Binary Module Test

AP-1645
Research Project 1094

Final Report, December 1980

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ABSTRACT

The objective of this project was to design and test a binary loop module representative of and scaleable to commercial size units. The design was based on state-of-the-art heat exchanger technology, and the purpose of the tests was to confirm performance of a supercritical boiling cycle using isobutane and a mixture of isobutane and isopentane as the secondary working fluid. The module was designed as one percent of a 50 MW unit. It was installed at Magma Power's East Mesa geothermal field and tested over a period of approximately 4 months. Most of the test runs were with isobutane but some data was collected for hydrocarbon mixtures. The results of the field tests are reported herein. In general these results indicate reasonably good heat balances and agreement with overall heat transfer coefficients calculated by current stream analysis methods and available fluid property data; however, measured pressure drops across the heat exchangers were 20 percent higher than estimated. System operation was stable under all conditions tested.

The project was cofunded by the Electric Power Research Institute and the Department of Energy/Lawrence Berkeley Laboratory.

EPRI PERSPECTIVE

PROJECT DESCRIPTION

This report for RP1094 contains the results of an effort undertaken to determine the overall performance characteristics of a binary module in geothermal service. The module was configured to be similar to that expected for full-scale commercial use and to be approximately one percent of the size required for a 50-MW(e) binary cycle geothermal power plant. The module was designed for a resource temperature of 360°F and to operate at supercritical boiling conditions using a light hydrocarbon as the secondary working fluid. The module was designed with state-of-the-art techniques, fabricated, installed at the East Mesa geothermal field in southern California, and tested with geothermal fluids from that field. The geothermal brine rate through the brine-to-hydrocarbon heat exchanger was about 200 gpm with a corresponding heat flow of about 16 million Btu/h.

The size of the unit was selected to be compatible with the view held by a number of heat exchanger manufacturers--that data from a one percent module would enable them to design a commercial-size unit [50 MW(e)] with a high level of confidence that it would perform as expected.

PROJECT OBJECTIVE

The objective of the project was to determine whether significant differences might be expected between actual performance of binary system heat exchangers and predicted performance, using state-of-the-art design methods. Areas of interest included heat exchangers, condensers, operating characteristics, and the effect of repetitive chemical cleaning on heat exchanger performance. It was also felt that these tests would help identify potential problem areas in design, control, and reliability.

PROJECT RESULTS

Most of the data collected were based on pure isobutane as the secondary working fluid, but some test runs were made with mixtures of predominantly isobutane with propane and isopentane. Several factors combined to limit the quantity and

quality of the data; however, a number of results are valid and should be made available to binary cycle power plant designers and operators. The factors that affected the data were (1) low brine temperature, (2) oversized condensing train, (3) leakage in one of the heat exchangers, (4) instrumentation problems, and (5) inadvertent contamination of the isobutane with propane preceding the hydrocarbon mixture runs.

Almost 300 data scans were completed at different operating conditions with the pure isobutane. Over the heating portion of the cycle there was relatively close agreement between the measured and predicted heat transfer coefficients. In the supercritical region, the measured values ranged about 10 percent higher than predicted; those measured in the subcritical region ranged about 10 percent lower. Although this suggests room for improvement in prediction techniques, and such improvement should be pursued, it also indicates that existing state-of-the-art methods and data bases can be used for engineering purposes with an accuracy of about 15 percent. Overall heat balances for the heating portion of the isobutane Rankine cycle agreed within about 7 percent. One unexpected finding was that the pressure drop across the heat exchangers in the isobutane Rankine cycle agreed within about 7 percent; another unexpected finding was that the pressure drop across the heat exchangers in the isobutane loop appeared to be somewhat greater than predicted.

The original intent was also to measure hydrocarbon condensing coefficients, but this data was not reliable. It was also the intent to develop data for a mixture of 90 percent isobutane and 10 percent isopentane and for a mixture of 80 percent isobutane and 20 percent isopentane. The mixture was inadvertently contaminated with propane; therefore, these results were interesting but did not establish the desired data points. The present plan is to relocate the binary module to the East Mesa Federal Test Facility and repeat these tests, the results of which will be published separately.

Some of the equipment and operational problems encountered during these tests may be indicative of those likely to be encountered with commercial systems and are worth reviewing. Some of the more significant ones were (1) residual particulate contamination in the hydrocarbon loop, (2) residual cleaning-liquid contamination in the hydrocarbon loop, (3) heat exchanger leakage, (4) system control, and (5) cooling-water impurities.

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Section 1

SUMMARY

INTRODUCTION

In recognition of the potential role of binary loop power cycles in the development of moderate temperature hydrothermal-geothermal resources and the fact that the process had not been demonstrated on a commercial scale, the Electric Power Research Institute and the Department of Energy acting through the Lawrence Berkeley Laboratory agreed in November 1977 to cofund this project. The concept was to design a binary loop module that was representative of commercial size units in terms of system configuration and performance and to test the system under representative field conditions. The system was designed for a resource temperature of 360°F and to operate as a supercritical boiling cycle using a hydrocarbon as the secondary working fluid. The intent was to run the tests with isobutane and isopentane. The system was sized as one percent of a 50 MWe system. This translated to a geothermal fluid flow of approximately 200 gpm from the Magma Power Company's East Mesa Geothermal Field.

Approximately five thermal megawatts were transferred from the geothermal brine to the hydrocarbon working fluids in a closed system. The turbine function was simulated with a throttle valve and the heat was removed from the working fluids in a condenser and subsequently rejected to the environment by a wet cooling tower. The thermodynamic performance of both the working fluids and the system components was measured during the test program.

The objectives for this field test program were as follows:

Broad Objectives

1. Determine the overall (or average) heat transfer parameters for one-pass, countercurrent flow brine secondary-working-fluid heat exchangers typical of those expected for a commercial-size power plant.
2. Observe the behavior of secondary fluids in processes similar to those planned for a commercial-size power plant.

3. Confirm fouling factors for the design of brine/secondary fluid heat exchangers appropriate to a commercial-size plant.
4. Establish heat exchanger tube-side cleaning procedures and the effects of procedures upon exchanger performance.

Specific Objectives

1. Measure temperatures and pressures of the brine and the secondary fluids along their flow paths.
2. Perform heat and mass balances on the heat exchanger and condenser.
3. Determine overall heat transfer coefficients for the following conditions:
 - Brine on the tube-side and water on the shell-side.
 - Hydrocarbon on the shell-side and brine on the tube-side.
 - Increasing fouling factors (tube-side and shell-side).
4. Analyze pressure drops to determine shell-side mean friction factors.
5. Compare experimental results with calculated results of heat transfer, pressure drop, and shell-side fluid properties (for each run).
6. Determine brine side scaling build-up, scale compositions, corrosion rates, and corrosion products and compare with those obtained in prior tube and annulus experiments.
7. Compare experimental heat transfer data with results predicted using well-known heat transfer correlations and examine their validity (at the local Reynolds number, Prandtl number, Mean Nusselt number data points corresponding to the heat exchanger sections used).
8. Study scale removal procedures and investigate possible sensitization of brine side (tube side) surfaces to repeated scale removal operations by measuring scaling rates, corrosion rates, scale composition, and corrosion products.
9. Confirm test loop design parameters and thus provide a benchmark for comparison with existing design methods.

OPERATIONS AND DATA ANALYSIS

Test operations and the analysis of the measured data began on October 31, 1979. All test operations were terminated January 16, 1980. Testing in accordance with

the Test Plan was prematurely shortened on December 5, 1979 due to loss of hydrocarbon fluid through three pinhole leaks in a heat exchanger tube to tube sheet weld. Subsequent test operations occurring in January, 1980, were based on knowledge gained during the prior test work and recognized that several of the goals and specific project objectives could not be met because the geothermal brine temperature was lower than expected. The reason for the lower temperature is attributed to the fact that most of the geothermal fluid came from the low temperature well in the field. It should be noted that the well was pumped throughout the test.

- The geothermal brine was found to be generally corrosive. No brine side fouling occurred. Hence the ability to monitor this fouling as a function of time and to investigate the effects of chemical cleaning of brine exposed surfaces was not practical.
- All test operations, both with subcritical and supercritical heat exchanger hydrocarbon fluid conditions, were found to be extremely stable and repeatable. Problems with pressure measurements from exchanger to exchanger curtailed the ability to further investigate the behavior of the hydrocarbon working fluid near its critical point. The stable operation, coupled by this lack of instrumentation precluded the need to vary heat exchanger test conditions over small increments.
- The extreme fouling apparent in the condenser, coupled with the approximately four times excess area in the condenser made it difficult to obtain detailed condenser performance data.

Accordingly, the test operations during January, 1980, were redirected to obtaining operating data with hydrocarbon fluid mixtures. Even though the isobutane obtained from a commercial supplier in December, 1979 to recharge the system after the heat exchanger repair had been contaminated with propane, two series of tests were run. The first series of hydrocarbon mixture tests included approximately 10 mole percent isopentane, the second approximately 20 mole percent isopentane.

All test operations were stable and easily controllable. Once the system was adequately charged with hydrocarbon fluid, predetermined process conditions could be set manually and repeatably. Little need existed to rely on the automatic control capability designed into the test systems.

Excessive manufacturing slag and machined chips were found in the heat exchangers. This caused uneven operation of the hydrocarbon throttle valve.

Also, excessive temperatures in the cooling water coupled with cooling water contamination caused a rapid buildup of scale on the cooling surfaces and on the cooling water control valve trim. These latter two problems were solved in the field by primary reliance on manual operation of the controllers. This manual operation enhanced the operational stability of the system.

In all, 281 data scans were completed before loss of hydrocarbon due to leakage caused premature termination of the hydrocarbon heat exchanger isobutane performance tests. An additional 44 data scans were completed in January, 1980 with various fluid mixtures. These included nineteen with approximately 5 mole percent propane in isobutane, twenty with about 10 mole percent isopentane in the prior fluid, and five with about 20 mole percent isopentane. Not all data scans were of sufficient quality to be used as input to the Data Processing System. A set of 119 data scans constitute the set of coherent data available from the runs. Table 1-1 summarizes the test runs.

Table 1-1
Operations and Data Analysis Summary

<u>Working Fluid</u>	<u>Test time (hrs)</u>	<u>Number of data scans</u>	<u>Number of qualified data scans</u>
>98% isobutane	582	282	119
4.8% propane 95.2% isobutane	24	19	note 1
4.3% propane 85.8% isobutane 9.9% isopentane	25	20	note 1
3.8% propane 76.5% isobutane 19.7% isopentane	5	5	3

Note 1: Physical properties not available for performance analysis, see Section 4

SIGNIFICANT RESULTS

The operational and analytic results from these tests can be summarized as follows:

- Heat exchange to hydrocarbon working fluids and mixtures of hydrocarbon working fluids has been shown to be very stable at both subcritical and supercritical boiling conditions. No

thermal or hydraulic instability was observed at the various fluid states in the Rankine cycle.

