ was fed to the cathode side and both anode and cathode exhaust flows were measured. At these conditions, the cathode pressure was about 2 iwc higher than the anode pressure. A higher pressure differential between the anode and cathode sides could not be obtained. As the cathode pressure was increased the anode pressure responded in a like manner.

Crossover Test with Leak Test Rig

The crossover tests that were performed in the dry room and in the ATF prior to stack conditioning were repeated after conditioning and cool down were completed. The test was performed while the stack was still in the ATF but after all of the process pipes were disconnected. Gas was supplied to the stack through the stack leak test rigs. The test consisted of three major steps completed for each of two different pressures. One test was performed at a differential pressure of 5 iwc and the other was performed at a differential pressure of 10 iwc. The procedure was the same as that used for the pre-conditioning leak test.

The acceptance criteria for this room temperature crossover test was ≤ 0.4 slpm/cell of any gas leakage in the stack at a clamping pressure of 40 psi and a differential pressure of 10 iwc.

Leak Test Summary

Table 3 summarizes the leak test data obtained during all of the leak tests performed before and after conditioning.

Table 3 Summary of Leak Tests

	Clamping Pressure (psi)	Differential Pressure (iwc)	Leakage between anode and cathode (slpm/cell)	Leakage from anode to atmosphere (slpm/cell)	Leakage from cathode to atmosphere (slpm/cell)
Pre-conditioning	40	5	0.036	0.164	0.116
	40	10	0.026	0.298	0.222
Post conditioning ATF test #1	40	0	0.796	0.172	0.032
Post conditioning ATF test #2	40	3	2.476	0.696	0.360
Post conditioning ATF test #3-no anode gas	40	2	1.696	0.00 - no anode gas supplied	0.240
Post conditioning ATF with leak test rig	40	5	0.127	0.115	0.113
	40	10	0.114	0.235	0.239
After power module delivery to NAS Miramar	40	5	0.103	0.033	0.035
	40	10	0.147	0.052	0.052

Subscale Cell Performance Verification Testing

Six subscale 100 cm² test cells were operated to aid in evaluation and performance analysis of the stack during assembly, conditioning, acceptance testing, and plant operation. These test cells are M0195001, 95004, 96001, 96002, 96003, 96004.

The purpose of cell test M0195001 was to determine the baseline operating performance of components under system conditions at 3 atm operating pressure. In addition, the test was used to verify the assumptions used in calculating the carbonate loading requirements for the PDT-1 250 kW stack. Testing with system gases demonstrated cell performance of 747 mV at 160 mA/cm² with 75% fuel utilization at an operating pressure of 3 atm. This test was terminated after 1851 hours of operation.

The purpose of cell test M0195004 was to simulate stack operating modes for identification of failure modes. The cell contained active area components meeting the Unocal demonstration stack specifications. This cell included the first subscale simulation of ATF performance evaluation conditions.

The purpose of cell test M0196001 was to acquire additional benchmark performance levels for a fuel cell operating in a simulated ATF and under power plant operating conditions. The cell contained active area components of the stack specifications. The cell demonstrated that state-of-the-art bench scale cell performance of this cell package under ATF operating conditions is 894 mV at 40 mA/cm². This verified previously obtained data from cell test M0195004 which indicated simulated ATF performance of 880 mV. Testing with system gases demonstrated cell performance of 705 mV at 160 mA/cm² with 75% fuel utilization at an operating pressure of 1 atm. The test was terminated after 1500 hours of operation.

The purpose of cell test M0196002 was to evaluate full scale conditioning procedures and to acquire performance levels at simulated ATF operating conditions. The cell contained active area components of the stack specifications. The cell was conditioned according to the schedule and demonstrated similar performance to that obtained from the stack during acceptance testing. Under ATF operating conditions the cell performance was 847 mV at 40 mA/cm². Testing with system gases demonstrated cell performance of 641 mV at 120 mA/cm² with 56% fuel utilization at an operating pressure of 1 atm. The test was terminated after 865 hours of operation.

The purpose of cell tests M0196003 and M0196004 was evaluation of plant start up and operating conditions on cell performance. The cells contained active area components of the stack specifications. Test M0196003 was also used to acquire additional cell performance levels at simulated ATF operating conditions.

During stack conditioning and acceptance testing, data analysis from the subscale testing focused on identification of acceptable ATF cell performance and projected system cell performance based upon ATF cell performance data. Table 4 summarizes data obtained at the subscale and full area testing at ATF and system conditions.

Table 4 ATF Cell Performance Summary

Test ID #	Operating Conditions				Cell Performance			
	Fuel		Oxidant		1 ATM Operating Pressure			
	Gas Mixture	U _{t,fuel} (%)	Gas Mixture	U _{t,CO2} (%)	Calc. Current Density (mA/cm ²)	Cell Voltage (mV)	IR Free (mV)	0.62 cm ² *ohm Projected Full Area (mV)
M0195004	ATF-F1	50	ATF-O2	30	39	880	900	876
M0195004	SF-01	75	SO-1	40	157	653	733	636
M0196001	ATF-F1	50	ATF-O2	30	39	894	906	882
M0196001	SF-01	75	SO-1	40	157	705	754	656
M0196002	ATF-F1	50	ATF-O2	30	39	847	861	837
M0196002	SF-01	75	SO-1	40	118	641	682	609
Unocal	ATF-F1	51	ATF-O2	28	37	816*	-	-
PDT-1	ATF-F1	49	ATF-O2	23	39	852*	-	-

* Average Cell Voltage

The tested stack was shipped to NAS Miramar in August, 1996.

Task 4 BOP Fabrication

Objective: Purchase major equipment items, shop test equipment, fabricate skid in accordance with design specifications, and ship the BOP skid to demonstration site.

Discussion: Stewart & Stevenson's primary responsibility was to design a unitized equipment module to incorporate the piping, valves, instrumentation, and equipment required to operate the 250 kW demonstration plant. The design requirements were derived from information developed by Bechtel and M-C Power, and represented on project PFDs, P&IDs, Instrument Data Sheets, and Description of Operation. Major components of the systems, such as the recycle blower, turbocharger, turbocharger start-up blower, cathode start-up heater, hot gas cooler, heat recovery steam generator, desulfurization vessels, electrical conditioning/protection, motor control center, and distributed control system were specified by Bechtel and procured by M-C Power. Material used in the fabrication of the BOP module was purchased by S&S. The equipment module was delivered to the plant site in February, 1996. M-C Power fabricated the fuel cell stack which was not a part of the BOP skid.

Subtask 4.1 Equipment Procurement

Discussion: The IHI plate reformer represents the largest piece of equipment procured for the PDT demonstration. To minimize reformer problems during pre-commissioning and start-up, all four 62.5 kW sections of the reformer were hot tested in the IHI factory. A purchase order was placed with an independent inspection company, Japan Inspection Company (JIC). JIC witnessed the hot tests and conducted a pre-shipment inspection on M-C Power's behalf. The results of the hot test showed that the reformer worked according to the original specifications. Reports on the hot test were received from JIC confirming the results of the hot tests. The reformer was placed on its foundation at the job site in February, 1996. IHI personnel were at the job site in March to inspect the reformer and its installation. The reformer was in good working order, as summarized by IHI's inspection report.

IHI personnel arrived back at the job site in September, 1996, to conduct training classes, to assist in start-up activities and to perform an acceptance test.

Subtask 4.2 BOP Assembly

Discussion: The BOP assembly was completed at S&S's plant in January, 1996.

Subtask 4.3 BOP Shop-Testing and Shipping

Discussion: The BOP equipment skids were shipped to the job site in February, 1996. Stewart & Stevenson personnel spent time at the job site to assist with the installation and testing of the BOP skids. They also were involved with the hot gas recycle blower shaft seal testing.

Task 5 Plant Site Activities

Objective: Prepare site, install, and commission 250 kW demonstration plant.

Subtask 5.1 Site Preparation

Discussion: SDG&E was responsible for all site preparation work and a topical report on the subject was delivered to DOE on January 22, 1996.

Subtask 5.2 Plant Installation

Discussion: SDG&E and its subcontractors completed the plant installation work on April 15, 1996. Bechtel verified all installation work to insure that it met the design criteria.

The plant is a fully integrated, natural-gas fueled 250 kW proof-of-concept power plant located at NAS Miramar in San Diego. The power plant imports pipeline natural gas and city water

(for boiler feed water preparation and cooling purposes) and exports electricity and steam. Since the cooling water requirement for the plant is very small (<2 gpm), city water is used for cooling on a once-through basis and discharged to the Naval base sewer system. Boiler blowdown and wastewater from the boiler feed water treatment is also discharged to the base sewer system.

The plant is organized into the following major areas:

- Area 1: Fuel Cell System
- Area 2: Instrument Air System
- Area 3: Inert Gas System
- Area 4: Boiler Feed Water Treatment
- Area 5: Electrical Systems
- Area 6: Plant Control System

A description of these plant facilities is provided below.

Fuel Cell System (Area 1)

A process flow diagram for this plant area is shown in Drawing B-101 along with major process flows for the plant nameplate capacity.

Fuel Preparation Section

The natural gas pressure at the plant boundary varies from 44 psig to 50 psig depending on the season and time of day. The incoming natural gas flow rate is controlled in order to maintain desulfurized natural gas feed available at 40 psig.

A two-stage desulfurization system reduces the natural gas sulfur concentration to 0.1 ppmv to protect the reformer catalyst and the fuel cell from poisoning. The first stage (C-100) is a bed of TOSPIX-II catalyst (a proprietary catalyst by Tokyo Gas) and serves to remove the major portion of the sulfur species. The second stage (C-101) is a carbon adsorption bed for removal of species not removed in the first stage.

The desulfurized natural gas is heated with flue gas in the feed preheat coil (E-102) of the heat recovery steam generator (HRSG) and then mixed with 1000°F superheated steam to achieve a 3-to-1 molar steam-to-carbon ratio. At on-load conditions, the mixed feed temperature is about 800°F. The mixed feed is processed in the IHI reformer (R-200) where the natural gas/steam mixture is converted to reformat (i.e., carbon monoxide and hydrogen) at approximately 1450°F and 43 psia.

- E-100A/B
NATURAL GAS COMPRESSOR
7.5 HP-CONSTANT
TYPE PISTON
CAPACITY: 45 SCFH

E-102
NATURAL GAS COOLER
6 MBTU/HR MAXIMUM

E-104
PULSATION DAMPENER
16 CU. FT.
- E-100
TOSPIX-II BED
12 CU. FT. BED
OD: 2 FT.
HEIGHT: 8 FT OVERALL

E-101
ACTIVATED CARBON BED
12 CU. FT.
OD: 2 FT.
HEIGHT: 8 FT OVERALL

E-103
BLOWDOWN DRUM
4 CU. FT.
- E-100
BFV PUMP
1/4 HP
4 GPM MAXIMUM

E-100
ECONOMIZER
300 MBTU/HR (RATED)
285 MBTU/HR (NORMAL)

E-101
BOILER
952 MBTU/HR (RATED)
907 MBTU/HR (NORMAL)
- E-102
STEAM DRUM
20 CU. FT.

E-102
NATURAL GAS PREHEAT COIL
20 MBTU/HR (RATED)
18 MBTU/HR (NORMAL)
- E-101
STEAM SUPERHEAT COIL
120 MBTU/HR (RATED)
3 MBTU/HR (NORMAL)

E-104
AIR PREHEAT COIL
928 MBTU/HR (RATED)
40 MBTU/HR (NORMAL)
- E-300
START-UP BURNER
1 MBTU/HR MAXIMUM

E-300
START-UP AIR BLOWER
CONSTANT SPEED
1/3 BHP (RATED)
335 SCFH (RATED)
- E-200
REFORMER PRESSURE VESSEL
OD: 1.2 FT
HEIGHT: 11.5 FT O
- R-200
1HI PLATE REFORMER
M/AL REFORMING
Pd/AL COMBUSTION
4 PARALLEL MODS

STREAM NO.		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
FLOW RATES, lbmol/hr	STREAM NAME	TOTAL NATURAL GAS TO UNIT	DESULFURIZED NATURAL GAS	GAS FR. E-102 PREHEAT COIL	BFV TO ECONOMIZER	BFV FR. ECONOMIZER	BLOW-DOWN	EXPORTED SATO STEAM	SATO STEAM TO SUPER-HEATER	STEAM TO E-200	E-200 SUPER-HEATED STEAM	SUPER-HEATED STEAM TO R-200	REFORMER FEED	ANODE FEED	ANODE EFFLUENT	CATHODE RECYCLE TO R-200	AIR TO COMPRESSOR	AIR TO R-200	AIR TO CATHODE	COMP. AIR BYPASS TO HRSG	COND. DRAIN
H ₂														22.39	6.10						
CO														4.36	1.92						
CO ₂		0.030	0.030	0.030																	
H ₂ O					58.33	58.33	1.75	35.53	180	21.04	21.04	21.04	0.030	2.46	23.61	5.10	0.05	0.02	0.03		5
N ₂		0.135	0.135	0.135									21.04	11.84	28.13	26.08	1.60	0.67	0.83	0.10	26
Ar													0.135	0.14	0.14	79.79	115.88	48.45	59.99	7.44	128
O ₂																0.94	1.39	0.58	0.72	0.09	1
SULFUR SPECIES, AS H ₂ S		5.00ppm	0.10ppm	0.10ppm												11.06	31.09	13.00	16.09	1.99	24
SO ₂													0.02ppm	0.02ppm	0.01ppm						
CH ₄		6.293	6.293	6.293									6.293	0.20	0.20						<0.01
C ₂ H ₆		0.248	0.248	0.248									0.248								
C ₃ H ₈		0.043	0.043	0.043									0.043								
C ₄ H ₁₀		0.011	0.011	0.011									0.011								
C ₅ H ₁₂		0.003	0.003	0.003									0.003								
C ₆ H ₁₄		0.001	0.001	0.001									0.001								
TOTAL, lbmol/hr		6.77	6.77	6.77	58.33	58.33	1.75	35.53	180	21.04	21.04	21.04	27.81	41.38	60.10	122.97	150.00	62.72	77.66	9.62	185
TOTAL, lb/hr		116	116	116	1,051	1,051	32	640	180	379	379	379	496	496	1,619	3,321	4,327	1,409	2,240	278	51
TEMPERATURE, °F		85	85	354	61	328	240	334	334	350	1000	1000	801	1250	1100	1290	60	334	334	334	
PRESSURE, PSIA		59.00	54.00	44.00	125.00	115.00	25.00	115.00	115.00	105.00	95.00	95.00	43.33	43.04	42.74	42.70	14.44	45.00	45.00	45.00	42

2220801.DCN

K-201
TURBOCHARGER
127 BHP (RATED)
960 SCFM (RATED)

K-202
TURBOCHARGER START-UP BLOWER
CONSTANT SPEED
25 BHP (RATED)
280 CFM (RATED)

The IHI reformer (R-200) is a "plate type" reformer consisting of four modules in parallel. Each module has multiple cells with each cell containing a reforming chamber sandwiched between two combustion chambers. The reforming chamber is packed with a nickel/alumina catalyst and the combustion chamber is packed with a palladium/alumina catalyst.

Reforming is an endothermic reaction and the necessary heat input is supplied by combusting the anode exhaust gas in the combustion chamber. Cathode recycle gas provides oxygen required for combustion as well as additional sensible heat. Compressed air is also introduced to control the inlet oxidant temperature to the combustor.