- Overall heat balance calculations done over the heating portion of a working fluid Rankine cycle showed agreement with calculated values to within $\pm 7\%$. This indicates:

--enthalpy property data of the hydrocarbon fluids both above and below their critical state are valid for heat exchanger design purposes, and

--these mass flow rate ratios, geothermal brine to hydrocarbon working fluids, can be used and relied upon in optimization studies for commercial plant design.

- Overall heat transfer coefficients deviate little from those predicted by current stream analysis methods. The heat transfer coefficients in the supercritical region are about 10% higher than predicted, those in the subcritical region about 10% lower. This is reasonably close agreement. The information developed by these tests, suggests that heater and condenser design to within 15% for applications with similar, nonfouling, geothermal brine characteristics is possible.
- Pressure drop data indicates that the pressure drops experienced by the hydrocarbon fluids across the heat exchangers are about 20% greater than calculated. Future heater and condenser design and overall system design should include this consideration.

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Section 2

CONCLUSIONS AND RECOMMENDATIONS

CONCLUSIONS

The operational and analytic results of this series of tests led to the following conclusions:

- Operation of the system in both supercritical and subcritical Rankine Cycles was shown to be inherently stable.
- The geothermal brine supply to the system came from only one production well and flow was steady. The lack of greater geothermal production caused a lower than design temperature brine supply to the test system limiting test operations at heat duties more than 5% above design. However, this does not affect the validity of the heat exchanger performance demonstrated by the test operations.
- The chemistry of the geothermal brine remained constant throughout the duration of the tests. The tests demonstrated the generally corrosive character of this particular brine. This corrosive character eliminated those goals of the test program directed towards determination and evaluation of scaling and chemical methods of 'insitu' cleaning of heat exchanger surfaces.
- The overall measured heat transfer coefficients of the heaters were determined to an accuracy which varies depending on the factors arising from data uncertainty considerations. Generally, the smaller the temperature driving force (pinch point) between the hydrocarbon working fluids and the geothermal brine the less accurate the results. Run number 110 is used here to illustrate the potential combined error inherent in the data and is illustrated in Table 2-1 which summarizes both the measured and calculated results for this run.

Table 2-1

SUMMARY OF RUN NO. 110

Exchanger Unit No.	Brine Duty (10 ⁶ Btu/hr)	Hydrocarbon Duty (Q) Difference (%)	LMTD °F	Measured Overall Coefficient (U) (Btu/hr-ft ² -°F)	Percent		
					Uncertainty Q	LMTD	Error U
B1	3.37	-4.2	33.53	348.5	5.8	4.4	10.6
B2	2.20	-9.2	19.0	401.0	9.0	7.7	17.9
B3	1.55	-17.8	12.72	422.4	12.8	11.2	25.9
B4	1.24	-25.9	10.59	404.1	16.1	13.4	31.2
B5	2.14	-32.4	15.51	478.2	9.3	9.9	21.9
B6	5.05	-8.5	33.12	528.0	4.0	4.6	10.0

From Table 2-1 it is clear that the data from heat exchanger units B1 and B6 will be the most accurate. All the heat exchange units have the same tube and shell side configurations except for B6. B6 has the same number of tubes but has double segmental baffles while the others have single segmental baffles. Since these units were in series, the mass flow rate was identical in each unit, but the velocity of the hydrocarbon increased unit to unit as it was heated. The main difference in the performance of these units is that the heating curve of the hydrocarbon bowed, nearly meeting the nearly straight heating curve of the brine. The minimum temperature difference "pinch" for this run was 10.2°F occurring between units B2 and B3. There were other runs with pinches of less than five degrees Fahrenheit. This low pinch point temperature difference generally makes the accuracy of the data for B2 through B5 quite low. This type of problem is inherent in the working fluid and the precision of the instrumentation and is not due to the manner in which the operating data was taken.

Table 2-1 shows that the overall heat transfer coefficient varied from 348.5 to 528.0 Btu/hr ft²-°F. These values are consistent with values which were obtained with the PFR shell and tube computer program, SAT-1. This comparison is summarized in Table 2-2.

Table 2-2

COMPARISON OF HEAT TRANSFER COEFFICIENTS - RUN 110

<u>Unit No.</u>	<u>U-Measured</u>	<u>U-Calculated</u>	<u>% Difference</u>
B1	348.5	390.0	+ 11.9
B2	401.0	397.0	- 1.0
B3	422.4	449.1	6.3
B4	404.1	406.7	0.6
B5	478.2	480.2	0.6
B6	528.0	460.0	- 12.9

The low percentage difference from units B1 to B6 indicates that with organizations specializing in the design of heat exchangers, given the basic process, thermodynamic, and transport data, the heaters and condensers could be designed to better than 15% accuracy for East Mesa and other geothermal sites with brines which have similar, low fouling characteristics.

The pressure drop of the isobutane through the exchangers increased from units B1 to B5 because the density decreases thereby providing an increased velocity. The pressure drop for these units was calculated using the conditions of run 110, using the above mentioned computer program, and compared to the overall measured values. A value of 90 psi was calculated as compared to the 109 psi measured. Therefore the specification for this type of heat exchanger should allow for pressure drops about 20% higher than otherwise expected.

The effect of fouling on the heat exchanger was examined to determine its significance on the data. Fouling could have occurred on the brine or the hydrocarbon side. The overall heat transfer coefficients were plotted as a function of time. From this plot, it appears there is a drop off in the performance starting in December. After visual inspection of the tube side it was concluded that the brine did not foul. This leads to the speculation that there may have been fouling on the hydrocarbon side. Inspection of the hydrocarbon working fluid indicated the presence of a black substance with the appearance of iron oxide and water mixed with the isobutane. The effect of metallic fouling on the data runs 53-165 appeared to be insignificant. Since the heat exchange units are of fixed tube sheet design, there is no way of verifying this fouling until the shell sides are cut open.

The data from the highest temperature heat exchanger (B6) is extremely important because the isobutane was heated beyond its critical temperature. Theoretically it can be shown that heat transfer should dramatically increase at the critical point. The

units were not instrumented to obtain internal, step-wise data; therefore this effect could not be confirmed.

The overall condensing heat transfer coefficients, 250 Btu/hr-ft²-°F for isobutane and 190 Btu/hr-ft²-°F for 80/20 isobutane-isopentane mixture are within expected industrial range for the process conditions. However, this condenser data is of limited value for two major reasons. First, fouling on the cooling water side was so high that it lowered the overall heat transfer coefficients much lower than can be considered normal by industry. Second, the water temperatures were much higher than would ever be allowed in a power producing cycle.

The chemistry of the cooling water in the cooling tower was easily monitored but required chemical treatment to minimize calcium carbonate fouling. The cleanliness of this water was difficult to maintain due to the light sand and dust conditions in the Imperial Valley environment.

Control problems discovered during test operations were solved in the field primarily by manual adjustment to the controller set points. These control problems were due in one case to slag and machined chips left in the heat exchangers during fabrication, and in the other by excessive fouling caused by higher than design cooling water temperatures.

RECOMMENDATIONS

As a result of these tests the following recommendations are considered appropriate. These recommendations are divided into two groups. The first addresses both practices and problems anticipated in commercial plant operation. The second suggests modifications if further tests are run.

1. Relevant to Commercial Practice

- Current Heat Transfer procedures should be used to design and optimize the heaters and condensers.
- Temperature differences (pinch points) between hot and cold streams as small as 5°F are thermally stable and may be used during optimization of the overall plant and heat exchanger design.
- A sump or pit should be included by design to:
 - handle any discharges during plant cleaning and washing;
 - have the capacity to hold cold brine from the system during plant initial heat up;
 - serve as a reservoir for any fluids removed from the system during maintenance and repair operations.

- Primary brine/hydrocarbon heat exchangers should be two pass on the tube side (U tubes) to eliminate expansion problems and simplify tube side piping. Shell and tube heat exchangers with removable bundles (or shells) should be used so that both the shell and tube sides can be mechanically maintained and to allow the bundle to freely expand.
- Heat exchangers of varying size in the heater train should be considered. As a hydrocarbon is heated, the size and configuration of each exchanger can then be optimized.
- The control system for cooling water to the condenser should operate by sensing the flow of cooling water rather than condenser pressure or condenser outlet temperature.
- A knock-out pot or drum should be installed at circulation pump suctions to enable removal of any foreign material which may be circulating in the system.
- A reflex gauge glass should be installed on all cooling heat exchangers and at system low points to verify that no liquid accumulates.
- Drip legs should be installed on the condenser shell bottom to collect any solids or liquids which may collect at the bottom of the condenser. A reflex gauge glass should be installed on these drip legs.
- If any vertical pumps are used, a bleed line should be installed from the bottom of their pump outside housing to grade to enable removal of any impurities which may settle at that point.
- A different design should be used for the welded joint between the tube sheet and tubes in all hydrocarbon heat exchangers. The tube sheets should be beveled to accept a fillet weld. Seal welds are not sufficient. Manufacturing procedures should be considered which eliminate the possibility of the brine entering the hydrocarbon loop and vice versa; double-tube sheet designs should be considered.
- Close inspection should be performed during manufacture of equipment. Stress should be placed on onsite inspection during manufacture and cleanliness after testing in the fabrication shop.
- Inspection openings should be provided in each heat exchanger and vessel.
- All threaded valves and fittings in brine service should have exposed male ends with end caps. All female threads and plugs should be eliminated.
- Paddle type flow alarms appear to be vulnerable in this environment. Differential pressure switches across orifice plates in conjunction with a flow meter might have a greater reliability.

- A small brine bleed line should be installed to the hydrocarbon heat exchangers so a controlled (small) amount of brine (heat) could be admitted to the exchangers to "boil-off" any hydrocarbons to the condensing side when repairs to or maintenance of the heat exchange system is necessary.
- An online gas chromatograph should be considered in any hydrocarbon system. This would serve to provide ready recall of fluid composition and eliminate expansion problems encountered during extraction of mixed hydrocarbon fluid samples.
- Consideration should be given in using an appropriate corrosion allowance on the brine side of the heaters for the East Mesa site if carbon steel tubes are to be used. A value of five mills per year appears appropriate.

2. Relevant to Future Testing

- The cooling tower water bypass should be eliminated and all the cooling water should be cooled to its lowest temperature. Should various temperatures be desired for experimental purposes, a bypass should be installed directly to the cooling water circulation pump suction.
- All the turbine meters used in the system should be carefully reviewed; larger sizes should be considered. The pressure drop through the meters in the cooling water service was excessive.
- Differential pressure gauges should be installed across the orifice plates used for brine and hydrocarbon flow control. These can be easily read and used with the orifice plate curves to confirm flow rates.
- The brine emergency shutdown valve on the brine outlet from the hydrocarbon heat exchangers should be eliminated. The inlet emergency shut down valve is sufficient.
- A manual bypass should be installed in parallel with the inlet emergency shutdown valve so that valve may be "exercised" daily without interruption of the brine flow.

Section 3

SYSTEM OPERATIONS

OPERATIONAL SUMMARY

Field operations began on January 11, 1979 with the execution of an agreement between Colley Engineers and Constructors, Inc., the project contractor to EPRI, and Magma Power Company, lessor of the East Mesa Geothermal site from the Bureau of Land Management. This agreement, together with Colley's designation as "Operator" by Magma Power Company, provided access to the field site. Construction and erection of the test module began shortly thereafter. Timing was planned to allow completion of the binary and check out procedures prior to a mid June start-up. When it became apparent that brine delivery on this date was impractical, field efforts were slowed to minimize cost exposure.

In order to place the system operations in perspective, a time line is shown on Fig. 3-1. This time line indicates the events discussed in this section and places them in chronological order. Erection of the test facilities and the initial preparation of the site accounted for 25 of the 53 week period included in the time line. Nearly 17 weeks time accumulated while test operations awaited a supply of geothermal brine for test purposes. Seven weeks actual time was devoted to test operations. During this test period, four additional weeks were required to allow modification by others to the geothermal brine supply systems. As seen on the time line, the total of 22 weeks down time was not without activity. Activity proceeded all-be-it at a reduced rate. This activity is discussed in more detail in the sub-section entitled System Check-out and Start-up. The seven weeks of test activities are discussed in the section entitled Description of Operations.

Erection was essentially completed by July 6, 1979. The three fluid loops were cleaned, acid washed, and the brine and cooling water loops passivated. By July 28, 1979, all loops had been hydrotested and their erection accepted based on their successful completion of these hydro tests. Various operation trips were

Figure 3-1a. Field Activity Time Line

Figure 3-1b. Field Activity Time Line

Figure 3-1c. Field Activity Time Line

Field Activity Time Line

Figure 3-1

set and upon attempts to couple the hydrocarbon circulation pump on July 25, 1979, its shaft was found to be bent. It was subsequently removed and straightened and a functional test run on September 13, 1979. A notable earthquake, 6.4 on the Richter Scale, occurred on October 15, 1979. However, no damage was sustained. Finally on October 30, 1979, we were informed by Magma Electric that brine was available. The subsequent test operations were divided into three periods: start-up, acceptance, and heat exchange performance tests. Start-up and the acceptance tests were done generally in accordance with section two of the Test Plan governing the project operations and approved by both the Electric Power Research Institute and Lawrence Berkeley Laboratory. The heat exchange performance tests were run in accordance with section four of this same Test Plan.

A tabular summary of the Binary Loop Test Operations and Data History is shown in Table 3-1. The different test operations are referenced by their particular designation in the Test Plan. A graphic portrayal of the test operations, including the start-up period is shown as Fig. 3-2. Also indicated on this figure is the comparable brine supply status. It is interesting to note that during the first five weeks of testing, 23 test start-ups and shut-downs occurred. Thirteen of these were caused by problems with the supply of geothermal brine. Since the testing only relied on the flow from one of two geothermal production wells, this brine supply situation is not considered indicative of a larger, commercial plant operating history.

Start-Up

The start-up period began at 1100 hours on October 30, 1979, and ended at 2230 hours on November 16, 1979. During this period, operational and mechanical problems were found and corrected and some of the instrumentation relocated on the test set up to facilitate manual setting of the desired test conditions. The paramount problem encountered was the malfunctioning of the level gauge (sight glass) on the accumulator due to the mispositioning of its liquid line, excess-flow valve. This problem was not diagnosed until the end of the start-up period when it was felt that with additional hydrocarbon fluid, the hydrocarbon pressure in the condenser could be brought up to its prescribed level for testing and thereby enable the hydrocarbon design flow to be established.

In all, the start-up period included 372 hours elapsed time. Operations were conducted during 136 of these hours or 36.6% of the total time. In accord with the Magma agreement, 312 hours of brine supply was purchased. The operational time represents 43.6% of this brine time.

In all, 44 data sheets were completed. These data sheets served two functions. First they provided the operators with on line experience in taking and recording the test data, and second they served as a diagnostic tool in correcting the operational and instrumentation problems encountered.

Eight start-up and shut-down operations were accomplished. These are divided into four hot start-ups and four cold start-ups. A hot start-up is one accomplished within four hours of the prior shut-down. This four hour period is chosen since after this time, the systems temperature drops sufficiently so that the vapor pressure in the hydrocarbon loop falls below the normal operating pressure in the brine loop. Three of these hot start-ups occurred after the lower pressure rupture disc had blown and one after a high cooling water temperature trip had occurred. The four cold start-ups were after 9 shut-downs in accomodation of Magma's well operations. During these shut-downs, other repairs and instrument relocation was also accomplished.

Acceptance Tests

The Acceptance Test period began November 14, 1979 and ended November 28, 1979. Six of nine scheduled conditions were run. During this period the brine production was switched to Magma's well 48-7 which showed a much lower temperature (335°F) at the inlet to the heat exchangers than had been planned (350-360°F). This lower temperature inhibited further testing at higher hydrocarbon heat exchanger outlet temperatures (>300 F°) principally due to heat exchanger approach temperature problems. During this period of 323¹/₂ hours, testing occurred for 309 hours. This testing required brine to be produced for 312 hours. This testing occurred over 95.5% of the total elapsed time. Six hot start-ups were accomplished with two of these due to blown rupture discs. The other four were caused by brine production problems.

During the initial acceptance test condition, which lasted for 147¹/₄ hours, two cold start-ups occurred; one necessitated by repair to two test system control valves and the other due to brine production problems.

Table 3-1
OPERATIONS AND DATA HISTORY

Total Elapsed Time (hrs)	Operation	Time and Date		Test Time (hrs)	Number of Start-Ups		Number of Data Sheets Recorded
					Hot	Cold	
372.2	START-UP	1040 hrs	10/30				
		2230 hrs	11/14	136.00	3	7	44
323.50	ACCEPTANCE TESTS						
	A	2230 11/14 - 1030 11/21		147.25	0	2	63
	B	1030 11/21 - 1200 11/24		73.50	1	0	29
	C	1200 11/24 - 1200 11/25		24.00	0	0	12
	D	1200 11/25 - 0825 11/27		38.75	3	-	16
	E	0825 11/27 - 1030 11/27		2.00	0	0	-
	F	1030 11/27 - 0000 11/28		11.50	2	0	6
	G	- - - - -		--			-
	H	0000 11/28 - 0230 11/28		9.00	0	0	3
	I	0900 11/28 - 1230 11/28		3.00	0	0	-
		S/T		309.00	6	2	
169.50	HEAT EXCHANGER PERFORMANCE TEST						
	No. 1	1230 11/28 - 2400 11/29		30.00	0	1	18
	2	0100 11/30 - 0855 11/30		8.00	0	0	8
	3	0855 11/30 - 2400 11/30		15.00	0	0	8
	4	1200 12/01 - 2020 12/01		8.33	2	0	8
	5	2140 12/01 - 0500 12/02		7.33	0	0	8
	6	0500 12/02 - 0700 12/02		2.00	0	0	-
	8	0700 12/02 - 1650 12/02		10.00	0	0	8
	7	0700 12/02 - 0900 12/03		16.00	0	0	13
	9	0900 12/03 - 1630 12/03		7.50	0	0	8
	10	1630 12/03 - 0030 12/04		8	0	0	8
	11	0330 12/04 - 1300 12/04		9.50	1	0	8
	12,15	2300 12/04 - 1400 12/05		15.00	1	-	-
		S/T		136.66	4	1

Figure 3-2. Field test and brine supply operations.

The purpose of the acceptance tests was two-fold. The first condition was run sufficiently long to enable determination of the initial fouling characteristic of the geothermal brine in the brine to hydrocarbon heat exchangers. Considering that these heat exchangers had been cleaned prior to test operations, an increasing fouling factor with time was expected. However, this initial fouling factor was found to become increasingly negative during this test run. Consequently, this acceptance condition was satisfied when the rate of fouling factor decrease with time did not increase.

The second purpose of the acceptance test was to determine the operational stability of the three loops (brine, hydrocarbon and cooling water) at off design flow and temperature conditions. Difficulty was encountered in operating the loops at 120% of the design heat exchanger duty of 16.1 million Btu/hr. This difficulty was caused by the lower temperature brine requiring a disproportionately higher brine flow, which exceeded the brine system's capability of 200 gpm. Additional difficulty was encountered at heat exchanger duties of less than 60% of design with hydrocarbon top temperatures of 300°F or more. While the principal cause of these difficulties was the heat exchanger approach temperature (pinch point temperature difference), the lower duties also gave rise to cooling water flow regulation problems. The lower cooling water flow rates caused large temperature increases in the cooling water flow across the condenser with concurrent excessive fouling. While the fouling was beneficial to condenser operations (and detrimental to the condenser performance data), it also caused the condenser cooling water flow control valve to stick. Cleaning of this control valve's trim was then required prior to setting the next test condition.

In all, 129 data sheets were completed during this test period. The first 23 of these have been discarded due to faulty flow rate data. The subsequent data sheets are used for analysis as noted in Section 4.

Heat Exchange Performance Tests

The Heat Exchange Performance Test, in accordance with Section 4 of the Test Plan, began November 28, 1979 and were prematurely terminated on December 5, 1979. The elapsed time period totalled 169¹/₂ hours; test operations occurred during 134 ²/₃ hours or 80.6% of this period. This testing caused 192 hours of brine production, hence testing occurred during 71.2% of the brine production time.

Twenty-three (23) heat exchange performance tests had been devised to monitor the heat exchanger's performance at different hydrocarbon, or shell-side conditions. These included attempts to bracket the hydrocarbon tricondentherm between different heat exchangers. Eleven of these performance tests were completed prior to the premature termination of this test series; however, of these eleven one was cancelled due to the same low brine inlet temperature problem as noted during acceptance test operations. The termination of this test series was caused by the loss of hydrocarbon from the hydrocarbon loop. Three pin-hole leaks were discovered in a tube to tube sheet weld at the brine outlet end of the heat exchanger train. The loss of hydrocarbon through these pin-hole leaks was not detected since the level gauge was inoperative due to the faulty orientation of its liquid line excess flow valve. The loss of hydrocarbon lowered the hydrocarbon liquid/vapor interface causing low pressures and high temperatures at the inlet to the hydrocarbon circulation pump. This condition limited the hydrocarbon flow required for a given pump discharge pressure. Without the ability to maintain test set conditions, further testing could not continue.