Mixed oxidant (cathode recycle gas and air) enters the combustion chamber and flows counter-current to the reformed gas. The inlet portion of the combustion chamber is a heat exchange zone which serves to preheat the incoming oxidant by recovering heat from the exiting reformat. During normal operation, the reformat enters the heat exchange zone close to 1450°F and exits the reformer between 800°F and 1260°F, depending on the plant load and the anode fuel utilization. Also in the heat exchange zone, fuel (anode exhaust) is dispersed into the oxidant to achieve a combustion gas temperature leaving the heat exchange zone of 1460°F (i.e., 2°F to 18°F greater than the incoming reformat temperature). The hot combustion gas exiting the heat exchange zone enters a reforming zone where the heat required for reforming is provided through counter-current heat exchange between the reforming and combustion chambers. During normal operation, the combustion gas cools to 1130°F before exiting the combustion chamber.

The reformer is enclosed inside an inert gas purged pressure vessel (C-200). Inert gas purging is desired to avoid the buildup of combustible gas mixtures inside the reformer pressure vessel (C-200). The purge gas flows through the reformer pressure vessel into the fuel cell pressure vessel (C-201).

Fuel Cell Section

Reformat from the IHI reformer is fed to the anode of the fuel cell (R-201). The anode preheater (P-203), an electric heater, on the anode feed line preheats the reformat entering the anode to 1250°F. A minimum anode inlet temperature of 1238°F must be maintained to avoid carbon formation at the anode inlet. This minimum temperature is calculated from equilibrium data assuming carbon formation via the Boudouard reaction ($\text{CO}_{(g)} = \frac{1}{2} \text{CO}_{2(g)} + \frac{1}{2} \text{C}_{(s)}$), based on the targeted reformat composition and the plant operating pressure.

In the anode, reformat reacts with carbonate ions transferred from the electrolyte to form carbon dioxide and water. At full load, this reaction consumes between 70% to 80% of the fuel. The anode effluent exits the cell at a minimum temperature of 1100°F and the composition of the anode effluent is governed by the water-gas shift reaction. This temperature increases at part load to a maximum of 1300°F. The remaining 20% to 30% of the fuel is consumed downstream in the combustion zone of the IHI reformer (R-200) to provide the energy for reforming. The flue gas from

the combustion zone is combined with recycled cathode gas and compressed air to provide cathode feed gas.

In the cathode, carbon dioxide and oxygen react to form carbonate ion. The formation of carbon ion at the cathode and its subsequent consumption at the anode establishes a flow of electrons between the anode and cathode (i.e., current). Carbon dioxide is recycled from the anode to the cathode in order to replenish the electrolyte consumed at the anode.

The overall cell reaction is that of combustion with a net production of carbon dioxide and water. Circulation of cooled cathode exhaust via the cathode recycle blower (K-200) provides the necessary mass flow to remove the dissipated heat and to maintain a 1290°F cathode exhaust temperature. With K-200 providing a constant cathode feed rate, the differential temperature across the cathode will be varied to achieve the desired heat removal rate. The system is designed for a minimum 1100°F cathode inlet temperature at the maximum rated plant design. This temperature increases at part load to a maximum of 1300°F. The inlet temperature is adjusted through heat rejection in the cathode recycle cooler (E-200) and the introduction of excess air to the cathode feed. The cathode start-up heater (P-202) located upstream of the recycle blower supplies heat input during start-up and provides a mixing zone for the cathode feed gases during operation.

The fuel cell operates close to 42 psia and consists of a 250-cell stack inside a pressure vessel. The pressure vessel (C-201) is internally insulated and contains electric guard heaters (P201 A-D) to keep the cell stack warm during off load states. The pressure vessel is purged with the purge gas exiting the reformer pressure vessel (C-200). The purpose of the purge gas is to prevent the buildup of any combustible gas mixtures inside the fuel cell vessel. The purge gas exiting the fuel cell pressure vessel is routed to the cathode inlet.

The mixture of molten carbonate salts serves not only as the electrolyte in the fuel cell, but also provides a "wet seal" to isolate the anode and cathode gases from each other as well as from the gas inside the pressure vessel. The plant has been designed to limit the differential pressure across the wet seals to a maximum of 20 iwc. Since inert gas purge discharges into the cathode inlet line, the pressure inside the fuel cell vessel is maintained essentially at the cathode inlet pressure. The purge gas flows to the fuel cell pressure vessel via the reformer pressure vessel. The purge gas flow rate is controlled to achieve two tasks, 1) to minimize back-diffusion of oxygen from the cathode inlet into the fuel cell pressure vessel and 2) to dilute and remove the combustible gases that leak from the fuel cell into the pressure vessel.

The inverter (P-204) regulates the amount of current drawn by modulating the impedance across the fuel cell stack. The fuel cell d.c. power is first transformed to higher voltage d.c. power and then inverted to a.c. power. The inverter has a solid state design and produces 3-phase 480 volt a.c. power with less than 3% total harmonic distortion. The fuel cell contains 250 cells connected in series and produces a maximum of 1822 amps and 182 volts d.c. at full load.

Energy Recovery Section

Heat and mechanical energy are recovered from the cathode exhaust to increase the efficiency of the power plant. The net cathode exhaust is expanded through the turbocharger (K-201), thereby recovering mechanical energy directly to supply compressed air to the fuel cell. The turbocharger exhaust gas (flue gas) is sent to the heat recovery steam generator (HRSG) where heat is recovered to preheat air (E-104), superheat steam (E-103), preheat natural gas (E-102), generate 100 psig saturated steam (E-101), and preheat boiler feed water (E-100). The plant is capable of producing more saturated steam than required by the process and exports this excess steam to the NAS Miramar steam distribution system. Boiler feed water (BFW) is delivered to the HRSG system at near ambient conditions. A BFW pump (G-100) is included to boost the pressure of the BFW to about 140 psig. The BFW flow is controlled to maintain the proper level in the boiler. The boiler blowdown is controlled by intermittently opening the boiler blowdown valve to achieve a predetermined flow rate. Boiling in the economizer is avoided at part load by overriding the boiler blowdown control to maintain a continuous blowdown and thereby increase the BFW flow rate through the economizer. The flow of exported steam can be controlled by bypassing flue gas in the HRSG around the boiler and economizer.

Heat rejected from the cathode recycle gas is utilized for superheating steam in the recycle gas cooler (E-200). During normal operation, the steam superheat coil (E-103) in the HRSG superheats 338°F saturated steam to 350°F. The 350°F is then sent to E-200 for further superheating to 1000°F. During standby or very low loads, steam bypasses E-200 totally and instead the steam superheat coil (E-103) in the HRSG superheats saturated steam at 338°F to 700°F. The low superheating duty for E-103 during normal operation is to insure that a some flow will be maintained through E-103 at all times. The temperature of steam exiting E-103 is limited by the temperature of the exhaust gas in the HRSG.

At part load, compressed air not required by the process is dumped from the compressor outlet to the expander inlet. This is necessary in order to keep the turbocharger operating at its rated speed and therefore providing 46 psia compressed air to the process. The dumped air flows from the compressor to the expander via the HRSG, where heat is recovered from the turbocharger exhaust to maintain a high expander inlet temperature. This feature extends the turndown range for the turbocharger, permitting operation with a lower flow of cathode exhaust gas from the process. The turbocharger can, with heat input from the start-up burner (F-300), operate completely on circulation of compressed air with no cathode exhaust flow.

Start-up Equipment

The plant is heated at start-up initially through heat input from the following electric heaters: reformer end heaters (P-200 A-H), reformer line heaters (P-205 A-E), reformer header heaters (P-206 A-D), fuel cell guard heaters (P-201 A-D), cathode start-up heater (P-202), anode preheater (P-203).

Only the anode preheater is used continuously during operation. The remaining electric heaters provide heat input at start-up and provide backup temperature control during operation.

The turbocharger start-up blower (K-202) circulates air from the compressor side of the turbocharger, through the air preheat coil (E-104) in the HRSG, to the expander side of the turbocharger. At start-up, the heat inputted in E-104 is sufficient to operate the turbocharger at full speed with all the compressed air circulated to the expander. The K-202 blower is sized to drive the expander at the minimum speed required for starting the turbocharger.

A gas-fired start-up heater (F-300) and its associated air blower (K-300) are at the inlet to the HRSG to provide hot flue gas during start-up. The HRSG can be started up to preheat air to start the turbocharger, generate saturated steam, and to provide superheated steam when insufficient heat is available in the cathode recycle cooler (E-200).

Instrument Air System (Area 2)

The major components of the instrument air system are depicted in Drawing B-102. The system consists of an air-cooled air compressor (K-400), an air receiver (C-400), and a regenerative dryer system (V-400). The instrument air system is sized to provide a maximum of 64 scfm air at 90 psig with the air compressor operating continuously at a maximum ambient temperature of 110°F. The actual instrument air required by the plant is estimated to be less than 36 scfm.

The K-400 air compressor is rated to compress 90 acfm air at ambient conditions to a maximum of 125 psig. The compressor operates intermittently with K-400 switching on when the pressure in the air receiver (C-400) drops below 115 psig and switching off when the pressure in C-400 achieves 125 psig.

The air receiver (C-400) has approximately 16 ft³ (120 gallons) capacity. Compressed air is supplied to the drying system at a minimum of 110 psig. In addition to supplying instrument air, a connection is provided to provide plant service air. A backpressure regulator on the service air line prevents the service air demand to draw down the receiver pressure below 110 psig. Plant nitrogen is used to back-up the instrument air system in the event of failure.

The drying system (V-400) consists of two alumina drying beds in parallel. The drying system utilizes one bed for drying while regenerating the unused bed with a portion of the dry air product. Approximately 15% of the system air capacity is consumed for bed regeneration. The dried instrument air has a maximum dew point of -40°F at operating pressure which is equivalent to 35 ppmv water at 90 psig.

BACK-UP NITROGEN
2X255 SCF CYLINDERS
COMPRESSED GAS
2,300 PSIG

V-500
NITROGEN SUPPLY
SYSTEM
6000 GAL.
LIQUID STORAGE

K-400
INSTRUMENT
AIR COMPRESSOR
TYPE: ROTARY, AIR COOLED
30 HP MOTOR
28 BHP, 88 CFM (RATED)

C-400
INSTRUMENT
AIR RECEIVER
16 CU.FT.
OD: 2 FT.
HEIGHT: 6 FT. OVERALL

V-400
AIR DRYING SYSTEM
TYPE: REGENERATIVE
2 BEDS ACTIVATED ALUMINA
0.6 CU.FT. BED VOLUME

C-901
SEAL POT
OD: 2 FT.
HEIGHT: 8 FT. OVERALL

801
RGE GAS RECEIVER
CU. FT.
12.5 FT.
100 FT. 1 FT. OVERALL

V-600
BOILER FEEDWATER
TREATMENT PACKAGE
CAPACITY: 2 GPM

T-500
WATER TANK
CAPACITY: 26 GAL.

G-500
WATER METERING PUMP
0.24 GPM
100 psid.

2-21-94	REVISED & REISSUED FOR PROJECT USE	BP	10/10
6-29-95	REVISED & REISSUED FOR PROJECT USE	ORIGINAL	SIGNED
1-12-95	ISSUED FOR PROJECT USE	ORIGINAL	SIGNED
11-1-94	REVISED AND REISSUED FOR NEW SITE	ORIGINAL	SIGNED
1-10-94	ISSUED FOR APPROVAL	ORIGINAL	SIGNED
DATE	REVISIONS	BY	CHKD BY
SCALE	NONE	DESIGNED	JJS/LMD
		DRAWN	PCL/CAD

BECHTEL CORPORATION
SAN FRANCISCO

M-C POWER SDG&E FUEL CELL
PRODUCT DEVELOPMENT TEST PROJECT
MIRAMAR U.S. NAVAL AIR STATION

PROCESS FLOW DIAGRAM
BALANCE OF PLANT

JCS No. 22200

DRAWING No. B-102

FEV 4

Inert Gas System (Area 3)

The inert gas system depicted in Drawing B-102 includes a nitrogen supply system and a bank of high pressure nitrogen gas cylinders. The system supplies nitrogen to the plant for three purposes: 1) to purge the fuel cell system, 2) to pressurize the fuel cell bellows (i.e., the clamping device in the fuel cell vessel to maintain wet seal integrity), and 3) to back-up the instrument air system.

The nitrogen supply system includes a liquid nitrogen storage tank, a nitrogen vaporizer, and a nitrogen delivery system. The nitrogen is delivered to maintain a nitrogen header pressure of approximately 200 psig. The nitrogen header supplies nitrogen to the purge gas receiver (C-801), the fuel cell bellows, and the instrument air header. Purge gas is withdrawn from C-801 to supply 45 psig purge gas to the plant.

Nitrogen sent to the bellows is regulated down to 100 psig. A bank of high pressure nitrogen cylinders serve as back-up for the nitrogen supplied to the bellows and is regulated to maintain a minimum pressure of 75 psig in the bellows header. Nitrogen used to back-up the instrument air system is regulated to maintain a minimum 80 psig pressure in the instrument air header in the event of a failure in the instrument air system.

Boiler Feed Water Treatment (Area 4)

The boiler feed water treatment system is a packaged plant designed to supply a minimum of 2 gpm boiler feed water continuously to the fuel cell system. The system consists of the following components: carbon prefilter, two-stage reverse osmosis unit, two mixed deionizing beds in parallel (1 active and 1 spare), and storage tank. An emergency bank of deionizing beds is also included to produce deionized water directly from city water. The boiler feed water system does not include a deaerator. The materials in the HRSG have been chosen to be compatible with non-deaerated water.

Electrical Systems (Area 5)

The major components of the electrical system include a direct current (d.c.) to alternating current (a.c.) inverter, a motor control center (MCC), heater control panels, power panels, adjustable speed motor drives, and an uninterruptable power supply (UPS). These components are housed in an electrical room located adjacent to the control room.

A d.c. cable bus connects the fuel cell to the inverter. The inverter converts the unregulated d.c. power generated by the fuel cell to regulated 480 volt a.c. power. The a.c. power output from the inverter is connected to a 4.16 kV 277 V/480 V transformer to supply power to the NAS Miramar substation. The inverter can not be operated in a load-following mode; therefore, the plant is shutdown if off site power is lost.

480 volt power is supplied to the MCC from SDG&E circuit 762 via a 12 kV 277/480 volt transformer. The MCC supplies all the power required by the balance of plant auxiliaries which include blowers, heaters, compressors, pumps, the distributed control system (DCS), building and area lighting and 208/120 volt miscellaneous power. A cable trench is provided for routing power and control cables from the electrical room to the equipment located on the various equipment skids. This trench also accommodates the control and instrumentation cables connecting the equipment skids to the DCS.

Two grounding systems are also provided. The first is an equipment grounding grid which connects all equipment to an earth ground and the second is an isolated and insulated grounding system for instrumentation and the DCS.

Plant Control System (Area 6)

The plant control system is designed to provide a central location for the operation of the plant with an appropriate level of automation. The primary control location is the central control room which is adjacent to the electrical room.

The plant control system is a microprocessor-based distributed control system (DCS). The DCS consists of: input/output (I/O) and process control unit cabinets, redundant primary operator consoles, color printer, shared data acquisition system equipped with modem and printer, dedicated modems for SDG&E and for the US Navy facility for remote monitoring of graphic displays, and a workstation (PC) with a printer.

The DCS performs control functions, continuous monitoring of the plant, data acquisition, data access and report generation, and miscellaneous tasks such as alarm, indication, trend recording and reporting.

The workstation is used to support changes in the plant configurations, testing and modification of graphic displays.

SDG&E and the US Navy each have an independent capability of accessing all graphic displays from a remote location for monitoring only. In addition, a common data acquisition system (DAS) is shared by SDG&E, M-C Power and other parties to access real time data, trends and historical files of predetermined points. Each user has defined their specific needs of data points, formats and software compatibility used in-house for their system.

The DCS power source is from the UPS. The UPS power is fed to a redundant power supply in the system. The DCS, including operator consoles, will remain in operation for a duration of 15 minutes in the event that site power is lost.

A stand alone programmable logic controller (PLC) based burner management system (BMS) is provided for the HRSG. The HRSG BMS interfaces with the DCS for control of the start-up burner (F-300) and start-up blower (K-300).

The BFW treatment package is provided with a stand alone PLC for control and monitoring purposes and provides status alarms to the DCS. The nitrogen storage system does not contain any automatic controls and does not interface with the DCS.

Subtask 5.3 Plant Commissioning

The demonstration test activities were initiated following completion of plant construction work. An audit was used to verify that the plant equipment and systems were installed in accordance with approved drawings and design documents. Deviations from the design bases were documented and subjected to in-depth analyses prior to acceptance. Remedial construction work was performed when dictated by the engineering review. Construction QA/QC inspection and testing procedures and reports were included in the construction documents.

Plant testing was performed sequentially in the following five areas:

1. Precommissioning Tests/Checkouts
2. Plant Subsystem Verification
3. System Startup and Plant Integration
4. Process and Control Testing without Fuel Cell Installation
5. Fuel Cell Startup and Plant Acceptance Tests

Precommissioning Tests/Checkouts

Pre-commissioning tests and checkouts were conducted in accordance with the pre-commissioning documents listed in Table 5.

Plant Subsystem Verification

Vendor data sheets, equipment specifications, operating and maintenance procedures were used for subsystem verification. These were supplemented by Bechtel specifications. Verification activities included equipment status, system cleanliness, control and alarm settings, leak and hydro testing results, and other pre-operational activities and checkout as required or recommended by equipment vendors.

System Start-up and Plant Integration

The plant start-up procedures are summarized in the start-up block diagram shown in Figure 12.

Table 5 Precommissioning Tests/Checkouts

Table 5 Precommissioning Tests/Checkouts			
	Test & Verification Documentation	Activity	Notes
System			Verification Documentation
			References
Electrical Power/DCS		Precommissioning	
Low & Medium AC Voltage	Megger, Phase Check, Grounding, Energize		Pre Comm Checkout Sheets
DC Voltage, UPS & Pwr Conditioner	Checkout and Energize		Pre Comm Checkout Sheets
DCS	Grounding, Shields, Termination, Fuse Inspection		Pre Comm Checkout Sheets
			Bailey Controls
Plant Utility			
Instrument Air	Checkout, Startup & Air blow		Airnetics Engineering/ Cleaness Specs
Nitrogen/Purge Supply	Checkout, Air Blow, Operation		Cleaness Specs
City Water	Checkout, Water Flush, Operation		Cleanliness Verification
			Cleanliness Verification
			Cleanliness Verification
Valves			
Instruments	Calibration, Set Points and Alarm Settings		Instrument Cal Verification Check Out Sheet
			Vendor Manual
Control Valves	Air Stroke, Set and Failure mod Verification		CV I/P Pos. Checkout Sheet
			Instrument Data Sheet Book & Vendor Manual
Safety Valves	Verification of Factory Setting/Seal		Safety Valve Name Plate Data Sheet
			Vendor Manual
Gas Chromatograph	Calibration, Setup		Inst. Cal Verification Sheet
			Applied Automation
Plant Sub-Systems			
Natural Gas	Hydro (pressure/leak test), Air Blow, Activate		Sub Sys Verification Sheet
H2 System	Leak Test, Activate		Sub Sys Verification Sheet
CO2 System	Leak Check, Activate		Sub Sys Verification Sheet
Demin Water Injection	Air Blow		Cleanliness Certification
Process Piping	Air Blow, Cleaning		Cleanliness Certification
BFW Treatment	Checkout, Operation		Sub Sys Verification Sheet
			Vendor Package

Table 5 Precommissioning Tests/Checkouts

TABLE-5		
Test & Verification Documentation	Activity	Notes
System	BFW Piping	Cleanliness Verification
	BFW Pump	References
	HRSG Startup Air Blower	Verification Documentation
	Bump Motor, Rotation Check, Alignment	
	Bump Motor, Rotation Check, Alignment	
	Boiler Hydro, Water Flush	
	HRSG BMS Checkout/Operation	
	Refractory Cure	
	Chemical Cleaning (Boil Out)	
	Hand Clean, Load Catalyst	
Reformer Vessel/Desul Bed	Pressure/Leak Test	Pre Com Procedure
	Insulation & Resistance Check	Heater Check Sheet
	Insulation & Resistance Check	Heater Check Sheet
	Reformer End Heaters	Heater Check Sheet
	Reformer Plate Heaters	Heater Check Sheet
	Anode Preheater/Cathode SU Heater	Heater Check Sheet
	Turbocharger SU Blower	Heater Check Sheet
	Turbocharger	Wallow
	Hand Clean Lube Oil Tank and Fill	Rotoflow
	Bump Motor, Rotation Check, alignment	Rotoflow & Cleaness Spec
Recycle Blower	Bump Motor, Rotation Check, alignment	IAP
	Insulation & Resistance Check	M-C Power Wallow
	Bump Motor, Rotation Check, alignment	
	Startup and Acceptance	
	Checkout and Program Verification	
	In Service Operation	
	Purge Gas System Test	Startup Procedure
	BFW Operation	Startup Procedure
	HRSG Heatup	Startup Procedure
	Turbocharger Startup	Startup Procedure
Gas Chromatograph	GC Inservice Operation	GC Operating Procedure
	Nat. Gas Desul Bed Operation	Startup Procedure
	Reformer	Startup Procedure
	Reformer	Startup Procedure
	Reformer	Startup Procedure
	Reformer	Startup Procedure
	Reformer	Startup Procedure
	Reformer	Startup Procedure
	Reformer	Startup Procedure
	Reformer	Startup Procedure
Recycle Blower	Operation in Cascade	Startup Procedure
	Reformer End Heaters Startup	Startup Procedure
	Operation in Cascade	Startup Procedure
	Reformer	Startup Procedure
	Reformer	Startup Procedure
	Reformer	Startup Procedure
	Reformer	Startup Procedure
	Reformer	Startup Procedure
	Reformer	Startup Procedure
	Reformer	Startup Procedure

Table 5 Precommissioning Tests/Checkouts

TABLE-5				
Test & Verification Documentation				
System	Activity	Notes	Verification Documentation	References
Anode & Cathode Startup Heater	Operation, Tuning and Heat Loss Check		Startup Procedure	Watlow
	PAC Test			
HRSG	HRSG Performance Fuel Consumption v/s Steam Production		Test Protocol	Broach Vol 1
Desulfurization	Sulfur Removal from Natural Gas		Test Protocol	Tospix II Manual
Reformer	Reformer Performance Methane Slip		Test Protocol	IHI
Miscellaneous Equipment	Recycle Cooler & Electric heaters Duties, Recycle Blower Performance		Test Protocol	Watlow & IAP Manuals
PAC Test (Simulation)	Integrated Plant Operation w/o Fuel Cell		Test Protocol	Test Protocol
Inverter	Electrical Checkout/Load Test		Test Protocol	InverPower/SDGE Eng.
	Fuel Cell Startup/Testing Acceptance			
Fuel Cell Guard Heater	Heater Operation		Test Protocol	M-C Power/Watlow
Fuel Cell	Heatup, Melting, Tilt Measurement		Test Protocol	M-C Power
Fuel Cell	Hot Hold - Cross Over Tests		Test Protocol	M-C Power
Fuel Cell	Standby - OCV Tests		Test Protocol	M-C Power
Fuel Cell	On Load @ 0-10%: Voltage Drop		Test Protocol	M-C Power
Fuel Cell	On Load @ 10, 25, 50, 75, & 100% and Sustained Operation		Test Protocol	M-C Power
Integrated Plant	Integrated Plant Emissions @ 50, 100%		Test Protocol	Emission Contractor
Notes:				
1. This is a suggested startup sequence. The Sequence can be altered based on construction completion and system turnover.				

Equipment/Systems Started

Establishing Plant Utilities

Plant start-up requires the availability of electricity, nitrogen, instrument air, and the DCS. These systems are essential and should remain operational at all times, even when the unit is placed in the Off state.

During initial start-up, the natural gas compression and desulfurization systems are purged with nitrogen and pressurized with natural gas. These systems remain filled with natural gas when the unit is in the Off state and are manually purged and vented, only as necessary, for repairs and/or replacement of desulfurization catalyst/adsorbent. Before routine start-up, cooling water flow to each individual piece of equipment is verified. In addition, the purge gas supply and boiler feed water treatment systems should also be verified to be operational.

System Purging

The plant is purged to lower the oxygen concentration throughout the process to less than 1% by volume using the procedures listed below.

1. The purge gas (nitrogen) flow to the natural gas preheat coil (E-102) is initiated at a predetermined rate. The purge gas flows through the reforming chamber, fuel cell anode, and combustion chamber where it is mixed with the cathode gas.
2. Purge gas is introduced to the reformer pressure vessel (C-200) at a predetermined rate. The purge gas flows through C-200 to a differential pressure control valve located in the purge line connecting the reformer vessel (C-200) to the fuel cell vessel (C-201). This valve regulates the pressure in the reformer vessel to be approximately 2" higher than the pressure in the reformer inlet line. The purge gas from this valve flows through the fuel cell pressure vessel (C-201) and is mixed with the cathode inlet gas.
3. Purging is continued for a predetermined time. The oxygen concentration measured at the flue gas to the waste gate and the fuel cell pressure vessel purge gas outlet serve as an indication of the completeness of purging.
4. The purge flow to C-200 continues at its maximum flow rate until the desired oxygen concentration at the C-201 outlet is attained. The flow rate of this purge gas is then set at its minimum flow rate.
5. Purge flow to E-102 is continued until the specified purge gas volume has been consumed and the oxygen analyzers confirm that the oxygen content in the system is less than 1.0 vol. % oxygen.

Heating to 1100°F with Electric Heaters

The fuel cell is heated to 1100°F, and the reformer is heated to 1000°F, using heat input from the electric heaters at the following locations: reformer vessel, fuel cell vessel, anode inlet, and cathode inlet. When the maximum fuel cell temperature is approximately 1100°F, melting (end point approximately 1040°F) can be reasonably assured in all portions of the cell stack. During this period, inert gas containing carbon dioxide by injection is circulated through the reformer combustion chamber and the fuel cell cathode by the cathode recycle blower. Nitrogen and reducing safeguard gases in accordance with Table 6 are swept through the anode system and vented through the anode exhaust vent valve. The cathode system pressure will be controlled at approximately 3 atmospheres by the waste gate controller. The anode system pressure will follow the cathode pressure using differential pressure control for the anode exhaust vent valve. The procedures are as follows:

1. The system is pressurized with inert gas to approximately 3 atmospheres by setting the cathode waste gate controller at nominally 28 psig. The blower (K-200) is then started and its speed set so as not to exceed the maximum allowable cathode manifold differential pressure, 15 iwc.
2. The cathode start-up heater (P-202) is activated and programmed at a nominal rate of 10°F/hr. The fuel cell is heated at a rate of 10°F/hr up to 842°F/hr. The power input to the heater is limited to maintain a differential temperature across the cathode of less than 50°F.
3. The IHI reformer heaters (P-200 A-E, P-205 A-E and P-206 A-D) are activated to heat the reformer at a nominal rate of 10°F/hr up to 1000°F.
4. The anode preheater (P-203) is activated and boosts the anode inlet gas temperature to match the cathode inlet gas temperature up to 1000°F, at which the operators maintain the 1000°F anode preheater setpoint in manual.
5. The fuel cell guard heaters (P-201 A-D) are activated and maintain a temperature equal to the measured cathode outlet temperature.
6. The cathode system waste gate controller will maintain the cathode system at 28 psig provided the purge gas rate is adequate to overcome system leaks. Therefore, the vessel purge rate is increased manually, if necessary to maintain a constant system pressure.
7. The cathode recycle gas blower is operated at a speed such that the cathode inlet to outlet differential pressure does not exceed 15 iwc. Plant operators will adjust the blower speed at intervals during the heat-up to maintain, but not exceed, this differential pressure.

Table 6 Reducing Safegas Compositions and Estimated Quantities

<u>Temperature Range:</u> <u>Temp <392°F</u>		
<u>Actions:</u> <u>1. Nitrogen sweeps through system.</u>		
<u>Composition:</u>	<u>Mole/hr</u>	<u>mole %</u>
H ₂	<u>0.0</u>	<u>0.0</u>
N ₂	<u>2.1</u>	<u>100.0</u>
H ₂ O	<u>0.0</u>	<u>0.0</u>
<u>Total</u>	<u>2.1</u>	<u>100.0</u>

<u>Temperature Range:</u> <u>392°F < Temp < 662°F</u>		
<u>Actions:</u> <u>1. Close isolation valve</u> <u>2. Activate anode exhaust vent valve on DP control</u>		
<u>Composition:</u>	<u>Mole/hr</u>	<u>mole %</u>
H ₂	<u>0.033</u>	<u>1.5</u>
N ₂	<u>2.167</u>	<u>98.5</u>
H ₂ O	<u>0.00</u>	<u>0.0</u>
<u>Total</u>	<u>2.20</u>	<u>100.0</u>

<u>Temperature Range:</u> <u>662°F < Temp < 1000°F</u>		
<u>Actions:</u> <u>1. Hold at 1000°F</u>		
<u>Composition:</u>	<u>Mole/hr</u>	<u>mole %</u>
H ₂	<u>0.033</u>	<u>1.5</u>
N ₂	<u>2.057</u>	<u>93.5</u>
H ₂ O	<u>0.11</u>	<u>5.0</u>
<u>Total</u>	<u>2.20</u>	<u>100.0</u>

8. Carbon dioxide from gas bottles is injected directly into the cathode system when the fuel cell has reached 662°F. The injection may be batch or continuous at a rate set by the operators at the CO₂ metering rotameter based on the vessel purge rate. The injection quantities will provide an average a 5% CO₂ in nitrogen mixture.
9. Nitrogen gas is injected upstream of E-102 on the natural gas feed line and passes through the reformer. At the anode preheater inlet, water is introduced in accordance with Table 6. At the anode inlet, hydrogen is introduced in accordance with Table 6 and the gas mixture then passes through the anode before venting to the seal pot. An isolation valve in the anode exhaust line prevents the reducing safeguard gases from entering the cathode.
10. Pressure differential control valves located in laterals from the anode inlet line and the cathode inlet line prevent overpressuring the fuel cell wet seals when the anode system is isolated from the cathode system. These valves feature fully pneumatic controllers to prevent loss of control action in the event of a power failure. Furthermore the valves are actuated pneumatically through a dedicated air tank which provides an independent air supply in the event of loss of plant air.

Starting the HRSG

While the system is being heated to 1000°F, the HRSG is started by firing natural gas in the start-up burner (F-300). Completion of the HRSG start-up period is timed to coincide with the fuel cell and reformer attaining 1000°F. The general procedures are summarized below. Detailed start-up procedures are specified by the HRSG vendor.

1. The valve on the boiler feed water line is set to control the level in steam drum (C-102). Boiler feed water flows through the economizer (E-100) to establish the desired water level in C-102. The air displaced by the incoming water is vented by a solenoid operated vent valve located on the steam drum or upstream of the steam back pressure regulating valve.
2. The start-up air blower (K-300) is started to purge the HRSG and to provide air to the start-up burner (F-300).
3. Natural gas is introduced to the start-up burner to heat the HRSG at a predetermined rate.
4. The boiler warms up and starts producing steam. The solenoid vent valve on C-102 is closed when the boiler steam pressure begins to rise.
5. A solenoid operated vent valve is opened just upstream of the superheated steam flow control valve to vent noncondensibles.