During the heat exchangers performance test period, four hot start-ups and only one cold start-up occurred. These hot start-ups were caused by the blowing of the lower pressure rupture disc. Under low hydrocarbon flow conditions, attempts to manually set the hydrocarbon throttle valve control point would easily cause a rapid hydrocarbon pressure increase, which increase terminated with rupture of this disc. The one cold start-up occurred after an area-wide loss of power. This loss of power lasted approximately 5¹/₂ hours. A large electrical surge in the Imperial Irrigation Districts system at the time power was lost apparently caused a large reverse torque on the hydrocarbon pump. The pump's seal lifted and leaked isobutane at a small but increasing rate until it was replaced January 9, 1980.

One hundred five data sheets were completed during the performance test period. The last ten data sheets taken just prior to the premature termination of this first series are not considered valid. A total of thirty-one data sheets were completed for test runs numbers one and seven. These two test conditions are at the systems design point. This design point also prescribed the test conditions for the first of the acceptance test runs. Hence identical test conditions were run three times during these test operations with 162 and 51 hour periods intervening.

Subsequent Test Operations

Additional testing was held in abeyance until after January 1, 1980 to allow Magma to revise their brine supply piping. When the heat exchange performance tests were terminated, the desuperheater, condenser, and subcooler water-box covers were removed. This showed that extreme fouling of all surfaces, hot and cold, had occurred. Prior to restart, the desuperheater, condenser, and the subcooler water sides were cleaned by hydroblast by Southwest Chemical Company of Brawley, California. This cleaning was accomplished on December 11, 1979.

On December 12, 1979, three pin-hole leaks were found in one tube to tubesheet weld at the brine outlet from the coldest brine to hydrocarbon heat exchanger (B1). The evidence that there was communication from the hydrocarbon side of the heat exchangers to the brine side was the equilibrium of pressure on both sides. Normally, under cold shut down conditions, the hydrocarbon pressure would be about 50 psi greater than the brine side pressure. All the brine, water-box covers for the brine to hydrocarbon heat exchangers were removed; no other leaks were found. The removal of these covers showed extremely clean surfaces on the brine side. This was indicative of a general state of corrosion in support of the initial decreasing trend noted for the brine side fouling factor.

On December 13, 1979, these pin-hole leaks were peened and a weld bead over-layed around the tube to tubesheet weld. An additional 2450 pounds of isobutane was loaded into the hydrocarbon system by January 2, 1980. The brine side water-box cover, covering the three fixed pin-hole leaks was removed and using the hydrocarbon pump, approximately 500 psi differential pressure was applied across this tubesheet. No evidence of leakage was found; consequently, testing was set to resume at the test conditions corresponding to run number one of the the heat exchange performance tests.

Subsequent difficulties were encountered in obtaining a supply of brine from Magma. These difficulties were due to their own, low-level requirements as they prepared their plant for start-up. Given this condition, recognizing that limited funds for testing remained, and recognizing that no heat exchanger fouling data would be obtained with the generally corrosive brine, it was decided to obtain heat exchanger performance data on hydrocarbon mixtures of isobutane and isopentane independent of the program outlined by the Test Plan.

Accordingly, successive heat-ups followed by cool-downs were begun January 14, 1980 to assure removal of a maximum amount of non-hydrocarbon fluid from the hydrocarbon loop. Two hundred fifty (250) pounds of isobutane was removed and 310 pounds of isopentane added. Four different test conditions were run for 24 hours from January 15, 1980 to January 16, 1980. An additional 385 pounds of isopentane was added to the hydrocarbon loop on January 16th and three of the same four test conditions run until all test operations were shut down at 1700 hours that afternoon.

In all, 24 hours of test operations were conducted with approximately a 90/10 percent molar mixture of isobutane/isopentane and six hours of test operations conducted with approximately an 80/20 percent molar mixture. No other start-ups or shut-downs occurred during any of these test runs.

Sampling of the hydrocarbon fluids was done during runs with each mixture. Vapor samples were taken to allow determination of the water content in the hydrocarbon fluid, and liquid samples were taken to enable determination of the mixture composition. This latter determination showed that the January 2, 1980 hydrocarbon recharge had been contaminated by propane.

However, after careful, subsequent analysis, the hydrocarbon mixtures were established as shown on Table 3-2. These mixtures were used for analysis of the data taken. Sixteen data sheets were completed for the 90/10 nominal mixture; five data sheets completed for the 80/20 nominal mixture. The results of analysis of the data taken are presented in Section 4.

The hydrocarbon fluid was removed from the hydrocarbon loop on January 17, 1980 and this loop flushed with 2250 SCF of nitrogen and secured on January 18, 1980. The water loop was drained on January 17th with the condenser and subcooler water-box covers removed and their tubes blown dry with air on January 18th.

Table 3-2

HYDROCARBON MIXTURE COMPOSITIONS

Period	Nominal Mole Percent (Liquid)		
	Propane	Isobutane	Isopentane
1/2/80 - 1/14/80	4.8	95.2	--
1/15/80 - 1/16/80	4.3	85.8	9.9
1/16/80	3.8	76.5	19.7

DESCRIPTION OF OPERATIONS

Stability

The first supercritical operation of the Binary Equipment Module Test was achieved on the evening of November 14, 1979. The capability to operate supercritically requires an inventory of hydrocarbon fluid sufficient to maintain a liquid-vapor interface in the accumulator located between the condenser and the subcooler. While the subcooler has the capability to condense the hydrocarbon vapor, which service it saw during the start-up period, its performance as a condenser left the hydrocarbon temperature too high and the hydrocarbon specific gravity too low, inhibiting operation of the hydrocarbon circulating pump. Accordingly, with the characteristic available from the hydrocarbon pump, insufficient head was developed across this pump to enable supercritical conditions.

Operation of the three Binary Module Test loops was found extremely stable. This stability is seen from the trace of pressure and temperature as shown in Figure 3-2. Test runs shown on this trace include both supercritical and subcritical brine to hydrocarbon heat exchanger outlet conditions. During design of the Binary Module Test, consideration was given to possible operating instabilities which would affect the stability of test operating conditions. Accordingly, provision was made for two bypasses installed on the cooling water loop, an accumulator enabling hold up time equal to about 15% of the hydrocarbon loop circuit time, and no accumulation or by-pass provision for the geothermal brine loop.

The two by-passes in the cooling water loop enabled the cooling tower basin, and hence the cooling water at the inlet to each heat exchanger, to be kept at an even temperature of about 90°F. One by-pass allowed hot water from the return to the cooling tower to be by-passed directly to the cooling tower basin. The second by-pass allowed cooling tower basin water to pass directly to the hot water return to the cooling tower, eliminating excessive carryover from the cooling tower. Both bypasses were used during test operations. The prime by-pass was the hot water by-pass to the cooling tower basin. It was automatically controlled and during operations held the tower basin temperature to within $\pm 1^\circ\text{F}$ of the set point. This provided for stable operation in the hydrocarbon to cooling water heat exchangers eliminating any operating instabilities arising from diurnal swings in wet bulb temperatures. Once the hydrocarbon fluid level was established above the subcooler, no flow or temperature transients were observed in the hydrocarbon loop.

The design hydrocarbon inlet temperature to the hydrocarbon pump was 120°F. With the subcooler full of liquid, this temperature was easily maintained; the pressure at the suction of the hydrocarbon pump was independent of the hydrocarbon temperature and the pump discharge pressure increase for the design flow easily established.

No operational instabilities were found in the geothermal brine to hydrocarbon heat exchangers due to thermal or hydraulic efforts. This is true for thermodynamic regimes including boiling (sub-critical) or transition across the tricendtherm (super-critical).

While no attempt was made with the instrumentation system to monitor short term transients, the time for the hydrocarbon to respond to a one degree Fahrenheit step change in the brine inlet temperature was estimated at about 30 minutes. Similarly, the response time for a mercury, glass bulb thermometer in its thermowell was estimated at about 50 minutes. An investigation of the temperature and flow rate scatter as shown in Figure 3-2 shows that no short term thermal or hydraulic transients existed. Indeed, no noise level change was noted during operations. No provision was made during either design or construction to mitigate any perturbations which might occur in the geothermal brine loop. One of the prime objectives of the Binary Module Test was to investigate the operational interface between the performance of a geothermal brine delivery system and the hydrocarbon energy extraction system. Variations in the temperature and production rates have been noted by others with geothermal well flows. Since the

geothermal wells providing production for the tests were to be pumped, it was felt that these variations would be minimal. No attempt was made to monitor or record any possible variations. Investigation of the heat exchanger brine inlet temperature and pressure shows the only variations to be from well to well, not from any one particular well. In the geothermal brine supply system, Magma Electric Company had installed a sand trap. This sand trap was also used to collect and vent excess non-condensable gases which may come out of solution in the geothermal brine with the pressure decrease from bottom hole to a brine booster pump. While it is possible that this sand trap dampened any hydraulic fluctuation, no evidence exists to show there was any fluctuation to dampen.

Performance Limits

As noted, performance was exceptionally stable at design conditions. Test operations were planned from 30% to 120% of design flow, from 300°F to 320°F in hydrocarbon exit temperature, and from 510 to 585 psig in the heat exchanger hydrocarbon exit pressure. Practical limits to operations were found below 60% of design hydrocarbon flow (60% of design duty) and to hydrocarbon duties (heat delivered) about 5% greater than the design duty of 16.1 million Btu/hr.

These practical limits existed for several reasons:

1. At low hydrocarbon flow rates and with the design hydrocarbon exit temperature, the low flow rate of cooling water was difficult to regulate through the condenser. This flow rate regulation valve was a six inch valve designed to pass 680 gpm. Due to the higher than expected isobutane condensing coefficient and the approximately four times excess area in the condenser, much higher cooling water temperature rises and lower flows were experienced. The condenser cooling water flow was typically about one quarter of its design rate. Hence, this control valve was operating at the bottom of its control range. As noted elsewhere, problems of fouling from the higher than normal top cooling water temperature further compounded the control problem. Regulation of this flow became impractical below this lower flow limit.
2. Two practical operational constraints became apparent for tests with heat transfer duties above the design duty of 16.1 million Btu/hr. The design inlet temperature for the geothermal brine was 360°F. Normal brine inlet temperatures experienced ranged from 335°F to 350°F. Since Magma Electric was not concurrently operating their plant and the heat loss during transmission of the brine to the Binary Module Test occurred at a nearly constant rate, a greater heat loss per unit of brine flow occurred. This problem was compounded because the brine also spent a longer period of time in transition. This lower brine inlet temperature required a higher flow rate to deliver the

same duty to the brine to hydrocarbon heat exchangers, and this, in turn, causes a higher brine exit temperature from the heat exchanger train. The brine heating curve is thusly rotated providing a lower mean temperature difference at the higher temperature end of the heat exchange train where about 60% of the duty occurs. This decrease in brine inlet temperature also decreases the temperature difference at the pinch point occurring on the hydrocarbon heating curve. Both these effects result in a decrease of the hydrocarbon flow rate for a given enthalpy rise. Hence, while several of the heat exchanger performance tests required hydrocarbon exit temperatures of 310°F and 320°F, these test conditions could not be achieved with the design hydrocarbon flow rate of 95,150 pounds per hour. These test conditions were run, but with approximately 80% of the design hydrocarbon flow rate.