6. Steam flows through the superheat coil (E-103), bypasses the cathode recycle gas cooler (E-200), and is vented to the atmosphere. A small flow of steam continues to be vented to warm the piping up to the steam flow control valve.
7. As pressure in the steam drum rises to 100 psig, steam is exported automatically from the plant.
8. The boiler blowdown valve is modulated to attain a boiler blowdown based on the quantity of boiler feed water sent to the HRSG. Additional blowdown is taken, as necessary, to avoid boiling in the economizer.

Starting the Turbocharger

The turbocharger (K-201) is started using the turbocharger start-up blower (K-202) to supply preheated air to the expander side of the turbocharger. The turbocharger must be started before the fuel cell system can be heated above 1000°F by burning reformed natural gas. The procedures are as follows:

1. The turbocharger start-up blower (K-200) is started. The blower operates at constant speed.
2. Air is drawn into K-202 from the atmosphere. The compressed air discharging from K-202 flows into the turbocharger air bypass line.
3. The air flows through the air preheat coil (E-104) and then to the expander side of K-201.
4. The air entering the expander side of K-201 impinges on the blades, causing the turbocharger shaft to rotate.
5. The firing rate of the HRSG start-up heater (F-300) is adjusted to preheat the air exiting E-104 at a predetermined ramp rate to a maximum temperature of 1100°F.
6. As the expander side inlet temperature increases, the expander side inlet pressure also increases as well as the turbocharger shaft rotational speed.
7. As the turbocharger shaft gains rotational speed, the pressure at the outlet of the compressor side of K-201 increases.
8. As the K-201 compressor side discharge pressure increases, compressed air from K-201 is joined together with that from K-202 to be heated in E-104 prior to passing through the turbocharger expander.

9. When the expander side inlet temperature is at 850°F, the compressor outlet pressure is approximately 5 psig and the turbocharger is at approximately 50% of its normal speed. At this point, the compressor flow rate is sufficient to maintain the turbocharger operating at steady state.
10. When the turbocharger compressor side discharge pressure exceeds the maximum outlet pressure of K-202 of about 6 psig, all the flow from K-202 is vented to the atmosphere from a pressure safety valve at the K-202 discharge.
11. The K-202 blower is stopped.
12. As the preheated air temperature is ramped from 850°F to approximately 1100°F, the turbocharger expander inlet pressure increases and the turbocharger speed increases from 50% to 100% of its normal speed.
13. The expander inlet block valve is opened when the turbocharger expander inlet pressure equals that upstream of the expander inlet block valve.
14. The expander inlet pressure (and speed) is maintained at its designed operating point by modulating the waste gate. This operating point corresponds to a compressor discharge pressure of 46 psia.

Heating the System by Burning Reformed Natural Gas

After the reformer combustor has been heated to greater than 1000°F, and the fuel cell has been heated to greater than 1100°F, the fuel cell system can transition to heating up with natural gas. Before natural gas can be fed to the system, combustion of the safegas is initiated by opening the isolation valve in the anode exhaust line. Reducing safegas combusts in the combustion chamber with turbocharger compressor exhaust air. When combustion is sufficiently stable, steam first and then natural gas are fed to the reformer while reducing safegas is withdrawn. The reformat product passes unconverted through the anode and is burned in the reformer combustion chamber. The fuel cell must be above 1100°F (adequately above the 914°F end point for the melt region) at this time to ensure that the fuel cell wet seals properly isolate the anode gas from the cathode and the pressure vessel. The procedures are as follows:

1. The cathode recycle control valves are modulating to meet the setpoint for the reformer combustor temperature.
2. The block valve at the compressor discharge side of turbocharger (K-201) is opened to permit air to be introduced to the process.

3. Air flow to the process is started at a predetermined rate. The temperature of the mixed oxidant fed to the combustor is controlled to a predetermined setpoint by controlling the amount of air introduced upstream of the combustor. The minimum oxidant inlet temperature is 932°F.
4. Once the concentration of oxygen on the cathode side has reached an acceptable level, the isolation valve in the anode exhaust line is opened to permit reducing safegas (60 lb/hr) to be introduced to the combustor. Immediately afterwards, the anode vent valve will close because it is interlocked with the block valve. A friction valve in the anode exhaust line then modulates to establish the anode outlet pressure to be equal to the cathode inlet pressure. This results in the anode manifold pressures being higher than the cathode manifold pressures. Therefore, any failures in the fuel cell wet seals will result in anode gas leaking to the cathode. The reformer maximum combustor temperature increases.
5. The combustion outlet temperature is allowed to increase above 1020°F by adjusting the reformer heater set points to 1130°F. This should result in a reforming temperature near 1000°F.
6. The vent valve upstream of the superheated steam control valve is closed. Superheated steam is then introduced to the reformer at a predetermined rate in preparation for introducing fuel to the reformer and to heat up the anode inlet piping.
7. When the temperature of the piping at the reformer outlet reaches 1020°F and the cathode outlet reaches 1188°F, natural gas is introduced slowly to the natural gas preheat coil (E-102) at a predetermined rate (7.5% of full capacity). As the natural gas is introduced, the nitrogen introduced upstream of the reformer and the hydrogen introduced upstream of the anode are withdrawn until the flow is stopped. The steam flow is controlled to provide a predetermined steam to carbon ratio (9.0, molar) based on natural gas flow.
8. The natural gas/steam passes through the reformer and the anode to the combustor.
9. The maximum combustor temperature is ramped up at a maximum rate of 45°F/hr by decreasing the flow rate of cathode recycle gas going to the combustor.
10. The flow rate of superheated steam is regulated to maintain a minimum 3-to-1 steam-to-carbon molar ratio. Excess steam is added to maintain a maximum 4-to-1 mixed oxidant-to-anode effluent molar ratio.
11. The anode preheater (P-203) output is ramped at 10°F/hr to achieve a 1250°F temperature at the piping at the anode inlet.

12. The fuel cell is ramped at a maximum rate of 10°F/hr with additional heat input from the anode preheater, the fuel cell guard heaters and the cathode start-up heater until the cathode outlet temperature is 1200°F. The fuel cell is then maintained at 1200°F.
13. The reformer combustor ramps up until the maximum combustor temperature is 1460°F and the fuel cell is at 1200°F (as measured by the cathode outlet temperature). At this point the reformer flue gas temperature is approximately 1250°F. The reformer combustor is then maintained at 1460°F.
14. The HRSG start-up burner (F-300) continues to supply heat to the HRSG to produce superheated steam and to maintain the turbocharger at its operating speed.
15. The cathode start-up heater (P-202) continues to heat the cathode recycle gas in order to maintain the desired cathode outlet temperature. The maximum exit temperature for P-202 should not exceed 1275°F or the electric heater will trip off.
16. The plant is now at standby condition.

Switching to Load

1. The inverter (P-204) is synchronized and connected to the 480 volt bus and a minimum load of 98 amps is initiated.
2. The heating value of the anode effluent gas going to the combustor decreases due to anode fuel consumption.
3. The flow rate of mixed oxidant is reduced automatically to maintain a 1460°F maximum reformer combustor temperature. The mixed oxidant temperature remains at its initial setpoint.
4. The combustor flue gas temperature drops from near 1250°F to approximately 1130°F.
5. The temperature controller at the exit of the reformer combustion chamber regulates the fuel flow rate to achieve a constant 1130°F combustor flue gas temperature.
6. The steam flow rate is adjusted to maintain a 3-to-1 steam-to-carbon molar ratio plus any additional steam to maintain the molar ratio of mixed oxidant to anode effluent less than 4.0.
7. The air flow to the process is set to achieve a minimum 70% excess air based on the fuel flow to the process.

8. The anode preheater (P-203) output is controlled to achieve a 1250°F anode inlet temperature. The maximum exit temperature for P-203 should not exceed 1300°F.
9. The cathode start-up heater (P-202) continues to heat the cathode recycle gas in order to attain a cathode outlet temperature of 1290°F. The maximum exit temperature for P-202 should not exceed 1275°F so as to insure that the cathode inlet temperature remains less than 1300°F.
10. The HRSG start-up burner (F-300) continues to supply heat to the HRSG to produce superheated steam and to maintain the turbocharger at its operating speed. The demand signal for F-300 is from the temperature controller at the discharge of the air preheat coil (E-104).

Ramping to Minimum Self-sustained Load

1. The current drawn for the inverter (P-204) is increased at a predetermined rate.
2. The heating value of the anode effluent gas going to the combustor decreases due to increased anode fuel consumption. This effect is minimized by using feed-forward control of the fuel flow rate to the process based on the measured rate of change of current.
3. The steam flow rate is adjusted to maintain 3-to-1 steam-to-carbon molar ratio plus any additional steam to maintain the molar ratio of mixed oxidant to anode effluent less than 4.0.
4. The flow rate of mixed oxidant is controlled to maintain a 1460°F maximum reformer combustor temperature. The mixed oxidant temperature remains at its initial setpoint.
5. The temperature controller at the exit of the reformer combustion chamber regulates the fuel flow rate to achieve a constant 1130°F combustor flue gas temperature.
6. The air flow to the process is set to achieve a minimum 70% excess air based on the fuel flow to the process.
7. The anode preheater (P-203) output is controlled to achieve a 1250°F anode inlet temperature. The maximum exit temperature for P-203 should not exceed 1300°F.
8. The cathode start-up heater (P-202) continues to heat the cathode recycle gas in order to attain a cathode outlet temperature of 1290°F. The maximum exit temperature for P-202 should not exceed 1275°F so as to insure that the cathode inlet temperature remains less

than 1300°F (Note that the cathode recycle gas temperature increases approximately 25°F through the K-200 recycle blower.)

9. As the load increases, the heat input required from the HRSG start-up burner (F-300) is reduced. The heat demand is the greater of two signals: 1) the temperature controller on the discharge of the air preheat coil (E-104) and 2) the temperature controller for the superheated steam exiting the HRSG. The setpoint for the air preheat temperature is calculated based on the measured air flow to the process.
10. As the load increases, the heat input required from the cathode start-up heater (P-202) is reduced in order to maintain the 1290°F cathode outlet temperature.
11. Steam is introduced into the cathode recycle cooler (E-200) when the heat input from the cathode start-up heater (P-202) is no longer required to maintain a 1290°F cathode outlet temperature. The setpoint for the steam temperature controller downstream of E-200 is slowly increased from 700°F to 1000°F, thereby slowly increasing the flow rate of steam to E-200.
12. At the same time, the setpoint for the temperature controller downstream of the steam preheat coil (E-103) is slowly decreased from 700°F to 350°F, thereby shifting the steam superheating duty almost completely to E-200. The temperature of 100 psig saturated steam is 338°F; therefore, a 350°F discharge temperature for E-103 insures a minimum flow of steam through E-103 at all times.
13. The power output from the plant continues ramping up until heat input is no longer required by the cathode start-up heater (P-202) and the HRSG start-up burner (F-300). The plant is then at a point of minimum self-sustained load.

Minimum Self-sustained Load to Full Load

1. The current drawn for the inverter (P-204) is increased at a predetermined rate.
2. The steam responds to maintain the following temperatures:
 - * 1460°F maximum combustion temperature
 - * 932°F mixed oxidant temperature (minimum)
 - * 1130°F combustor outlet temperature
 - * 1290°F cathode outlet temperature
 - * 1000°F superheated steam temperature
 - * 1250°F anode inlet temperature

3. The maximum reformer combustion temperature is controlled by varying the flow rate of mixed oxidant sent to the combustor.
4. The mixed oxidant temperature is maintained by varying the amount of compressed air introduced upstream of the reformer combustor.
5. The combustor outlet temperature is maintained by varying the reformer feed rate.
6. The cathode outlet temperature is maintained by 1) controlling the mixed oxidant temperature going to the reformer and 2) controlling the amount of compressed air sent to the cathode.
7. The superheated steam temperature is maintained by varying the amount of steam introduced to the cathode recycle gas cooler (E-200).
8. The anode inlet temperature is maintained through heat input from the anode preheater (P-203).
9. Heat input from the fuel cell guard heaters (P-201 A-D) is no longer required. The heaters remain active to provide back-up heating when the plant is switched off load or when the stack can not maintain its temperature for any reason.
10. Heat input from the reformer heaters (P-200 A-E, P-205 A-E, and P-206 A-D) is no longer required. The heaters remain active to provide back-up heating when the plant is switched off load or when the reformer can not maintain its temperature for any reason.
11. Heat input from the cathode start-up heater (P-202) is no longer required. This heater remains active and provides back-up heating in the event that the cathode outlet temperature can not be maintained.

Process and Control (PAC) Testing without Fuel Cell Installation

Prior to integrating the balance of plant equipment and systems with the fuel cell, it is necessary to have them tested and accepted as suitable for full scale operation. In some instances it is not feasible to simulate full load conditions without an operating fuel cell. Therefore it is necessary that some acceptance tests be performed under partial process flow/load conditions. Final acceptance was deferred until the fuel cell was integrated with the rest of the plant and full load conditions were established.

Prior to initiating PAC testing the following prerequisites/conditions must be met to ensure that comprehensive and meaningful testing can be accomplished.

1. Plant construction activities for the system/equipment under test are completed.
2. Precommissioning activities as outlined in the Plant Operating Manual are completed to the extent they are required for system operation.
3. Plant utilities and permanent power are available to support the testing operation.
4. Required plant instrumentation is installed and calibrated.
5. Plant Distributed Control System (DCS) is configured to control the system under the stipulated test conditions. Shift Log Reports are able to be printed in either automatic or manual mode. The DCS-Data Acquisition System (DAS) and the Gas Chromatograph (GC)-DAS are operational and provide necessary data downloads.
6. Necessary insulation work has been completed to satisfactorily perform the required heat/mass balance for acceptance test.
7. Temporary instrumentation, if required, is installed and calibrated.
8. Contracted services for sample collection and analysis is in place.

The following test conditions were used in support of the PAC tests. Tests were conducted under steady state conditions. Plant systems were at a normal (i.e, non-transient) operating state where feasible. During the PAC tests, it was necessary to override those trip alarms and/or pre-alarms which are interlocked with fuel cell related logic since some of these tests simulate on-load conditions while the plant logic is actually in the natural gas heat-up state. Otherwise, all system alarms would be activated.

Test duration for each test condition is outlined under the test protocol. Where required, test duration was extended in order to collect additional data. Process parameters measured on-line (such as flow, pressure, temperature) were continuously trended and posted to the DCS-DAS files at 15 minute intervals. Sufficient data was collected for statistical significance. The sequence in which gas samples are automatically taken by the GC was specified by the operator.

PAC testing procedures for the following equipment/subsystems are described below:

1. Heat Recovery Steam Generation (HRSG)
2. Desulfurization Performance
3. Reformer
4. Recycle Blower K-200
5. Cathode Recycle Cooler
6. Cathode Startup Heater

7. Anode Preheater
8. Inverter Load Test w/o Fuel Cell (optional) and Remote/Manual Control with DCS
9. DCS and Pneumatic Controller Loop Stability
10. Gas Chromatograph (GC) Analysis and Flow Measurement Accuracy

Heat Recovery Steam Generation (HRSG) PAC Test

The HRSG PAC Test covers the HRSG operating on startup burners, on burners with turbocharger, on burners with reformer in service and the HRSG at full load with no burners. The PAC Test for each of these operating conditions is described below.

HRSG on Startup Burners

The objective of this test was to verify that the steam production rate (FI-206) equals the BFW flow (FI-201) when there is no blowdown. The blowdown rate was set at 0.5% of the BFW rate. A gas temperature profile and heated fluids temperature profile was established in order to determine the approach temperatures at each coil and to determine the individual coil exit temperatures. The system was operated to detect air in-leakage. HRSG burner (F-300) and coils (E-102, E-103, and E-104) duties were calculated to establish baseline data. The nameplate duties of E-100 (economizer) and E-101 (boiler) and export steam capabilities and PAC test steam line capabilities were verified in order to generate nameplate steam from boiler E-101).

HRSG on Burners w/Turbocharger

The objective of this test was to verify that the turbocharger runs and to determine heat losses in the pre-heated air line from the T/C expander exhaust to the HRSG inlet. Heat losses from pre-heated air reduce the overall HRSG duty. A gas temperature profile and heated fluids temperature profile was established in order to determine the approach temperatures at each coil, and to determine the individual coil exit temperatures. HRSG burner (F-300) and coils (E-102, E-103, E-104) duties were calculated to establish baseline data. The compression ratio and air flow rate from the turbocharger compressor was also determined.

HRSG on Burners w/Reformer in Service

The objective of this test was to determine/plot gas temperature profile and heated fluids temperature profile in order to establish the approach temperatures at each coil. HRSG burner (F-300) and coils (E-102, E-103, E-104) duties were calculated to establish baseline data. The compression ratio and air flow rate from the turbocharger compressor was also determined.

HRSO @ Full Load--No Burners

This test was conducted when the Fuel Cell was on line and the cathode exhaust was recycled as flue gas to the HRSO.

The objective of this work was to determine/plot gas temperature profile and heated fluids temperature profile in order to establish the approach temperatures at each coil. Gas flow and flue gas composition were established through the HRSO as baseline emissions from the fuel cell system. Measurements of CO, NO_x, and SO_x from the fuel cell system were made per EPA guide line 40 CFR 60, Emissions from Stationary source using natural gas as primary fuel. The overall heat balance for the HRSO - heat input through the recycle gas, heat loss through the stack, and heat transferred to individual heated fluid was determined.

Desulfurization Performance

Desulfurization performance tests were performed after the beds were pressurized with natural gas and prior to the reformer. The objective of this test was to verify desulfurization effectiveness of each bed at full capacity. Using GC-ESD techniques, sulfur component concentrations were measured in the natural gas feed to the plant. Measurements of the effluent sulfur content by sulfur containing component (i.e., mercaptans, thiophanes, etc.) for each bed configuration were made and the effectiveness of the Tospix-II catalyst was effective. A measurement of the inter-bed sulfur content by sulfur containing component (i.e., mercaptans, thiophanes, etc.) was made. Verification tests showed that the calgon activated carbon is effective. Measurements of the inter-bed sulfur content by sulfur containing component (i.e., mercaptans, thiophanes, etc.) were also made.

Reformer PAC Test

The reformer was tested at four different capacity levels as described below.

Reformer at 24% Capacity (Case 1):

The objective of this test was to verify desulfurization at the turn-down of this test. Methane conversion was verified by comparison of gas analyses for the reformer feed and effluent streams. Inlet gas composition was calculated from the (assumed) constant desulfurized natural gas assay (taken once just before Reformer PAC test case 1 only for a verification of desulfurization). Reformate gas composition was recorded immediately after the start of reformate flow (through the on-line GC) and once again at the end of each Reformer PAC test. One grab sample of the reformate was taken (during Reformer PAC test case 1 only) from the GC sample conditioner and analyzed for comparison with the on-line GC results. The test verified that reformer outlet temperature can be maintained using a simulated anode effluent prepared by diluting the fuel to the reformer combustion chamber with purge nitrogen and/or steam in accordance with the proportions of

nitrogen to fuel shown on the test table. Baseline data was gathered on the reformer temperatures, extinction of combustibles, overall air to fuel ratio and heater performance.

Reformer at 50% Capacity (Case 2):

Same as Test Case 1. All verification was performed when reformer load was at 50% of the capacity.

Reformer at 75% Capacity (Case 3):

Same as Test Case 1. All verification was performed when reformer load was at 75% of the capacity.

Reformer at 100% Capacity (Case 4):

Same as Test Case 1. All verification was performed when reformer load was at 100% of the capacity.

Recycle Blower K-200

The recycle blower was tested during the blower seal test for operation and blower integrity. Additional testing took place during PAC test.

Miscellaneous Equipment Performance

Cathode Recycle Cooler E-200

Performance testing verified that adequate steam superheat at full capacity was available. The tests also verified that adequate cooling of cathode inlet gas was available.

Cathode Startup Heater P-202

Performance testing verified that cathode start-up heaters heat output was adequate. The tests also verified that heater output was adequate to heat the system.

Anode Preheater P-203

Performance testing verified that anode preheater output was in accordance with duty specifications. The tests also verified that heater output was adequate to heat the system.

Inverter Pre-commissioning and Acceptance Test Protocol

The Inverter Test table provides the outline for the required pre-commissioning verification and final acceptance test protocol for the inverter package. Items 1 through 8 are pre-commissioning activities that were performed without special setup and with standard multi meter and a scope. Meters and relays were verified for calibration and settings. Step 7 was performed under simulated conditions with a d.c. power source and an a.c. resistor load bank. Once inverter basic checks were verified, the inverter was connected to the grid. A test procedure by InverPower, included with the subsystems verification documentation, was used to prove step 7, 8 and 10 -inverter performance.

Inverter Test Table Steps

The following test steps were followed:

1. Tie-in line to 4.16 kV substation and grid protective relay settings and checkouts
2. WH Metering Station calibration and checkout
3. Generation MCC inspection, wiring checkout, relay trip settings, verification
4. Generation MCC and 4.16 kV X-fmr cable meggering, grounding checks
5. P-204 inverter inspection, wiring checks and I/O cable Hi-Pot
6. P-204 functional tests and local controls/diagnostics (logic power supply check, Logic Board settings, Gating pulse check)
7. Inverter test with Rectified d.c. Source (optional)
 - a.) Test with Inverter Output to resistive load bank, isolated from the grid
 - b.) Test with inverter connected to Grid
8. Inverter control remote/manual with DCS under conditions of Step 7 a) and b)
9. Inverter with fuel cell - operational control and protection verification
10. Inverter Performance Test

DCS and Pneumatic Controller Loop Stability

The objective of this test was to verify that DCS-based controller setpoints are correct (i.e., in manual, auto/cascade and cotracking modes) for each operating state (except on-load) and during the reformer acceptance tests. The tests verified that the valves under DCS control were configured

correctly. Tuning constants (gain, integral and derivative) for each DCS-based controller were determined. System response to anode overpressure (i.e., reducing cathode pressure during Heat up to 1150°F) were verified and the tuning constants for PDIC-551 were determined. System response to cathode overpressure (i.e., reducing anode pressure during Heat up to 1150°F) were verified and the tuning constants for PDIC-552 were determined. DCS-based lag constants were verified that they were compatible with real process lags (i.e., valve positioning lags, thermal lags, instrumentation lags (thermocouples and gas analyzers). Controller rate limiters, de-limiters and action (i.e., forward/reverse) were adjusted and their performance was verified.

GC-Based Composition Analysis and Flow Measurement Accuracy Tests

The objective of these tests were to verify that the gas chromatography analyzer unit's on-line composition analysis and the flow instrument measurements were both accurate. The range of flow rates over which the flow instrument response gives accurate measurements was determined. The accuracy of the GC was also determined. Testing was used to determine quirks (instrument tubing/sequence valve leaks) and bugs (electronic malfunctions in the instrument transmitter or the DCS).

Task 6 Demonstration Plant Testing

Objective: Prepare plant operating manuals, test plans, execute test programs, and document the PDT demonstration plant's performance.

Subtask 6.1 Operating and Maintenance Manuals

Discussion: The operating and maintenance manuals for the demonstration plant were completed and delivered to GRI on December 14, 1995 as this was a task funded by GRI.

Subtask 6.2 Operating Test Plan

Discussion: The operating test plan for the demonstration plant was completed and a topical report on the subject was delivered to DOE and GRI on March 20, 1996.

Subtask 6.3 Performance Testing

Discussion: Performance testing of the demonstration plant was completed on December 22, 1996 and normal test operation was initiated on January 10, 1997.

Subtask 6.4 Emissions Testing

Discussion: Emission testing of the demonstration plant was not conducted because the plant was not operated for a sustained period of time under conditions specified in the emission testing

program. Also, the cooperative agreement completion date did not allow for additional plant modifications and operation. Emission testing is scheduled to take place when the plant is restarted under a plant component verification testing program which is beyond the scope of this cooperative agreement.

Subtask 6.5 Analysis and Test Reporting

Discussion: The results of the demonstration plant performance testing is described below.

Operating History

The history of stack temperature, current, voltage and power during the demonstration are summarized in Figures 13 - 16. The initial heatup of the stack to operating temperature of about 1200 °F was started at noon of January 10, 1997 (as zero operating hour) and the cooldown of the stack was complete at about 17:00 of May 21, 1997 (3150 operating hours). By defining "hot" hour as the operating time in which Cell 5 temperature exceeded 1100 °F, a total of 2360 "hot" hours was achieved during this demonstration test. During this period, the power plant produced a total of 160 MWH electricity and exported 346,000 lbs of 100 psig steam.

Polarization Curve

Figure 17 shows the stack performance characteristics during the first 1575 hours of operation. The highest load achieved during the demonstration test was 1200 amps operation with d.c. power output of 210 kW. Based these actual operating data, the projected d.c. power output at the nameplate load of 1822 amps would be about 267 kW. This projected power output is about 81% of the design target of 330 d.c. kW, which is to result in 250 kW a.c. power.

Cell Voltage Distribution

The cell voltage distribution for OCV and 1200 amps load during steady-state operation are shown in Figure 18. The end-plate effect can be seen in the cell pack containing cells 245 to 250. Excluding this cell pack, the standard deviations for the OCV and 1200 amps operation are 9 and 23 mV, respectively.

Cathode Temperature Profile

During the test, cathode inlet and outlet temperatures in the pipes were monitored and controlled to avoid excessive temperature gradient in the stack. The goal was to keep the cathode inlet temperature above 1100° F and cathode outlet temperature below 1290° F. Figure 19 summarizes the history of cathode temperatures during the test.

Cathode Pressure Drop

According to plant design, cathode pressure drop was not to exceed 30 iwc during operation to protect the wet seal of the stack, which was operated in a counter-current mode. Most of the time during the test, the vessel pressure was one or two inches above the anode and cathode manifold pressures, and the anode outlet manifold pressure was maintained slightly above the cathode inlet

Figure 13 History of Stack Temperature at NAS Miramar

Figure 14 History of Stack DC Current at NAS Miramar

Figure 15 History of Stack DC Voltage at NAS Miramar

Figure 16 History of Stack DC Power at NAS Miramar

Figure 17 Polarization Curve for PDT-1 Stack at NAS Miramar

Figure 18 Voltage Distributions for PDT-1 Stack at Miramar

Figure 19 History of Cathode Temperatures at NAS Miramar

pressure. Because the anode pressure drop was very small compared (less than 3 iwc) to the cathode pressure drop, the wet seal between the anode inlet and cathode outlet manifolds could be broken if cathode pressure drop became too large. The same is true for the wet seal that separates the cathode from the perimeter of the stack.

Figure 20 shows the history of the cathode pressure drop. During operation, it was found that the cathode pressure drop was much greater than anticipated. Figure 21 shows the actual pressure drop data for the PDT-1 Stack in the ATF and during the demonstration testing, the actual pressure drop data for the stack at the Unocal demonstration, and the anticipated pressure-drop correlation line based on previous 20 kW stack data. As shown in these data, the cathode pressure drop for the stack at the ATF were consistent with the anticipated correlation line. The pressure drop data obtained during the Miramar demonstration were twice or three times the anticipated pressure drop. Based on the correlation line, at the nameplate load of 1822 amps, the pressure drop should only be about 20 iwc at the design flow of 402 standard liter per minute (slpm) per cell. The actual data from the Miramar demonstration shows 30 iwc at a flow of 320 slpm/cell (about 1200 amps).

It was also found that the cathode pressure increased with operating time. While maintaining the highest allowable cathode pressure drop of 35 iwc, the amount of cathode inlet flow rate got progressively smaller. Cathode flow being the primary means of removing heat from the stack, its reduced flow led to a higher temperature differential between the cathode inlet and outlet. In order to keep this temperature differential within proper limits, the maximum permitted stack load became smaller with operating time.

To quantify the extent of increase in cathode pressure drop with operating time, the data shown in Figure 21 were further delineated in Figure 22 and Table 7. By segregating the data in narrow flow ranges, the pattern of increase in cathode pressure drop can be seen clearly. As shown in Figure 23, the ΔP s increased with processing time from OCV to 700 amps load. A linear regression of these data was performed with respect to time. Based on the regression correlations, over a one-thousand-hour period, the increase in ΔP for the 430-485 amp load was about 13 iwc, and for the 520-710 amp load was about 14.7 iwc. Although there are insufficient data for the 1185-1200 amp load for regression analysis, its ΔP increase was expected to be higher.

Stack Performance

Based on the actual power output, the projected power output at the nameplate load of 1822 amps was shown to be about 80% of the plant design power output. For the plant design, the average cell voltages at different stack load were predicted using the PSI model, using 0.57 Ohm cm^2 for cell internal resistance. To verify these design voltages, several bench-scale tests (4" x 4" active area, isothermal operation) were run at the conditions used in the plant design. As shown in Figure 24, the bench-scale results match the PSI model predictions.