Control

Seven controllers were used to control the system operations. One controlled the geothermal brine flow rate at the brine exit from the brine to hydrocarbon heat exchangers. Two were used to independently control the hydrocarbon flow rate and heat exchanger exit pressure. Three were used to control the cooling water flows through the hydrocarbon desuperheater, the condenser, and the subcooler; and one was used to control the cooling water temperature out of the cooling tower basin. Each controller had the capacity to automatically maintain a preselected process value or to allow each process value to be set by manual adjustment to the controller air output pressure.

The only controller to be always used in the automatic mode was that controller controlling the temperature of the cooling water out of the cooling tower basin. The controller controlling the hydrocarbon outlet temperature from the subcooler by regulating its cooling water flow was used in the automatic mode once a hydrocarbon liquid-vapor interface was established in the accumulator.

Operations were found to be so stable, and automatic control unnecessary, that the other five controllers were used in the manual set mode.

Referring back to Figure 3-2, the geothermal brine conditions were nearly constant. No control capability was needed beyond a manual adjustment of the flow control valve to set the desired brine flow rate. The flow of hydrocarbon was likewise nearly constant, requiring no automatic control. Manual adjustment of the throttle valve was found necessary due to the weld slag and machined chips which periodically became caught in this valves' trim. During the time attempts

were made to allow this valve to automatically adjust to its upstream pressure conditions, an increasing air pressure on the controller diaphragm caused the valve adjustment to overshoot resulting in a burst rupture disc. Consequently, to achieve fine control over the hydrocarbon pressure, it was decided to change the controller air output pressure manually and then mechanically aid the movement of the control valve stem. This procedure was found successful. Since there was no variation in the hydrocarbon pressure, no further need existed to refine the control valve operation to the automatic control mode.

Severe control problems were encountered with two controllers, one controlling the hydrocarbon exit temperature from the desuperheater and the other the hydrocarbon condensing pressure. Both control valves exhibited full stroke swings while trying to maintain the preselected process value. These swings were caused by attempts to control the process value to a point about which there was poor regulation. In the former case, sufficient regulation occurred above the vapor saturation temperature and zero regulation occurred below the saturation temperature. In the latter case, no regulation exists by definition across a fluid's thermodynamic saturation dome. In both cases the selection of the control variable was incorrect. In both cases, the control variable should have been the water temperature out of each of these two exchangers. Consequently, operations dictated that control be effected by manual adjustment to the controller air output pressure. Again, operations were so stable in this mode, there was no need to revise the control scheme to enable control in the automatic mode. Periodically each of these valves needed mechanical aid in moving from one set point to another due to the build up of fouling on their trim from the high cooling water outlet temperature.

SYSTEM CHECKOUT AND START-UP

Completion operations and checkout of the Binary Module Test began with the end of erection, July 6, 1979. This period lasted until start-up began October 30, 1979, with the first "all-up" systems test. Acceptance of the completed brine and hydrocarbon fluid loops was based on their successful completion of hydrostatic testing at pressures of 1.5 times their maximum operating pressure, 563 psig and 953 psig respectively. No hydro test was required of the cooling water loop.

Acceptance of the hydrocarbon loop required that the test pressure, corrected for any temperature change, show no change during a one hour period and that no leakage be observed at any weld, flanged joint, valve stem packing, pump packing,

or equipment or pipe surface. Acceptance of the brine loop simply required no more than 1% hr. decay in indicated pressure (less than 5.6 psi/hr.) after temperature correction. Verification of the flow direction within each loop was presumed until start up given correct rotation of the two circulation pumps. All control valves were witnessed to cycle over their full range and the fail open or fail closed characteristics visually verified. In addition, each emergency shut down (ESD) system signal was shown to independently cause each ESD ball valve in the brine loop to close. An initial hydro test of all three loops was used on June 12, 1979 to indicate any gross system erection problems and provide the basis to direct completion of the erection process. Both the brine and cooling water loops were accepted at this time. The 563 psig hydro of the brine loop, however, exceeded the rated capability of the brine side of the brine to hydrocarbon heat exchangers. Accordingly, new brine side heat exchanger covers and a gasket system were devised by ThermXchanger, their manufacturer, the channel flanges remachined in the field, and the covers replaced and acceptably re-hydroed on July 20, 1979 at a pressure of 890 psig.

On June 22, 1979, 565 psig pressure of nitrogen at 125°F was applied to the hydrocarbon loop. This enabled verification of the welds, flanges, gaskets, and the hydrocarbon pump seal all of which exist within the two isolation valves surrounding the hydrocarbon pump. Additional leaks were found and fixed. On July 18, 1979, the hydrocarbon loop was successfully hydroed at 985 psig to within the requirements of the Test Plan. One additional leak was found in the casting for the isolation valve on the downstream side of the hydrocarbon turbine flow meter. The valve casting was peened and the leak stopped. In accordance with the plan prepared to direct the preparation of the fluid loops for test operations, four drums of inhibited oil well acidizing fluid had been procured from Hallibuton Company. This acid is a 15 weight percent hydrochloric acid solution in 50-60 weight percent acetylenic alcohol and 10 weight percent formaldehyde. This solution was used to remove any grease or oil and mill scale from the pipe and equipment surfaces in each fluid loop. In order to check the effacacy of this inhibited acid and to check for any undue corrosion on the steel heat exchanger tubes, six samples of the heat exchanger tubes were cut and delivered to GHT Laboratories, Inc. of Brawley, California for weight loss determination. The acid solution was diluted to about a $2\frac{1}{2}$ weight percent hydrochloric acid strength (one part acid into five parts water) and the six samples immersed for varying periods up to 180 minutes. A picture of these samples after this immersion is shown as Figure 3-3. The results of the weight loss determination are shown in Figure 3-4; negligible weight loss occurred after 30 minutes immersion and the

weight loss that did occur was equivalent to 0.0015 pounds/ft.² or less than 4 percent of one mil in thickness. Hence it was concluded that the inhibitor was extremely effective and use of this acid as diluted presented no corrosion hazard to the pipe or equipment. This testing was done on July 3, 1979, the temperature of the dilute acid solution about 80°F.

On July 10 and 11, 1979, acid washing of the three fluid loops was accomplished. Since the brine and cooling water loops would be subsequently exposed to air, an additional passivation of these loops' surfaces was done with a zinc chromate solution to control possible oxygen corrosion. A summary of the procedure is shown on Table 3-3. A portable, five horsepower gasoline engine driven pump was used to circulate the wash and passivating fluids at about 50 gpm. The systems average temperature during these operations was approximately 85°F.

Cleaning connections were provided on each loop. Cleaning of the cooling tower itself, together with the main cooling water circulation pump was not done due to the galvanized metal construction of the cooling tower. Likewise, the cleaning connections for the hydrocarbon loop were located to enable cleaning of all this loop except that part within the two hydrocarbon pump isolation valves. Subsequent to the fluid circulation and soaking, each loop was flushed with water until a pH of approximately 6 was measured. The indicated amount of zinc chromate was then added to the flush water for continued circulation and soaking. At the conclusion of the passivation treatment, the used solution was discarded.

Figure 3-3 Tubular Coupons After Weight Loss Test

Figure 3-4. Weight loss test results.

TABLE 3-3
ACID WASHING AND CHROMATE PASSIVATION PROCEDURE

	<u>Brine Loop</u>	<u>Hydrocarbon Loop</u>	<u>Cooling Water Loop</u>
System volume, gals.	200	1100	400
Fluid volume, gals.	250	1150	500
Acid dilution from 15 w/o HCl	1/4.67	1/7	1/9
Circulation time (@ 50 gpm)	30 Min.	60 Min.	60 Min.
Soak time	30 Min.	-----	60 Min.
Exposure time	60 Min.	60 Min.	120 Min.
ZnCrO ₄ , 45 w/o solution	1 Gal.	-----	3 Gal.
Circulation time (@ 50 gmp)	30 Min.	-----	30 Min.
Soak time	30 Min.	-----	30 Min.
Exposure time	60 Min.	-----	60 Min.

The hydrocarbon pump barrel was acidized for cleaning August 8, 1979. Five gallons of the 15 w/o HCl were added to 56 gallons of water as the pump barrel was filled. This was allowed to stand for about one hour after which it was flushed with fresh water until a pH of 6 was measured. A vacuum truck was used to pull the flush water from the pump barrel.

On July 24, 1979, instruments from the process instrumentation system were used to calibrate and set trip points in the emergency shutdown system. The high pressure trip in the hydrocarbon loop, located downstream from the highest temperature heat exchanger was set at 650 psig, the inlet cooling water temperature trip was set at 120-124°F.

On July 25, 1979, attempts to couple the hydrocarbon circulation pump to its motor lead to the discovery that the pump shaft had arrived at the site bent. This pump and motor were removed from the system and taken to Process Equipment Company, Los Angeles. A new pump shaft was installed and the pump and motor returned to the

site and reinstalled August 7, 1979. This pump shaft and motor shaft were successfully coupled to within the manufacturer's tolerance of 5 mils, TIR. From this point on, start-up appeared imminent. Consequently, a total of 761 (3260 lbm) of isobutane was loaded into the hydrocarbon loop by September 7, 1979. On September 9, the low level emergency shutdown system trip located at the accumulator was found not working. This low level trip was replaced with a pneumatic level switch on September 18, 1979. Meantime, a cold circulation test was run on the hydrocarbon pump, September 13, 1979. A lack of electric signal continuity was found in the hydrocarbon turbine meter circuitry. Otherwise all safety systems appeared properly functional and all flow directions verified.

A functional check of the cooling water systems had been made August 8, 1979 at which time a pump curve (Figure 6-29) was field determined. Several problems were identified at this time. Of prime concern was the higher than anticipated flow resistance of the four cooling water turbine flow meters. While this served to decrease the overall cooling capability of the cooling water system, the system's initial over design margin was decreased and the flow deemed sufficient for test purposes. Also at this time and again on October 29, 1979, one of the two cooling water low flow trip switches were found to be broken. Stiffening of the paddle assemblies enabled this switch to continue functioning in the Emergency Shut Down System. The cooling water system flow directions, control, and instrumentation were verified acceptable at this time.

The initial production of geothermal brine for the benefit of the Binary Module Test was scheduled for September 19, 1979 by Magma Electric Company. However, on the morning of September 19, 1979 one dike in Magma's main deep cooling pond was discovered damaged and the planned start-up postponed until repairs to this dike were made.

A joint inspection of the condenser tubes was made by the Electric Power Research Institute and the Lawrence Berkeley Laboratory on October 19, 1979. This inspection showed small traces of what appeared to be "gutter" dirt inside the bottom of each tube. This indicated that the rate of cleaning fluid circulation may not have been high enough or that the collection of dirt by the cooling tower was higher than anticipated. (Each cooling tower in similar locations should include capability for sand and dirt removal during operation. This could either be by a "shaker table" technique or by a fluid cross ruff caused by a properly shaped cooling tower basin floor.) Accordingly, plans were made to enable condenser tube cleaning by means of a steel wire brush when analysis of the condenser performance

indicated the condenser fouling factor exceeded design or would exceed design during the next test run.

Magma Electric Company provided notification at 10:40 AM on October 30, 1979 that geothermal brine was now available for the System Test start-up. Consequently, the brine supply and return to Magma Electric's System were filled and pressurized at system operating pressures. A leak check showed no leaks and the brine to isobutane heat exchangers were filled. A start-up was attempted but quickly resulted in the bursting of the lower pressure rupture disc installed around the hydrocarbon throttle valve. At low hydrocarbon flows, adjustment of this throttle valve was very sensitive and difficult. This rupture disc was quickly replaced and the start-up continued.

Initially, the six controllers were pre-set to the desired start-up condition operating point. No attempt had previously been made to determine the correct control valve setting for the given flow, pressure or temperature condition. Accordingly, the pneumatic controllers began competing, causing the control valves to cycle, and increasing the air consumption rate beyond the capacity of the air compressor. The air compressor starter overheated and tripped; the system automatically shut-down.

On October 31, a new and smaller diameter pulley drive wheel was obtained and installed on the air compressor motor to reduce the electric load drawn by the air compressor. Additionally, the controllers were removed from their automatic control mode and manual adjustment of the controller air output pressure used to set the control valves in their appropriate start-up operating position. This start-up was very smooth. The initial loading of isobutane had indicated the proper operating level of isobutane was reached with about 500 gallons of isobutane inventory. Consequently, with the smooth start-up, 220 gallons of isobutane were removed and replaced in four of the bottles from which it had been loaded.

This removal of isobutane from the hydrocarbon loop was an unfortunate mistake. Although not realized at that time, the level gauge at the accumulator was giving erroneous readings due to a mal-positioned excess flow valve in its liquid leg. This valve was installed with the handle pointing down. If the handle were pointing up, the excess flow check would be open allowing flow communication. With the handle pointing down, no flow occurred. This problem was not discovered until the internals of this valve were removed in Mid-December.

The start-up period ended November 14, 1979, at which time, it was deduced from the inability to raise the condenser pressure, and hence the inability to reach system design pressure, that insufficient hydrocarbon inventory existed in the system and that the level gauge readings were not accurate. Accordingly, two of the four bottles of isobutane that had been removed from the system were replaced into the system. This resulted in the first super-critical operation of the system at 2300 hours that evening. The level gauge was not used or monitored after November 14, 1979.

A sufficient inventory of hydrocarbon fluid is the key to successful stable operation of the system. A lack of sufficient inventory required condensation to occur in the subcooler causing an increasing hydrocarbon circulating pump suction temperation with increasing pump discharge pressure. Had the isobutane inventory been sufficient, the 14 days spent on start-up would have been reduced to two or three days.

During the start-up period, several other operational problems became apparent:

- The measured pressure difference across the hydrocarbon side of the brine to hydrocarbon heat exchangers was found to be about 100 psi. The design pressure drop had been calculated as 50 psi, and consequently five of the six differential pressure gauges were rendered inoperative due to stretched diaphragms.
- A brine surge during a start-up, due to the noncondensable gas content of the geothermal brine served to damage four of the brine to hydrocarbon heat exchanger brine side differential pressure gauges. The remaining two of these gauges were removed to accurately measure the pressure difference across the brine and hydrocarbon orifice plate flow meters.
- Without all of the hydrocarbon side differential pressure gauges, a complete set of heat exchanger data was impossible. However, assumptions made as to the distribution of the overall pressure drop among the five lower temperature heat exchangers, based on the square of the isobutane specific volume, and together with the differential pressure accross the double segmetal baffled heat exchangers, allowed for on-site heat exchange performance analysis.
- The three Ashcraft digigauges supplied by the Lawrence Berkeley Laboratory suffered stretch movements and burst diaphragms due to leaky instrument valves in their sensing manifold systems, as assembled. Consequently, this data, absolute pressure measurements at four hydrocarbon locations and three geothermal brine locations, and use of the geothermal brine venturi flow meter were lost for the duration of the tests. The process instrumentation, however, enabled this particular data to be obtained with slightly impaired accuracy.

- Operational difficulties were found with the condenser. The hydrocarbon condensing coefficient was quickly found to be nearly twice the expected value of about 180 Btu/hr ft²°F, and the total tube area about twice again more than needed. Consequently, at low hydrocarbon flow conditions, the flow of cooling water was also low giving rise to problems of controllability. Additionally, the cooling water experienced a large temperature rise which caused the water side surfaces to foul at a much greater rate than originally anticipated. Coupled with the high cooling water outlet temperature was a higher cooling water return temperature to the top of the cooling tower. This presented a flow balance problem in the cooling water loop, and adverse fouling of the cooling tower. Accordingly, no condenser performance data was possible at other than design flow conditions. A related problem was the buildup of basically a carbonate scale on the surfaces of the superheater and condenser cooling water control valve's trims. This required disassembling of these valves for cleaning and remachining of their trim to increase their tolerance to this fouling.
- Machined metal chips and weld slag collected in the hydrocarbon throttle valve trim inhibiting any adjustment of this valve. The acid waste and cleaning operations had been formulated to provide a normal commercial cleaning of all surfaces. The existence of these chips and slag required the system to be shut down and the isobutane fluid removed twice from the system in order to clean the throttle valve. Approximately 24 hours were required to shut down and empty the system so this problem was time consuming. While enforcement of manufacturing cleanliness may decrease the magnitude of this problem, a trap for these metal pieces should be provided in front of each critical piece of equipment.
- The control valves controlling the cooling water flow through both the desuperheater and the condenser were found to cycle from full open to full closed. This caused excessive air consumption and occurred because the choice of control scheme and set point was improper. While this problem was easily corrected in the field through use of manual adjustment of the controller air output pressure, a proper control scheme would recognize the lack of regulation afforded by a constant temperature or pressure saturation line and would regulate the flow of cooling water based on its temperature at the water outlet from each exchanger.

SAMPLING

Sampling of geothermal brine was accomplished using two Whitey 500 ml sample bombs each isolated by a 1/2 inch Hills McCanna ball valve (F602 FS6F) at each end. Sampling of the hydrocarbon working fluid was accomplished using a two section sample bomb to enable a pressure reduction of the hydrocarbon fluids prior to sample extraction for analysis. The second section of this hydrocarbon sample

bomb contained a rubber septum holder connected by a 'tee' between two Hills McCanna ball valves. Pictures of both these sample bombs are shown on Figure 3-5.

To characterize Magma's geothermal brine production, a total of four samples of Magma's production from wells 44-7a, b and c were obtained on 5/23/79. The first sample bomb loads, about 250 ml each showed

pH	Temperature
5.6	117°F
5.7	118°F
5.75	140°F

These three bomb loads were comingled into one 500 ml sample (101) for submittal to GHT Laboratories, Inc. of Brawley, California. The fourth sample (101 NCG) was submitted to GHT for noncondensable gas analysis.

A total of seven samples of Magma's production from wells 44-7 a and b were obtained on 5/25/79. The first three showed a pH of 5.7 at 100-116°F. The next group of three showed a pH of 5.7-5.8 at temperatures of 118-130°F. These were comingled as 500 ml samples (102 and 103, respectively) for submittal to GHT. In addition, sample 102 NCG was obtained this same day.

Subsequent samples of production from Magma's south well pad (wells 48-7 and 48-7a) were obtained courtesy of Magma. These samples and a sample of the south pad production down stream of the sand separator were analyzed by GHT during July 1979. Additional check samples of Magma's north pad production were taken July 12th. No further brine sampling was done until start-up October 31, 1979. The final brine sampling was done January 15, 1980. The results of sample analysis by GHT Laboratories, Brawley, California are reported on Table 3-4. The results of GHT's analysis of the noncondensable gas sample is also reported on Table 3-4.

Figure 3-5 Brine and Hydrocarbon Sample Bombs

The sample analyses listed on Table 3-4 indicate very little change in brine composition from well to well. Sample 101 is high in chlorides due to recent work on well 44-7; the well had not yet managed to clear itself of the weighting fluid (principally NaCl) used during well rework. The X sand separator analysis indicates that the sand separator, installed by Magma to trap any sand produced along with the geothermal brine, had little effect on the geothermal brine chemistry. The pH indicated for both the X sand separator analysis and that listed as 48-7, 7a was determined by GHT. No immediate determination of this was made; and, due to the time lag between sample collection and analysis, it is not considered representative. This last comment is also applicable to the two samples labelled as 107.

Further analysis of GHT's results was done by Battelle Northwest Laboratories using the 'Equilib' computer program developed for EPRI to help predict geothermal plant operational problems due to scaling caused by the geothermal brine production.

Battelle reported in June 1979, that no scaling or brine side fouling was anticipated due to use of the geothermal brine as reported by samples 101, 102, 103 and 101NCG and 102NCG. Their determination can be documented by observing the relatively clean heat exchanger water boxes shown in Figure 3-6. These pictures were taken December 12, 1979 after approximately 700 hours of test time had accrued.

In January 1980, verbal advice was received from Battelle indicating that by their own sampling and analysis, it appeared that the brine was mildly corrosive. They had found a greater concentration of iron (Fe) in the brine flow at the injection well head than at the production well head. Accordingly, samples of the brine, labeled 107, were obtained from both the brine inlet to heat exchanger B6 and the brine outlet from heat exchanger B1. These two samples were taken sequentially and show approximately a doubling of the iron concentration between these two points. Consequently, it may be advisable to periodically caliper on otherwise access for inspection all brine exposed surfaces.