Figure 20 History of Cathode Pressure Drop for PDT-1 Stack at NAS Miramar

Figure 21 Cathode Pressure Drop Correlation

Figure 22 Cathode Pressure Drop vs. Flow for PDT-1 Stack at Miramar

Table 7 Regression of Cathode Pressure Drop Data for PDT-1 Stack at NAS Miramar

			Cath. DP	Cathode DP @ 1202 F and 44.7 psia				
			Adjusted to	OCV &			OCV	
Stack			1202 F &	Flow	~140-285	Regression Output:		Regressed
Load,Amp	Op. Hr	FI-524,acfm	44.7 psia	slpm/min	slpm/cell	Constant		DP
0	300	2203	3.69	146.49	3.69	Std Err of Y Est	3.27001	4.34
0	615	1539	6.05	142.05	6.05	R Squared	0.54885	5.46
0	655.5	1875	6.72	177.39	6.72	No. of Observations	6	5.60
0	845.5	2480	6.15	205.30	6.15	Degrees of Freedom	4	6.28
0	850	2053	5.80	170.10	5.80			6.30
0	854.3	1767	5.89	147.51	5.89	X Coefficient(s)	0.00356	6.31
						Std Err of Coef.	0.00161	
			Cath. DP	Cathode DP @ 1202 F and 44.7 psia				
			Adjusted to	~130-290 A &			~130-290 A	
Stack			1202 F &	Flow	~90-260	Regression Output:		Regressed
Load,Amp	Op. Hr	FI-524,acfm	44.7 psia	slpm/min	slpm/cell	Constant		DP
264.4	432	906.14	5.10	86.66	5.10	Std Err of Y Est	0.30991	5.96
267.13	432.3	906.14	5.38	87.37	5.38	R Squared	0.72394	5.96
267.01	432.5	906.14	5.43	87.57	5.43	No. of Observations	9	5.96
264.4	432.8	906.14	6.84	88.06	6.84	Degrees of Freedom	7	5.97
286.29	495	2124	7.59	201.97	7.59			6.78
287.16	495.3	2604	7.52	247.28	7.52	X Coefficient(s)	0.01307	6.78
161.93	539	2488	7.24	235.55	7.24	Std Err of Coef.	0.00305	7.36
161.05	540	2734	7.38	260.18	7.38			7.37
131.28	678	2268	8.82	244.93	8.82			9.17
			Cath. DP	Cathode DP @ 1202 F and 44.7 psia				
			Adjusted to	~430-485 A &			~430-485 A	
Stack			1202 F &	Flow	~200-260	Regression Output:		Regressed
Load,Amp	Op. Hr	FI-524,acfm	44.7 psia	slpm/min	slpm/cell	Constant		DP
431.67	435	2069	6.52	201.21	6.52	Std Err of Y Est	4.15223	6.36
	800					R Squared	0.99318	8.22
467	1291	2517	10.35	249.82	10.35	No. of Observations	3	10.72
	1500					Degrees of Freedom	1	11.78
485	1945	2645	14.25	255.06	14.25			14.04
						X Coefficient(s)	0.00509	
						Std Err of Coef.	0.00042	
			Cath. DP	Cathode DP @ 1202 F and 44.7 psia				
			Adjusted to	~520-710A &			~520-710A	
Stack			1202 F &	Flow	~200-310	Regression Output:		Regressed
Load,Amp	Op. Hr	FI-524,acfm	44.7 psia	slpm/min	slpm/cell	Constant		DP
521.62	830	1947	9.57	198.33	9.57	Std Err of Y Est	-0.57932	11.62
659.2	1005	2824	14.47	276.02	14.47	R Squared	0.96094	14.20
640.83	1006	2838	14.13	274.10	14.13	No. of Observations	6	14.21
640.83	1007	2875	14.43	279.15	14.43	Degrees of Freedom	4	14.23
709.23	1070	3007	17.26	312.01	17.26			15.15
697	2074	2385	29.45	243.64	29.45	X Coefficient(s)	0.0147	29.91
						Std Err of Coef.	0.00148	

Figure 23 Cathode Pressure Drop vs. Time for PDT-1 Stack at NAS Miramar

Figure 24 PDT-1 Stack Design vs. Bench-Scale Cell Voltages

* Bench-Scale test results adjusted to cell resistance 0.57 Ohm.cm². BS tests conducted at conditions matching the conditions in Bechtel's process Description for Miramar Power Plant Design.

The plant operating conditions were different from the design conditions. The steam/carbon ratio was higher, system pressure was lower, and oxidant concentration was lower when compared to design conditions. Therefore, the PSI model was run to obtain predictions at the actual operating conditions and compare them to the actual performance, as shown in Figure 25. The actual performance is always lower than the predicted by 30 to 90 mV. The difference is attributed to scale-up effects.

Previously the best cells in a 20-kW stack were always very close to their "twin" in bench-scale cell performance, but the average cell in any stack always had a lower performance. The lower performance for the 20 kW stacks can be ascribed to intrinsic differences (e.g., physical differences) between the 20-kW stacks and bench-scale cells, such as non-isothermal conditions vs. quasi-isothermal conditions, non-uniform gas distribution vs. quasi-uniform gas distribution, presence vs. absence of end plates effects, difference in electrolyte distribution, etc.. One should also expect that the stack performance would be even lower when the number of cells increases from 20 to 250 in the stack. To some extent, the use of intermediate plates have reduced the scale-up effects, but not completely. To check whether the scale-up effect was already present during the acceptance test of the stack at the ATF, the PSI model was run. As seen in Figure 26, the PSI model predictions (infers bench-scale cell performance) were indeed higher than the actual performance by 60 to 100 mV.

In addition to the scale-up effects, the higher than expected cathode pressure drop also prevented the plant from delivering the required compressed air to reach the design flow for the name plate load of 1822 amps. This also caused the oxygen level in the cathode inlet stream to be lower than the design values and contributed to even lower stack voltages. It appears to be reasonable to conclude that the power plant could only reach 80% of the name plate power output is primarily caused by the scale-up effects and high cathode pressure drop problem.

Fuel Cell Material Balance

Figure 27 shows available instrumentation for collecting the data to calculate the stack material balances. Because there was no flow meter to measure the anode inlet (AI) flow rate, it was calculated assuming 100% carbon balance and 100% mass balance for the IHI reformer. Knowing the amount of steam (F-301) and natural gas (FI-125) fed to the IHI Reformer and the GC analysis of the reformer product (AP-518), the anode outlet stream was defined. However, due to instrumentation problems, the water content in the reformer product could not be determined. To estimate the amount of water content, the elemental distribution (i.e., elemental mass flow rates) of carbon, hydrogen, and oxygen of the combined feed stream was calculated first. Assuming the distribution of non-water component measured by GC was correct, the percent of water was then adjusted iteratively until the elemental carbon mass flow rate of the reformer product stream matched that of the combined reformer feed stream. Table 8 shows an example of this calculation. The fact that the feed/product ratios for elemental hydrogen and carbon are very close to 1.0 implies this

Figure 25 Predicted vs. Actual Average PDT-1 Stack Cell Voltages at NAS Miramar

Figure 26 Predicted vs. Actual Average PDT-1 Stack Cell Voltages at ATF

75

Table 8 Example of Estimating Anode Inlet Flow and Composition from Reformer Operation

Rfmr Prod (Anode Inlet) GC Date						* Avg for 1132 to 1204 operating Hours			
This spreadsheet backcalculates %H2O in reformer product to match:									
1) total mass (natural gas + steam) into the reforming chamber,									
2) total carbon feeding into the reforming chamber (i.e., 100% carbon balance)									
The calculations in this spreadsheet assume the distribution of non-water components as measured by GC to be correct.									
These calculations are iterative. The %H2O in reformer product (adj. %, wet) is iterated until its corresponding carbon matches the total amount of carbon fed into the reformer. These iterative calculations can be performed automatically by pressing CTRL, SHIFT and G keys at the same time (after entering necessary data) .									
STREAM		Purge Gas	Nat. Gas	Steam	NG&Purge	NG&Stm &Purge	Rfmr Prod Raw %	Rfmr Prod Norm. %	Rfmr Prod Adj. %, wet
	MW	mole%	mole%	mole%	Mole%	Mole%	wet	dry	Adj. %, wet
CO2	44.01	0.00	0.45	0.00	0.45	0.08	11.58	11.69	7.21
H2O	18.02	0.00	0.00	100.00	0.00	81.19	0.93	0	38.29
O2	32.00	0.00	0.00	0.00	0.00	0.00	0	0.00	0.00
N2	28.01	100.00	2.00	0.00	2.00	0.38	0.15	0.15	0.09
CH4	16.04	0.00	93.02	0.00	93.02	17.49	0.57	0.58	0.36
CO	28.01	0.00	0.00	0.00	0.00	0.00	10.42	10.52	6.49
H2	2.02	0.00	0.00	0.00	0.00	0.00	76.35	77.07	47.55
C2H6	28.05	0.00	3.67	0.00	3.67	0.69	0	0.00	0.00
C3H8	44.10	0.00	0.64	0.00	0.64	0.12	0	0.00	0.00
N-C4H10	58.12	0.00	0.22	0.00	0.22	0.04	0	0.00	0.00
Total		100.00	100.00	100.00	100.00	100.00	100	100.00	100.00
Avg MW		28.0134	17.1209	18.0152	17.120946	17.847		9.7785	12.9327
		(FI-821)	(FI-125)	(FI-301)					
Flow, SCFM		0							
Flow, lb/hr		0.00	79.35	360.5	79.35	439.85			439.850
Flow, lbmole/hr		0.0000	4.6347	20.0109	4.6347	24.6456			34.0108
							Checking		
Species, lbmole/hr							Adj. Prod/Feed		
C		0.0000	4.8020	0.0000		4.7811	1.0000		4.7811
H		0.0000	18.6045	40.0218		58.6263	1.0043		58.8787
O		0.0000	0.0417	20.0109		20.0526	1.0042		20.1376
							0.3428		

method of estimating the anode feed flow rate and composition is very accurate. As an additional check, the expected reformer composition based on actual reforming conditions was also calculated. As shown in Table 9, there is an excellent agreement between the expected composition and the composition estimated from the 100% carbon balance and 100% mass balance assumption.

The cathode inlet (CI) flow rate and composition were measured by FI-524 and AP-520, respectively. Because the vessel purge stream (FI-506) flow rate was very small compared to cathode inlet flow rate, it was ignored in the material balance calculations. Based on 100% WSE assumption and information on anode and cathode inlets, the anode outlet (AO) and cathode outlet (CO) flow rates and composition can be calculated. Note that because CO flow was not directly measured, it was therefore estimated from the 100% WSE CO flow and measured GC (AP-521) by assuming 100% nitrogen balance. There was also an instrumentation problem for determining the water content in the AO stream. Therefore, based on the water shift reaction, the percent water in the measured AO GC was adjusted such that the sum of the adjusted percent H₂O and adjusted percent CO₂ is equal to the sum of expected sum of percent H₂O and percent CO₂ based on 100% WSE. An example of such calculations is shown in Table 10.

The material balance calculations were made for periods in which steady-state operation was achieved and the required instrument readings were stable. A summary of these material balances is shown in Table 11 and Figure 28. For time up to 2074 operating hours, the mass recovery was between 96% and 102%. Based on the accuracy of instrument readings of the demonstration plant, the mass recovery is believed to be $\pm 2\%$.

Power Plant Energy and Material Balances

The material balance envelope for the demonstration plant is shown in Figure 29. To have a true overall material balance, all streams external to plant boundary should be measured. Since this is not the case for the demonstration plant, some output stream flow rates had to be calculated from the measured flow rates of related streams inside the plant boundary. As a result, the accuracy of flow measurement for internal streams also affect the overall material balance. Therefore, the history of those measured flow rates that affect the overall material balance were reviewed and the following flow measurements were found to be questionable:

FI-128 (NG to startup burner F-300) : always at ~80% of capacity regardless of stack load

FI-201 (BFW): very erratic, even gives negative backcalculated blow down rates.

FI-222 (System Stack): never worked; always zero value.

To perform the overall plant material balance, a steady-state period for 1200 amps operation was selected. The result of this plant material balance calculation is shown in Table 12. For this period, the overall plant mass balance was 94%. Using the fuel cell mass balance data, the flow rate for cathode exhaust measured at FI-516 was also back-calculated. The measured FI-516 flow rate was found to be 13% less than the back-calculated value, indicating a fair agreement.

Table 9 Example of Calculating Reformer Product Flow Rate and Composition

Summary of Reformer Operatio										
					Oper. Dat	*				
Reformer Operating Conditions					Oper. Tim	*	P.S.T.			
Reformer Feed Temp.		698 F (TI-422A)		Approx. Reformer Pressure		43.60 psia				
Reformer Rxn. Temp		1390 F (TI-402)				28.90 psig (e.g., PI-303 or PI-603)				
Sample Line Shift Te		1390 F								
Temp/App=		0 F (Refm)		Note: T/A positive for endothermic reactio						
		-5 F (Water Gas Shift)								
Natural Gas Flow at Desulfurizer Outlet (FI-125)				79.35 lb/hr	4.6347 lbmole/hr	1759 SCFH				
Steam Flow to Reformer (FI-301)				360.50 lb/hr	20.0109 lbmole/hr	7594 SCFH				
Purge Gas (FI-821, nitrogen)				0 SCFM	0.0000 lbmole/hr	0 SCFH				
SV-823 Reading (either 0 or 1; 1=purge)				0 NG+Purge	4.6347 lbmole/hr	1759 SCFH				
Feed Composition (Natural Gas + Purge)					Reformer Feed Component Flow Rates (NG + Purge + Steam)					
	% Vol	SCFH	lbmole/hr	SCFH	lbmole/hr	lb/hr	% Vol		MW	
H2	0.00	0	0.00	0	0.00	0.00	0.00		2.0158	
CO	0.00	0	0.00	0	0.00	0.00	0.00		28.01055	
CO2	0.45	8	0.02	8	0.02	0.92	0.08		44.00955	
H2O	0.00	0	0.00	7594	20.01	360.50	81.19		18.0152	
CH4	93.02	1636	4.31	1636	4.31	69.16	17.49		16.04275	
C2H6	3.67	65	0.17	65	0.17	4.77	0.69		28.0539	
C3H8	0.64	11	0.03	11	0.03	1.31	0.12		44.09665	
N-C4H10	0.22	4	0.01	4	0.01	0.59	0.04		58.1236	
N2	2.00	35	0.09	35	0.09	2.60	0.38		28.0134	
O2	0.00	0	0.00	0	0.00	0.00	0.00		31.9988	
	100.00	1759	4.63	9353	24.65	439.85	100.00			
MW of Nat'l Gas		17.12095								
Calculated Reformer Input and Output										
					Calculated					
Reformer Flow			Reformer Flow		Rfrmr Prod. Comp.					
at Inlet			at Outlet		at Sample Point					
					(AP-518) (AP-518) (AP-518)					
					Measured Norm.wet Norm.,dr					
					GC, % GC, % GC, %					
H2	0.00	0.0000	16.0803	0.4730	32.41	47.30	76.66	47.56	47.56	77.07
CO	0.00	0.0000	2.3677	0.0696	66.32	6.96	11.29	6.49	6.49	10.52
CO2	0.02	0.0008	2.3301	0.0685	102.55	6.85	11.11	7.21	7.21	11.68
H2O	20.01	0.8119	13.0246	0.3831	234.64	38.31	0.00	38.29	38.29	0.00
CH4	4.31	0.1749	0.1041	0.0031	1.67	0.31	0.50	0.36	0.36	0.58
C2H6	0.17	0.0069	0.0000	0.0000	0.00	0.00	0.00	0.00	0.00	0.00
C3H8	0.03	0.0012	0.0000	0.0000	0.00	0.00	0.00	0.00	0.00	0.00
N-C4H10	0.01	0.0004	0.0000	0.0000	0.00	0.00	0.00	0.00	0.00	0.00
N2	0.09	0.0038	0.0927	0.0027	2.60	0.27	0.44	0.09	0.09	0.15
O2	0.00	0.0000	0.0000	0.0000	0.00	0.00	0.00	0.00	0.00	0.00
Total	24.65	1.0000	33.9996	1.0000	440.19	100.00	100.00	100.00	100.00	100.00
Predicted CH4 Conversion based on input steam and NG (Vol%)					97.58					
Actual CH4 conversion based on reformer product GC (Vol%)					97.44	(formula is based on no CO or CO2 or C2+ in feed gas)				
Net Heat Required for Reforming					592,503	Btu/hr				
Kp of Shift Rxn=(%H2*%CO2)/(%H2O*%CO)=					1.2150	(@immediate exit of reforming section)				

Table 10 Example of PDT-1 Stack Material Balance for NAS Miramar Operation

MCFC INPUT DATA									
Stack Volt. (VDC)	175.000	Volt, DC	(EI-539)						
Cell Volt. (VDC)	0.700	Volt, DC							
No. of Cells	250								
Active Area/Cell	11.39	ft ²							
Curr. Den., ASF	105.70676	ASF	1204	<-dc Amp	(IIC-550)				
Anode Fuel Util. =	0.673	mole frac.							
Stk pwr, kW (DC)	210.70	kW, DC							
Anode Feed Temp (TI-509), F	1225		1519.67		(TI-509)				
Anode Exhaust Temp (TI-501), F	1060				(TI-501)				
Temp/App for shift, F	-5	<-Negative=Exothermi							
Cathode Feed Temp (TI-502), F	1079				(TI-502)				
Cath. Exhaust Temp (TI-504), F	1279				(TI-504)				
Fuel Cell Vessel Outside Wall Temperature			140.00	F					
Fuel Cell Vessel Surface Area (excluding bottom)			1056.00	ft ²					
Average Ambient Temperature			60.00	F					
Heat Loss from Fuel Cell Vessel ***			(64,444)	Btu/hr					
Back-Calc. Heat Transfer Coefficient of free-convection ***			-0.76	Btu/ft ² -F-hr					
*** assuming fuel cell guard heaters stay inactive									
Cathode Inlet Flow			12482	lb/hr					
FCV N2 Purge (FI-506)			13.2	SCFM	(FI-506)				
Cathode Blower Outlet Flow (FI-524)			2844	ACFM	(FI-524)				
Cathode Inlet Flow (FI-506+FI-524)			452.72	lbmole/hr					
Cathode Inlet Pressure (PI-303)			28.9	PSIG	(PI-303)				
Cathode Inlet Pressure			43.6	psia					
Calculated Cathode Outlet Flow			11738.81	lb/hr	(no FI for cathode Outlet)				
Calculated Cathode Outlet Flow			434.15	lbmole/hr	(no FI for cathode Outlet)				
Anode Outlet Flow			1482.0917	lb/hr					
Anode Outlet Flow (FI-515)			319.8	ACFM	(FI-515)				
Anode Outlet Flow			51.31	lbmole/hr					
(% H2O in GC is the least accurate measuremen									
Cathode Inlet Composition (from AP-520) Cat. Out. Cmp. (AP-52) Anode Outlet Comp. (AP-519)									
	Raw %	Norm. %	Raw %	Norm. %	Raw %	Norm. %			
CO2	8.35	8.35	4.89	4.89	45.49	45.49			
H2O	19.76	19.76	20.11	20.11	41.24	41.24			
O2	4.94	4.94	4.22	4.22	0	0.00			
N2	66.94	66.94	70.77	70.77	1.19	1.19			
CH4	0	0.00	0	0.00	0.43	0.43			
CO	0.01	0.01	0.01	0.01	3.08	3.08			
H2	0	0.00	0	0.00	8.57	8.57			
Raw Total	100.00	100.00	100.00	100.00	100.00	100.00			
Avg. MW	27.57		26.95		28.89				
	(AP-520)		(AP-521)		(AP-519)				
MCFC STREAM SUMMARY (based on 100% sealing)									
	Anode Feed StreamS		Anode Exhaust Stream		Cathode Feed Stream		Cathode Exhaust Stream		
Temp, F	1225		1060		1079		1279		
Press, psia					43.6				
Component	lbmole/hr	m.f.	lbmole/hr	m.f.	lbmole/hr	m.f.	lbmole/hr	m.f.	Net
H2	16.1755	0.4756	4.9190	0.1060	0.0000	0.0000	0.0000	0.0000	
CO	2.2073	0.0649	1.0856	0.0234	0.0453	0.0001	0.0453	0.0001	
CO2	2.4522	0.0721	15.9520	0.3439	37.8020	0.0835	25.4239	0.0586	
H2O	13.0227	0.3829	24.2792	0.5234	89.4573	0.1976	89.4573	0.2061	
CH4	0.1224	0.0036	0.1224	0.0026	0.0000	0.0000	0.0000	0.0000	
C2H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	
C3H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	
C4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	
N2	0.0306	0.0009	0.0306	0.0007	303.0501	0.6694	303.0501	0.6980	
O2	0.0000	0.0000	0.0000	0.0000	22.3643	0.0494	16.1752	0.0373	
Total	34.0108	1.0000	46.3890	1.0000	452.7191	1.0000	434.1518	1.0000	-6.1891 lbmole/hr
lbs/hr	439.79		1182.59		12481.62		11738.81		-0.00 lb/hr
g-moles/s-cell	0.017141		0.023380		0.228170		0.218813		
Sens. Heat, Btu/hr	313864		422.364		3,552,249		4,096,794		653,045
Std HOF*, Btu/hr	(1,879,051)		(5,283,183)		(15,711,256)		(13,614,844)		(1,307,720)
* HOF= Heat of formation									
S No FI for anode inlet; input based on calculated reformer product flow									

Table 10 Example of PDT-1 Stack Material Balance for NAS Miramar Operation (Cont'd)

MCFC STREAM SUMMARY (based on 100% sealing)									
	Anode Feed Stream\$		Anode Exhaust Stream		Cathode Feed Stream		Cathode Exhaust Stream		
Temp, F	1225		1060		1079		1279		
Press, psia					43.6				
Component	lbmole/hr	m.f.	lbmole/hr	m.f.	lbmole/hr	m.f.	lbmole/hr	m.f.	Net
H2	16.1755	0.4756	4.9190	0.1060	0.0000	0.0000	0.0000	0.0000	
CO	2.2073	0.0649	1.0856	0.0234	0.0453	0.0001	0.0453	0.0001	
CO2	2.4522	0.0721	15.9520	0.3439	37.8020	0.0835	25.4239	0.0586	
H2O	13.0227	0.3829	24.2792	0.5234	89.4573	0.1976	89.4573	0.2061	
CH4	0.1224	0.0036	0.1224	0.0026	0.0000	0.0000	0.0000	0.0000	
C2H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	
C3H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	
C4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	
N2	0.0306	0.0009	0.0306	0.0007	303.0501	0.6694	303.0501	0.6980	
O2	0.0000	0.0000	0.0000	0.0000	22.3643	0.0494	16.1752	0.0373	
Total	34.0108	1.0000	46.3890	1.0000	452.7191	1.0000	434.1518	1.0000	-6.1891 lbmole/h
lbs/hr	439.79		1182.59		12481.62		11738.81		-0.00 lb/hr
g-moles/s-cell	0.017141		0.023380		0.228170		0.218813		
Sens. Heat, Btu/hr	313864		422,364		3,552,249		4,096,794		653,045
Std HOF*, Btu/hr	(1,879,051)		(5,283,183)		(15,711,256)		(13,614,844)		(1,307,720)
* HOF= Heat of formation									
\$ No FI for anode inlet; input based on calculated reformer product flow									
Anode Exhaust Stream									
Temp, F	1060								
Press, psia									
Component	Calculated Values		Based on AP-519						
	lbmole/hr	%mole	lbmole/hr	Norm. %			%Fuel Util		
H2	4.9190	10.60	4.3969	8.57	@100%wse	67.34			
CO	1.0856	2.34	1.5802	3.08	"Measured"	67.49			
CO2	15.9520	34.39	23.3390	45.49					
H2O	24.2792	52.34	21.1585	41.24					
CH4	0.1224	0.26	0.2206	0.43					
C2H6	0.0000	0.00	0	0.00					
C3H8	0.0000	0.00	0	0.00					
C4H10	0.0000	0.00	0	0.00					
N2	0.0306	0.07	0.6105	1.19					
O2	0.0000	0.00	0.0000	0.00					
Total	46.3890	100.00	51.3059	100.00					
lbs/hr	1182.5883		1482.0917						
Cathode Exhaust Stream									
Temp, F	1279								
Press, psia									
Component	Calculated Values		Based on AP-521						
	lbmole/hr	%mole	lbmole/hr	Norm. %					
H2	0.0000	0.00	0.0000	0.00					
CO	0.0453	0.01	0.0428	0.01					
CO2	25.4239	5.86	20.9399	4.89					
H2O	89.4573	20.61	86.1147	20.11	@100%wse	% CO2 Util	% O2 Util.		
CH4	0.0000	0.00	0.0000	0.00	"Measured"	33	28		
C2H6	0.0000	0.00	0.0000	0.00					
C3H8	0.0000	0.00	0.0000	0.00					
C4H10	0.0000	0.00	0.0000	0.00					
N2	303.0501	69.80	303.0501	70.77					
O2	16.1752	3.73	18.0708	4.22					
Total	434.1518	100.00	428.2184	100.00					
lbs/hr	11738.81		11541.84						

Table 10 Example of PDT-1 Stack Material Balance for NAS Miramar Operation (Cont'd)

Mass Balance Around the Fuel Cell			
	Mass Flow Rate (lbs/hr)		
Anode Feed	439.79	(based on calculated reformer product stream)	
Cathode Feed	12,481.62	(based on FI-524, AP-520, PI-303, and TI-502)	
Total Mass In	12,921.40		
Anode Exhaust	1,482.09	(based on FI-515, AP-519, PI-303, and TI-501)	
Cathode Exhaust	11,541.84	(based 100% N2 balance over cathode and AP-521)	
	13,023.93		
% of unaccounted mass (100*(mass out-mass in)/mass in)			
	0.79	%	
FUEL CELL PERFORMANCE			
Stack Volt. (VDC)	175.000	(EI-539)	
Cell Volt. (VDC)	0.700		
No. of Cells	250		
Active Area/Cell	11.39	ft ²	
Curr. Den., ASF	105.70676	ASF	
Anode Fuel Util =	0.673	mole frac.	
Stk pwr, kW (DC)	210.70	kW, DC	
(H2+CO) Cons.	12.378	lbmol/hr	
(H2+CO) Fed	18.383	lbmol/hr	
Fuel Cell HEAT BALANCE			
Anode in	313,864	Btu/hr	Btu/h
Anode Out	422,364	Btu/hr	
Cathode In	3,552,249	Btu/hr	
Cathode Out	4,096,794	Btu/hr	
Std Ht-of-Rx	(1,307,720)	Btu/hr	
Heat Loss	(64,444)	Btu/hr	(positive means heat loss)
Power	719,119	Btu/hr	
% Fuel Cell Heat Loss (relative to power output)			
		-8.96	%
Fuel Cell Vessel Outside Wall Temperature			
Fuel Cell Vessel Surface Area (excluding bottom)		140.00	F
Average Ambient Temperature		1056.00	ft ²
Heat Loss from Fuel Cell Vessel ***		60.00	F
Back-Calc. Heat Transfer Coefficient of free-convection ***		(64,444)	Btu/Hr
*** assuming fuel cell guard heaters stay inactive		-0.76	Btu/ft ² -F-hr

Table 11 Summary of Material Balance for PDT-1 Stack at NAS Miramar

			Mass	Stack	Stack	System	Anode**	Oxygen**
Operating			Recovery	DC Load	DC Volt	Pressure	Fuel Util.	Utilization
Hours	Date	Time	%w	Amps	Volts	Psig	%	%
339.5	10/24/97	15:30	97.3	333	208	24.5	20	5
459	01/29/97	15:30	96.7	515	199.9	24.5	55	12
748	02/10/97	16:00	96.8	0	241.81	27.7	0	0
822	02/13/97	18:00	97.4	805	192.54	25.4	78	17
931.75	02/18/75	07:45	100.4	1211	180.15	27.9	75	26
940	02/18/75	16:00	98.9	1125	178.43	28.3	71	19
1070	02/24/97	02:00	99.3	709	193.32	25.7	67	18
1106	02/25/97	14:00	99.2	1095	178.65	28.1	57	33
1168	*	*	99.2	1204	175	28.9	67	28
1291	03/05/97	07:00	102	467	200.34	23.8	64	15
1402	03/09/97	22:00	101.3	533	201.5	24.6	54	17
1945	04/01/97	13:00	101.3	485	193.21	22.8	56	14
2074	04/06/97	23:00	99.4	697	179.68	24.3	63	17
* Average for period from 1132 to 1204 operating hours.								
** Based on 100% wet seal efficiency								

Figure 28 History of Material Balances for PDT-1 Stack at NAS Miramar

Figure 29 NAS Miramar Plant Material Balance (All numbers are in lb/hr)

Table 12 Overall Mass Balance for NAS Miramar Plant at 1200 Amps Operation

Overall Power Plant Material Balance between **1132** and **1204** op. hrs.

Input Streams at **1204** Amps (avg.)

NG (FI-125 + FI-128 + FI-821)		83	lb/hr
NG Flow at Desulfurizer Outlet (FI-125)		79	lb/hr
NG to S/U Burner (FI-128)	1.75 scfm	4	lb/hr
N2 (FI-506)	13.19 scfm	58	lb/hr
N2 (FI-821, purge to NG feed, NNF)	0.0 scfm	0	lb/hr
N2 (FI-822, purge to IHI Comb., NNF)	0.0 scfm	0	lb/hr
BFW (FI-201)		440	lb/hr
Air (FI-604 + FI-608 + Air to S/U Burner)	833.3 scfm	3801	lb/hr
Air to S/U Burner (normally no flow — N/A in database)	0.0 scfm	0	lb/hr
Air to Cathode and Combustor (FI-608)	559.0 scfm	2550	lb/hr
Air to Expander (FI-604)	274.3 scfm	1251	lb/hr
Total Input		4382	lb/hr

Molecular Weight of Air = 28.8503 lb/lbmole

Output Streams

Flue Gas (FI-222 = FI-604 + FI-516 + FI-128 + Air to S/U Burner)				4011	lb/hr
Air to S/U Burner (normally no flow — N/A in database)				0	lb/hr
NG to S/U Burner (FI-128)				4	lb/hr
Air to Expander (FI-604)		274.3	scfm	1251	lb/hr
Cathode Exhaust to Expander (FI-516)		646.5	scfm	2756	lb/hr
Turbocharger Inlet Temperature (TI-605)	735.1		acfm		
Turbocharger Inlet Pressure (PI-303-PDI-303)	1220.5		deg F		
Recycle Blower Outlet Pressure (PI-303)	41.80		psia		
Recycle Blower DP (PDI-303)	28.91		psig		
Cathode Outlet Gas Composition	50.23		inwc		
CO2	4.89		%	Check for 100%	
H2O	20.08		%	99.99	
O2	4.29		%		
N2	70.72		%	Avg MW(lb/lbmole)	
CO	0.01		%	26.96	
Export Steam (FI-211)				90	lb/hr
BFW Blowdown (assumed to be 0.5% of BFW)				2	lb/hr
Anode Outlet to Sealpot (FI-503)				0	lb/hr
Total Output				4103	lb/hr
% Weight Recovery	93.63	%	Total Input - Total Output =	279	lb/hr

Back Calculated FI-516

shaded areas above must be filled for calculated values in this section to be correct

Total Cathode Exhaust (from Fuel Cell material balance)			11489	lbs/hr	A		
Cathode Exhaust going to IHI Combustor (C - D)		329.4	scfm	1404	lbs/hr	B	
Average Molecular Wt of Cathode Exhaust	27.0	lb/lbmole					
Air/Cathode Exh. Stream to IHI Comb. (FI-431)	473.0	acfm	504.4	scfm	2202	lbs/hr	C
Air/Cathode Exh. Temp. to IHI Comb. (TI-430A)	986.1	deg F					
Air/Cathode Exh. Pressure to IHI Comb. (PI-303)	28.9	psig					
Air to IHI Combustor (FI-430)		174.9	scfm		798	lbs/hr	D
Sum of Cathode Exhaust to Cathode Inlet & Turbocharger (A - B)					10085	lbs/hr	E
Cathode Inlet (Blower Inlet + FI-506)					12480	lbs/hr	F
Cathode Hot Blower Outlet Flow (FI-524)	2844.0	acfm	2849.8	scfm	12422	lb/hr	F1
Cathode Inlet Temp. (T-502)	1079.0	deg F					
Cathode Inlet Pressure (PI-303)	28.9	psig					
Cathode Inlet Gas Composition							
CO2	8.35	%	Check for 100%				
H2O	19.76	%	100.00				
O2	4.94	%					
N2	66.94	%	Avg MW(lb/lbmole)				
CO	0.01	%	27.57				
Air to Cathode Recycle Line (FI-608 - FI-430)		384.1	scfm		1752	lbs/hr	G
IHI Combustor Outlet Stream to Cathode Recycle Line (B + D + I + J)					3685	lbs/hr	H
PAC Steam (FI-307)					98	lbs/hr	I
Cathode Exhaust going to IHI Combustor					1404	lbs/hr	B
Air to IHI Combustor (FI-430)		174.9	scfm		798	lbs/hr	D
Anode Outlet Stream to IHI Combustor (FI-515)	319.8	acfm	324.3	scfm	1385	lbs/hr	J
Anode Outlet Temperature (TI-501)	1060.4	deg F					
Anode Outlet Pressure (~PI-303)	28.9	psig					
Anode Outlet Gas Composition (from Fuel Cell material balance)							
CO2	36.83	%	Check for 100%				
H2O	53.17	%	100.00				
N2	0.95	%					
CH4	0.16	%	Avg MW(lb/lbmole)				
CO	2.86	%	27.00				
H2	6.03	%					
Cathode Recycle to E-200 Inlet (F1 - G - H)					6985	lbs/hr	K
Back calculated FI-516 (E - K)					3100	lbs/hr	
Measured FI-516					2756	lbs/hr	
% Difference from Measured Value					12.48	%	
MISC. Stream Flow							
Steam to Reformer NG Feed (FI-301)					361	lb/hr	

* Measured FI-516 is assumed to be correct for overall plant material balance. Its accuracy will affect this calculation.

ovt1_mb1.vb3

Using the material balance data and actual conditions of streams and operating units, energy balance for the overall power plant was also calculated. The energy balance estimates are less accurate than the material balance estimates because many of the required energy components, such as heat loss through the pressure vessels and pipe insulation, heater duties, etc., had to be estimated. The stream enthalpies as calculated by a commercial process simulator HYSIM are based on ideal gas behavior with zero enthalpy at absolute zero temperature. The result of the energy balance calculations, related mass flow rates, and various efficiencies are shown in Table 13.