Table 3-4

SUMMARY OF GHT LABORATORY GEOTHERMAL BRINE SAMPLE ANALYSES

	101 5/23/79 approx. 200 feet down- stream of wells 44-7a,b,c	102 5/25/79 approx. 200 feet down- stream of wells 44-7a,b,c	103 5/25/79 approx. 200 feet down- stream of wells 44-7a,b,c	X sand separator 7/2/79 downstream from sand separator production from wells 48-7,7a	48-7, 7a 7/2/79 comingled well head samples at wells 48-7,7a	104 7/12/79 approx. 200 feet down- stream of well 44-7a	105 7/12/79 approx. 200 feet down- stream of well 44-7a,b	106 10/31/79 at brine inlet to heat exchangers production from 48-7 well	107 1/15/80 at brine inlet to heat exchangers production from well	107 1/15/80 at brine outlet from heat exchangers
TDS	11354	7821	7696	7836	8272					
pH @ °F	5.7 @ 127	5.7 @ 113	5.75 @ 124	7.2 @ 80	7.15 @ 80	5.8 @ 140	5.8 @ 120	5.75 @ 130	6.4 @ 80	6.4 @ 80
HCO ₃ (Mg/Li)	362	375	432	337	334	350	422	300	386	302
Cl	6353	4281	4256	4500	4293	5360	4612	4800	4750	4810
SO ₄	774	115	76	62.5	63.4	74.1	105.3	-	91	65
SiO ₂	277	285	263	232.9	233.3	246.9	255.6	255	230	257
Fe	5.9	13.7	11.4	16.2	1.72	.85	trace	7.3	12	25.5
Sr	29	6.5	6.3	11.1	14	19.4	11.8	22	20	18.9
Ba	3.9	1.2	.9	trace	trace	ND	ND	trace	2.0	1.4
NH ₄	15.9	14.6	16.6	17.8	15.6	18.3	20.3	39	31	31
B	7.1	7.6	7.6	8.50	8					
Al	trace	trace	trace	ND	trace					
Mn	.20	.07	.09	.031	.023					
Zn	.05	.05	.02	.035	.014					
Sb	ND	ND	ND	.045	.042					
Bi	ND	ND	ND	ND	ND					
As	.26	.29	.40	.368	.292					
Mg	4.1	.72	.68	1.56	1.28					
Ba	3.9	1.2	.9	trace	trace					
Cu	.02	.03	.03	.011	.015					
Pb	.41	ND	ND	trace	ND					
Sn	ND	ND	ND	trace	trace					
Li	12.2	7.3	6.9	9.11	9.32					
Na	3600	2630	2650	3100	2800					
K	390	275	270	290	270					
Ca	186	38	38	57	47					
	101 NCG	102 NCG			48-7a					
H ₂ , % by volume	.29	1.68			-					
O ₂	.92	1.45			2.18					
N ₂	9.35	7.65			5.15					
CH ₄	3.80	4.40			6.06					
C ₂ H ₆	trace	ND			0.099					
CO ₂	85.54	84.88			86.16					
H ₂ S	ND	ND

Figure 3-6 Brine to Hydrocarbon Heat Exchanger Waterboxes,
December 12, 1979

Table 3-5

HYDROCARBON SAMPLE ANALYSIS SUMMARY

	Bottle iC ₄ 8/6/79	After Methanol Purge 8/15/79	After Propane 8/16/79	#3 8/21/79	#4 (1) 8/24/79	#5 (2) 8/30/79	#6 (3) 9/5/79	207 (5) 11/5/79	208 11/20/79	209 12/5/79	210 (6) 1/15/79	210A (7) 1/15/79	212 (6) (8) 1/16/79	213 (7) (8) 1/16/79
H ₂ O	-	.14	.022	.36	<.03	.010	<.01	.015	.007	.14	-	117 ⁽⁹⁾	.084	335 ⁽⁹⁾
CH ₄ OH	-	remainder	.061	.95	<.20	1.19	<.08	.06	-	.18	-	.	.12	
O ₂	.03	.66	.18	-	<.05	.09	<.05	1.6	-	.01	-		.004	
N ₂	.14		4.55	3.9	<.35	8.6	2.4	7.3	-	7.76	.01		.056	
CH ₄	.10		-	-	-	-	-	-	-	-	-		-	
C ₂ H ₆	-		2.5	-	-	-	-	-	-	-	-		.89	
C ₃ H ₈	-		89.0	2.1	-	.59	1.4	3.8	.44	.39	15.2		15.86	
nC ₄ H ₁₀	-		4.9	-	-	-	-	-	-	-	-		-	
iC ₄ H ₁₀	99.70		2.1	93.8 ±2	98.5 ±2	87 ±2	95 ±2 ⁽⁴⁾	92.8	98.3 ±2	90.5 ±2	82.6		77.8	
nC ₅ H ₁₂			-	-	-	-	-	-	-	-	-		-	
iC ₅ H ₁₂			-	-	-	-	-	-	-	-	1.8		3.1	

No Hexanes No Hexanes

< means less than

% is by volume

(1) taken after loop fluid flush of sample bomb

(2) taken after N₂ purge of sample bomb and unmixed loop contents(3) taken after low pressure N₂ purge of sample bomb with well mixed fluid loop contents

(4) adjustable to 98.4% by subtraction

(5) may contain some air contaminants

(6) vapor sample from bomb - not representative of liquid mixture

(7) sample taken from hydrocarbon high point vent, 250 psi, 300°F; analyzed by Karl Fisher method

(8) About 15 ml oil collected from sample 213, oil insoluble in Methanol and didn't evaporate at 55°C (131°F).

(9) Results expressed as (mg)_{H₂O} per Li. of sample bomb volume. This corresponds approximately to 0.2 and 0.6% by weight of water.

The non-condensable gas samples likewise held no surprises. H_2S was sought in these analyses but was not detected. The CO_2 was estimated as the remainder of the sample due to its large proportion in relation to the other gases. Earlier work performed by NARCO, a subsidiary of San Diego Gas and Electric Company, indicates that these non-condensibles constitute between 0.02 and 0.03 percent by weight of the well production. Although sample 106NCG was taken, it was lost due to a leaky valve on the sample bomb used to capture that sample.

Hydrocarbon samples were taken periodically to keep track of the composition of the fluid in the hydrocarbon loop. This was particularly useful during the initial drying, purging and loading of isobutane. The first sample was taken from a random bottle of isobutane as supplied by Phillips Petroleum Company, Borger, Texas. Subsequent samples were taken from a $1/2$ inch side connection to the six inch pipe between the suction of the hydrocarbon circulating pump and the hydrocarbon exit from the subcooler. However, samples 210a and 213 were taken from the high point vent in the hydrocarbon loop. The purpose of these two samples was to obtain another measure of the amount of water mixed within the hydrocarbon fluid. A summary of the results of GHT's hydrocarbon sample analyses is shown on Table 3-5. In general, the hydrocarbon sample bomb was purged with nitrogen and left with a nitrogen over pressure so that when connected to the hydrocarbon loop, any air could be flushed from the point of connection. This procedure primarily avoids using the loop hydrocarbon fluid as the purge fluid with its sudden release in the vicinity of the hydrocarbon pump and its motor. Additionally, since isobutane is the major component of the hydrocarbon fluid, its dilution by nitrogen enables its measurement to be more precise. Determination of the fluid composition is then done by subsequent adjustment of the composition to remove the nitrogen present.

The initial goal for the purity of isobutane in the hydrocarbon loop was 98% by volume. This level of purity was documented with sample 6 after several successive fill and vent operations with the bottled isobutane. Sample 207 and 208 were taken as noted on Table 3-5. Sample 209 was taken just prior to shutdown on 12/5/79. Post analysis emptying of sample 209 showed about 10-15 ml of fluid left in the sample bomb.

Periodic draining of the hydrocarbon system had always resulted in small amounts of non-hydrocarbon fluid being driven out ahead of the isobutane. In fact it became an operational procedure to periodically check the low point drains located both at the accumulator bottom and on the hydrocarbon circulation pump suction

pipng. The non-hydrocarbon fluid was presumed to result from incomplete drying of the hydrocarbon system prior to the final loading of isobutane. This fluid always contained a fine black suspension, which suspension also appeared as a froth on top of the liquid collected from those low points. Samples of this liquid and its froth were collected early in December, 1979 and submitted as sample 211 to GHT Laboratories for analysis. The results of this analysis are shown on Table 3-6.

This analysis was designed to aid in identification of the source(s) of this liquid. Accordingly, the analysis sought each of the components used in cleaning the hydrocarbon loop, the Mercaptan odorant that was added, and a finger print of the geothermal brine was used to check if some of the liquid might be due to leakage of this brine. In addition, the solid matter contained as a suspension and froth was also fired to check its carbon content. The high level of trace metals indicates corrosion and possible dissolution of the slag left in the equipment from fabrication. It does not appear that any geothermal brine leakage occurred. Minor amounts of each cleaning fluid additive and the mercaptan odorant were found as expected. The relatively large amount of carbon present may have been due to either grease left from fabrication or dissolution of graphite, a major component of the valve packing material. The non-hydrocarbon fluid thus appears to be primarily water left from the initial cleaning operations done prior to hydrocarbon filling.

Table 3-6

Hydrocarbon Sample Number 211

	<u>Liquid</u>	<u>Solid</u>
Iron (Fe)	9030 mg/li	63.0% Fe
Sulphate (SO_4)	(Total) 82 mg/li	---
Formaldehyde	Approx. 70 mg/li	---
Sulphites (SO_3)	not detected, less than 10 mg/li	---
Manganese (Mn)		0.4%
Copper (Cu)		0.6%
Chromium (Cr)		210 ppm
Lead (Pb)		910 ppm
Zinc (Zn)		915 ppm
Silicon (SiO_2)		2.7%
Acetylenic Alcohol	not detected, less than 5 ppm	---
Mercaptans	present, approx. 0.05%	---

Carbon	4.1%
Hydrogen	1.7%
Ash	88.0%

Samples 210 through 213 were taken of the two fluid mixtures used for test purposes. Since the fluids mixed are completely miscible and exhibit a fugacity of less than unity, subsequent correction to their vapor analysis is needed to indicate the true composition of the fluid sampled. The resultant mixture are shown on Table 3-7. These are the mixtures used for heat exchange performance analysis purposes.

During May, 1979 samples of cooling water were taken from Magma's deep pond and submitted to GHT Laboratories, Inc. for analysis. The cooling water comes from the East Highline Canal. Magma's only treatment to this water during the course of operations was the addition of triazole to a concentration in their main cooling pond of 15 ppm to control algae growth. This analysis of the cooling water is shown on Table 3-8. This water was used as make-up to the system cooling towers. The cooling water system was filled August 2, 1979, emptied, and subsequently refilled. After filling its pH was found to be 8.2. Sufficient 15 w/o HCl (supplied by Halliburton Company) was added to bring the cooling tower basin pH to 7.4 at 88°F. An additional 400 ml of 45 w/o zinc chromate was also added resulting in a pH of 7.2 at 90°F. Periodic pH measurements were made and small amounts of HCl added to keep the cooling water at a pH of between 7.0 and 7.4.

Shortly after test operations began on October 30th, a problem was encountered with a sticky control valve in the cooling water system. It stuck due to a scale build-up in its trim. This was due to two facts:

- a. Operations with low cooling water flows through the desuperheater together with its high surface temperatures, resulted in the flashing to steam of the cooling water; and
- b. The cooling water blowdown valve had been inadvertantly shut causing the total dissolved solids content of the cooling water to 15,410 mg/li (as determined by GHT November 7th).