According to the demonstration plant design for the nameplate case, where the stack is to operate at 1822 amps with fuel utilization of 70%, the power plant electrical efficiency is 34.2% HHV. The actual operation, as shown in Table 13, shows an electrical efficiency of 33.0% HHV that corresponds to 1200 amps operation with anode fuel utilization of 75%. It appears that if it were not for the high cathode pressure drop problem and over-estimated stack performance by the PSI model, the plant would have achieved the design efficiency.

The power plant efficiency as defined by the Federal Energy Regulatory Efficiency Commission is as follows:

$$FERCEff = \frac{NetElectricity + 50\% * (ExportedSteamNetEnthalpy)}{FuelLHV}$$

For the design case of full load and maximum efficiency (1822 amps with anode utilization at 80%), the FERC efficiency is estimated to be 54.7%. For the Miramar demonstration plant operation, it did not achieve the full load. Using the actual operating conditions at 1200 amps and 75% fuel utilization, the corresponding FERC efficiency is estimated to be 43.1%, which slightly exceeds the 42.5% minimum FERC efficiency required for connection of a cogeneration facility of this type to a public utility grid.

LESSONS LEARNED

Based on the operating experience gained during the Process and Control (PAC) test and demonstration test, problems, solutions implemented, or solutions identified for future resolution were documented by Bechtel. The important lessons gained from operating this demonstration plant are summarized below.

BOP Equipment

Recycle Hot Blower

The major problems encountered with the Hot Cathode Recycle Gas Blower were vibration at a lateral critical speed within the operational speed range and leakage through the shaft packing gland seal. The critical speed always occurred between 2300 and 2800 rpm and was avoided during

Table 13 Summary of Overall NAS Miramar Plant Energy Balance during 1200 Amps Operation ****

	Flow lb/hr	Latent Heat MM Btu/Hr	Sensible Heat MM Btu/Hr	HHV MM Btu/Hr	Electrical Energy MM Btu/Hr	Total Energy MM Btu/Hr	Fuel Cell No. of Cells	250
INPUT STREAMS							Active Area/Cell, ft2	11.39
Natural Gas	67.74		0.018387	1.61634		1.63473	DC Voltage/Cell, V	0.7206
Air	3801		0.474851			0.47485	DC Stack Volt., V	180.15
BFW	440		-0.37155			-0.37155	DC Load, Amps	1211
Inert Gas Purge	58		0.00744			0.00744	DC Load, mA/ft2	106.32
Cathode Recycle Blower					0.0798	0.07983	Stack DC Power, kW	218.16
Anode Preheater					0.0108	0.01075	Anode Fuel Util., %	74.8
Cathode Preheater					0.0798	0.07983	% of Ht of Rxn as Electricity	56.45
BFW Pump					0.0000	0.00001	Fuel Cell Mat'l Bal., %w	99.6
Turbo-compressor					0.2383	0.23830		
TOTAL	4367					2.15419	Fuel to Plant	
							Flow, lb/hr	67.74
							HHV, MM Btu/Hr	1.616344
OUTPUT STREAMS							Design Efficiency	
Stack Gas*	4276.74		0.450763			0.45076	Cathode Rcy. Blower,%	65
Export Steam	90		0.05107			0.05107	Turbo-expander, %	75
Blowdown	0		0			0.00000	Turbo-compressor,%	72
Reformer Heat Loss**					0.0400	0.04000	Inverter, %	95
Fuel Cell Heat Loss**					0.1000	0.10000	BFW Pump,%	60
Inverter Loss					0.0741	0.07410		
Gross Power, a.c.						0.67046	Electricity	
Turbo-expander					0.3512	0.35120	Stack a.c. Power, kW	196.45
TOTAL	4367					1.73759	Cathode Rcy. Blower, kW	23.39
Overall Plant Heat Bal, %						80.66	Anode Preheater, kW***	3.15
							Cathode Preheater, kW***	13.5
							BFW Pump, kW	0.004
							Net Turbo Output, kW	33.08
							Net Electricity, kW	156.406
							Plant Elect. Efficiency, %	33.03
							Plant Heat Rate, HHV, Btu/kwh	10332.73
							Actual Efficiency	
							Cathode Rcy. Blower,%	-
							Turbo-expander, %	-
							Turbo-compressor,%	69.15
							Inverter, %	90.05
							BFW Pump, %	-

* Backcalculated assuming 100% overall material balance. Measured value (FI-516) is 82 % of the backcalculated value. FI-608 Air Flow might also be reading too high, since cathode inlet GC data shows much lower %O2 than design values. Overall plant material Bal. was about 94%.

** Estimated value

*** Estimated from rated duty at 15% output

**** Fuel Cell Balance data @ 931.75 operating hours (2/18/97 7:45 a.m.) were used for turbocharger and cathode recycle blower calculation.

operation by setting the blower speed above or below the critical speed range. The blower was designed to operate at 3500 rpm continuously. The sporadically leaking packing gland required placing the plant in warm hold for at least one calendar day to replace the packing.

It is known that sustained excessive vibration will also cause the packing to deteriorate, but this was not how the blower was operated at Miramar. The blower's vibration signature and operating experience showed that this blower ran smoothly. Since the glands wore out at 3500 rpm blower speed (well outside of the critical speed area), and the glands wore down slowly, it is most likely that the gland failure was caused by the frictions between the gland and the shaft. The remedy to these vibrations is to replace the blower shaft with one which does not cause the critical speed. The remedy to packing gland leakage is to replace the packing gland seal with a labyrinth seal.

Inverter

The major problems encountered with the inverter were sporadic shutdowns due to susceptibility of the logic board to electromotive interference (EMI) and incorrect inverter action when the fuel cell voltage dropped due to the controller logic based on constant power instead of constant current. A loose set of a.c. power connections within the inverter cabinet, occluded from view, caused random over-voltage transients (power spikes) which emitted electric fields that interfered with the control logic. When adequate noise suppression devices were installed to protect the logic board, it was possible to pin-point the problem immediately. The connections were tightened down and the problem was solved. It was also found that the controller action was specified incorrectly and was revised to respond properly.

Turbocharger

The major problems encountered with the turbocharger were poor reliability and low efficiency. The nine units installed at the NAS Miramar demonstration plant were not reliable for reasons ranging from thrust bearing failures, lube oil contamination with process gases, impingement of the speed pick-up's fiber optic lens on the compressor blade when the lens dislodged from its holder, and lubricating oil feed pump foul ups. These problems were resolved when the fifth turbocharger was operating. The remaining turbochargers displayed about the same behavior. Within 10-days of service they began to flow less air at a given system pressure which caused the turbocharger to surge. Investigation of the expander blades showed that a rough hygroscopic coating had developed which could contribute to resistance to flow through the expander.

The turborcharger efficiency was also low and resulted in lower system pressure than desired. The problem was exacerbated at temperatures higher than the iso-point (60° F) as the compressor and expander horsepowers balance. The problem stems from using an off-the-shelf turbocharger not specifically engineered for this application. The remedy for both of these problems is either procure a turbocharger that is specifically designed and manufactured for this power plant application or to try a different type of unit, such as a microturbine. A microturbine appears to have proper operating

characteristics and also has the advantage of converting the excess mechanical power from the turbine into usable electricity.

IHI Reformer

The problem encountered with the IHI Reformer was hot/cold spots in the combustor. Whenever the combustion-to-reforming heat balance was not near the IHI's design point associated with an apparent anode fuel utilization of 65%, a large temperature differential would develop in the combustor. Due to the tight process integration and the control logic used, it was difficult to keep the combustor running within the desired temperature differential.

Typically, a reduction in the anode fuel utilization would cause an immediate rise in the average combustor temperature due to an increase in the heating value in the anode outlet stream. The control logic would then adjust the compressed air flow to the combustor to maintain a proper ratio of mixed oxidant to the fuel at the combustor. The combustor average temperature used to control the mixed oxidant/fuel ratio was the average temperature of all forty plates in the combustor, but this average temperature is not adequate to reflect the unacceptable temperature differential in the combustor. Such a temperature differential would cause pre-trip alarms (high-high combustor plate temperature). This problem was especially prevalent at combustor temperatures averaging less than 1410 °F (design was 1460 °F), and was solved by operating the fuel cell near 65% fuel utilization, and the combustor at about 3:1 mixed oxidant/fuel ratio, while maintaining the average combustor temperature above 1410 °F.

Fuel Cell Stack

Scale-up Effect on Stack Voltage

There are inherent scale-up effects for scaling up a singular isothermal 100 cm² bench-scale cell to a non-isothermal tall stack of 250 cells with 10,500 cm² area per cell. Based on the PSI model predictions which have been verified by bench-scale data, the PDT-1 Stack operations at the ATF and at the Miramar demonstration have shown that the actual average stack voltage performance was always lower than those predicted. At the ATF, the performance was lower by 60 to 100 mV and at Miramar it was lower by 30 to 90 mV. Considering the uncertainties of flow measurements associated with the Miramar demonstration operation, both ATF and Miramar operations show equal magnitude of scale-up effects. Therefore, scale-up effects should be considered in the future plant designs.

Cathode Pressure Drop

The cathode pressure drop at the beginning of the Miramar demonstration operation was already about twice the pressure drop of the ATF operation, and it deteriorated with time. For operation at 700 amps, the cathode pressure drop increased by a factor of five over about 1500 hours

of operation (Figure 23). It is expected to be even worse for operations at higher currents. This problem and scale-up effects are believed to be the primary reasons for preventing the power plant reaching the design power output.

Task 7 ARPA Enhancements

Objective: Provide support to the PDT demonstration work related to the change from the original demonstration site at the Kaiser Permanente Medical Center to NAS Miramar and in areas which will improve chances of project success.

Discussion: All the work on the following subtasks was coordinated with and complementary to the cognizant tasks and subtasks discussed earlier.

7.1 Site Design, Permits and Manuals

7.2 Site Preparation and Installation

7.3 Stack Conditioning and Shipping

7.4 NEPA Documentation, Update and Approvals

7.5 Full-Area, 20 kW Stack Verification

7.6 Advance Separator Plate Qualifications

7.7 Pressure Vessel Design Improvement

The pressure vessel for the stack was reduced in weight by 30% and the diameter was reduced by 15%. This was accomplished by employing a reusable cage for stack lifting. A change in the specification of the inverter eliminated the need to maintain electrical isolation between the plenum and pressure vessel. Consequently, ceramic flanges were eliminated, simplifying the internal piping. Additionally, the fuel cell stack instrumentation was significantly reduced. An on-line gas chromatograph was purchased for better control of start-up and material balances.

The pressure vessel was successfully hydro tested after modifications to accommodate a camera and light source were completed. The camera system provides a back-up measurement system for the stack to observe and verify digital readouts of stack shrinkage during conditioning.

4.0 CONCLUSIONS AND RECOMMENDATIONS

- ▶ The use of 10 intermediate plates in the PDT-1 250 kW stack at NAS Miramar proved to be successful in reducing the accumulative tolerances previously seen in the Unocal

250 kW stack and resulted in a very flat stack. For the PDT-1 stack, the difference between the average anode-side stack heights and the average cathode-side stack heights was only 0.3 mm, compared with 13.3 mm for the Unocal stack.

- ▶ The PDT-1 stack met the M-C Power acceptance criteria for leakage: <0.4 slpm/cell at ambient temperature, 40 psi top clamping pressure and 10 iwc leakage pressure. The ambient leakage tests were conducted after the stack assembly, after the ATF conditioning and acceptance testing, and after the stack was shipped to NAS Miramar. All leakage test results showed acceptable leakage rates. The leakage rates to the atmosphere were also reduced significantly after the conditioning step, probably due to better sealing resulting from carbonate impregnation into the matrix.
- ▶ The changes in instrumentation fittings and thermocouple assembly specifications based on the Unocal 250 kW stack operating experience has resulted in no instrumentation failures in the ATF test and minor instrumentation problems at the NAS Miramar demonstration.
- ▶ The PDT-1 stack tested in ATF showed no sign of leakage (i.e., no loss of any manifold pressures) during the electrolyte melting step, indicating successful implementation of conditioning strategy. The total stack shrinkage for the electrolyte melting step was 294 mm, or 1.18 mm/cell.
- ▶ The OCV data from the ATF test also imply that the PDT-1 stack had good sealing both before and after the acceptance test. The predicted OCV based on actual gas compositions was 268 volts for the stack, and the measured OCVs before and after were 268 and 267 volts, respectively.
- ▶ The PDT-1 stack tested in ATF met the acceptance criteria: > 207.5 volts at 40/cm². The stack voltage was 210 volts at 40/cm² (423 amps).
- ▶ The ATF test data for PDT-1 stack indicates an estimated average anode wet seal efficiency of 86% and estimated average cathode wet seal efficiency of 98%. The relation between wet seal efficiencies and the pressure differential between anode and cathode also indicate possible development of cracks between anode and cathode during the ATF test.
- ▶ The NAS Miramar demonstration plant was successfully started up and reached a maximum power production of 210 kW d.c. Power. The power plant produced a total of 160 MWH d.c. electricity and 346,000 lbs of 100 psig steam for export.
- ▶ High cathode pressure drop and scale-up effects caused the demonstration plant to reach only 63% of the design power output: 330 kW d.c.

- ▶ Three pieces of BOP equipment - cathode recycle blower, inverter and turbocharger, did not have reliable performance, especially the turbocharger which caused numerous plant shutdowns for repair or replacement.
- ▶ The tight operating constraints required by the IHI plate-type-reformer, also the first prototype tested for commercial operation, restricted the plant operations to a very small operating window.
- ▶ The PDT-1 stack material balances at NAS Miramar varied between 97 to 102% for the first 2100 operating hours. The overall plant material balance at 210 kW operation is estimated to be 94%. For this period, the net a.c. electrical efficiency for the demonstration plant was 33% and the corresponding FERC efficiency was 43.1 which exceeded the 42.5% minimum FERC efficiency for connecting a cogeneration facility of this type of power plant to a public utility grid.