Table 3-7
HYDROCARBON FLUID MIXTURES

	Nominal mole percent		
	<u>Propane</u>	<u>Isobutane</u>	<u>Isopentane</u>
Sample Number 210	15.2	82.6	1.8
Liquid corresponding to No. 210	4.3	85.8	9.9
Sample Number 212	15.9	77.8	3.1
Liquid Corresponding to No. 212	3.8	76.5	19.7

TABLE 3-8
COMPOSITION OF MAGMA POND WATER

HARDNESS (as CaCO_3)	385 mg/l
ALKALINITY (as CaCO_3)	126 mg/l
pH	7.95
TOTAL DISSOLVED SOLIDS	973
SULFATE (SO_4)	408 mg/l
BICARBONATES (HCO_3)	153 MG/l
SODIUM (Na)	0
POTASSIUM (K)	7.9 mg/l
CALCIUM (Ca)	90 mg/l
AMMONIA (NH_3)	0.7 mg/l
SPECIFIC CONDUCTANCE	1440 micromhos/cm
CHLORIDE (Cl)	156 mg/l

This scale on the trim of TCV 301 was analyzed by GHT and found to contain:

Carbonates	45.7% by weight
Calcium	32.0% by weight
Sulfate	6.8% by weight

This scale is estimated to be 29/32 calcium carbonate and 3/32 calcium sulphate.
To assure adequate blowdown:

- a. a TDS meter was borrowed from Magma to enable continual monitoring of the total dissolved solids level; and
- b. the blowdown flow, expressed in gallons per minute, was set numerically equal to the brine heat duty divided by a factor of one million; i.e.

$$\text{Blowdown (gpm)} = \frac{\text{Duty}}{10^6}$$

The TDS meter was obtained on November 16th and a sample of cooling water taken for both a meter reading and analysis by GHT. This calibration of the TDS meter showed that;

$$\text{Meter reading} \times 1,360 = \text{TDS}$$

Consequently instructions were left to keep the meter reading of cooling tower basin samples less than 2.5. This would assure a cooling water TDS of 3400 mg/Li or less. This maintains cooling tower operations at 3.5 cycles of concentration or less.

Severe fouling of both hot and cold cooling water surfaces was found on December 5, 1979. This fouling is shown in Figure 3-7. Samples of the fouling found showed:

Figure 3-7 Subcooler Waterbox, December 5, 1979

Iron	28.6% by weight
Calcium	16.3% by weight
Carbonate	28.9% by weight
Sulphate	2.5% by weight

The relatively large amount of iron present indicated corrosion was occurring. This corrosion can be seen as pitting on the cooling water heat exchanger water box covers, Figure 3-8.

Betz Laboratories (a water treatment specialist) was contacted regarding this problem with cooling water system fouling and corrosion. A Mr. Ed Allen of Betz visited the site on December 13, 1979, and took a sample of cooling system water for analysis. Betz recommendations developed on the 13th of December and from their subsequent analysis were to:

1. Feed one pint per shift of their chemical 419, and
2. Maintain cooling tower basin TDS at about 2000 ppm

These instructions were incorporated into operational procedures for all testing subsequent to December 5, 1979. The effect of this treatment can be seen in Figure 3-9, a picture taken of the condenser and subcooler cold end water boxes on January 17, 1980, after 65.3 additional test hours had accrued.

Figure 3-8 Pitting Evidenced by Condenser Waterbox Cover,
December 5, 1979

Figure 3-9 Condenser and Subcooler Waterboxes,
January 12, 1980

Section 4

TEST RESULTS

HEAT EXCHANGER PERFORMANCE, ISOBUTANE WORKING FLUID

Description of Operation

The purpose of the Binary Module Test was to determine the performance of geothermal brine to hydrocarbon heat exchangers in the field.

Desired information included the determination of the overall heat transfer coefficient for individual heat exchangers, the effect of time on the brine side fouling, log mean temperature differences determination, shell side pressure drop, and fluid property comparisons.

Data recorded included temperatures, pressures, differential pressures and flow rates. Table 4-1 lists those stations where data could be obtained for each fluid loop, brine, hydrocarbon and cooling water. The locations of these data stations is schematically shown in Figure 4-1. Generally, data was recorded once each hour and this data was processed within the following 24 hour period.

Tests for heat exchanger performance were performed from November 14, 1979 to December 5, 1979, and included runs number 1 through 281. Runs number 1 through 50 were taken during system start-up and were not processed due to inconsistent and incomplete data. Most of this start-up time was spent in calibrating and debugging the instrumentation and system's components. The main problem areas included thermometers being over or under range, differential pressure gauges, the turbine flow meter in the hydrocarbon loop, the venturi flow meter in the brine loop, and the digital pressure gauges. These problems were solved on-site and the facility equipment was either eliminated, replaced, or repaired. The flow meters, for instance, were not repaired and orifice plates with differential pressure gauges were used to measure the flows. Similarly industrial bourdon tube pressure gauges were used in place of digital pressure gauges.

Table 4-1
(as shown on Figure 4-1)

BRINE/HYDROCARBON DATA STATIONS

<u>Brine</u>	<u>Hydrocarbon</u>	<u>Cooling Water</u>
TW 101	TW 201	TW 301
102	202	302
103	203	303
104	204	304
105	205	305
106	206	306
107	207	308
	208	
	209	
	210a	
	210b	
	211	
	212	
PI 101	PI 201	PI 301
107	207	302
108a	204a	303
108b	204b	304
108c	204c	305
	204d	306
	208	307
	209	308
	211	
	212	
DPI 101	DPI 201	
102	202	
103	203	
104	204	
105	205	
106	206	
M 101	M 201 (or 201s)	M 301
		302
		304
		305
		307

Figure 4-1. Data stations.

Once the operational problems were solved and the system brought to a steady state condition, the processing of data began. Runs number 51 through 281 were processed during this test period. Out of these 231 runs, a total of 112 runs were rejected.

Before processing, all the data was checked and re-checked, both manually as well as by computer. Manual checking of data for missing or inconsistent data was done by the PFR Engineer on duty. Computer programs checked the data in the following manner:

- Each fluid stream is followed in its flow wise direction and the data points (temperatures and pressures) were checked for a consistent increase or decrease in their values.
- Temperatures for brine and hydrocarbon taken at the same location were checked. If the temperature of hydrocarbon was higher than the temperature of brine for the same location an error message would be printed.
- A summation of the differential pressures measurements was compared with the overall prime drops for both the shell and tube sides of the heat exchangers. If a difference was found, again an error message would be printed.
- On the cooling water side the overall flow rate was compared with a summation of the individual flow rates through each of the three heat exchangers.

Data which was not subject to an error message from these checks was deemed acceptable. This accepted data was reduced by means of computer programs and before the results were printed a final check was made for any unacceptable behavior of one or more heat exchangers. A good example would be the primary heat exchanger B6. If the quality of hydrocarbon stream leaving from B6 was less than 1.0, an error message would be printed. Likewise, if the vapour quality of hydrocarbon leaving the condenser was greater than 0, an error message would be printed.

A final qualification of the data was made with consideration of the results of heat and mass balance determinations.

The principal reasons for rejection of a set of data were unreasonable flow rates and/or inconsistent temperature readings. The results reported here were obtained from those sets of data which were qualified under the above restrictions.

The Data Processing System is described in detail in Section 5. The basic operations performed included the determination of heat duties for each fluid stream for each exchanger.

A heat balance between the brine and hydrocarbon duties was calculated and reported as percentage difference. Also overall heat duty calculations are performed with the temperature and pressure conditions of the brine and hydrocarbon streams at the inlet and exit of the primary heat exchanger field. From the overall heat duties, an overall heat balance was made and is reported as percentage difference. The log mean temperature differences (LMTD) were calculated for individual heat exchangers.

From the calculated values of LMTD and brine heat duty, the exchanger by exchanger heat transfer coefficients were calculated.

Use of the LMTD function assumes a linear temperature-duty relationship. This requires a constant specific heat. These assumptions are most valid for the initial and final heat exchangers and contributed to the difficulty in establishing the performance of the intermediate heat exchangers.

Heat and Mass Balance

The overall system was divided into two parts. In the first part, heat was delivered from the hot geothermal brine to the cold hydrocarbon fluid in six shell and tube heat exchangers placed in series. The second part enabled the heat in the hydrocarbon fluid to be delivered to the cooling water in the three heat exchangers.

Heat and mass balances were performed for each exchanger and for each part. Since a closed loop can be assumed for both the brine and hydrocarbon fluids, the flow rate of each stream is the same through each exchanger. In determining heat exchanger performance, the heat balances for the brine to hydrocarbon heat-exchanger are of importance. These balances were done using the inlet and exit enthalpies of the brine and hydrocarbon streams and their flow rates. The results of overall heat balances for individual runs are reported in Figure 4-2.

Figure 4-2. Overall heat balance results.

According to the test plan, an error of $\pm 3\%$ in the overall heat balance was acceptable. However, considering the operational problems involving the data acquisition system instrumentation, this error limit was further increased to $\pm 7\%$, which limit is still less than the $\pm 10\%$ limit used for pilot plants of similar sizes. From Figure 4-2 it appears the data taken before run number 200 falls within the allowable accuracy of $\pm 7\%$, however, from run number 200 onwards this error increased to more than $\pm 20\%$. This high percentage error increases the uncertainty beyond usual industrial limits of accuracy.

One source of this uncertainty was the hydrocarbon flow meter. As mentioned earlier the turbine flow meter ceased functioning at this time. The flow was subsequently measured by an orifice plate and the accuracy of the orifice plate is questionable due to the changes in the temperature of hydrocarbon fluid.

Another source of this uncertainty was the propagation of a tube to tube sheet leak in exchanger B1. This resulted in a small but continuous loss of hydrocarbon to the brine. Ultimately, the tests were shut-down due to excessive loss of hydrocarbon from the system. It should be noted that this loss of hydrocarbon fluid also resulted in an increasing hydrocarbon temperature at the flow orifice noted above.

Another source of uncertainty is the possibility of brine or water in the hydrocarbon loop. While water could have been left in the hydrocarbon loop as a result of the initial cleaning operation, laboratory reports on samples of hydrocarbon taken during the test period showed negligible traces of water.

During system operation, the direction of leak must be from the hydrocarbon to the brine side due to the higher pressure on the hydrocarbon side. The only time that brine could leak into the hydrocarbon is when the system is shut-down and during start-up. From the size of the leak (pin-hole) and the small pressure differential favoring brine leakage from tube to shell into the hydrocarbon during these times one concludes that negligible additions of brine occurred. This supported by the laboratory tests previously noted. Water will evaporate when its partial pressures is equal to the vapor pressure at temperature and the latent heat of evaporation will cause a decrease in the heat received by hydrocarbon. This will result in a smaller hydrocarbon temperature rise in that heat exchanger where the evaporation takes place. No such effect was observed.

It is important to note that a very large percentage difference between the heat duties for exchangers B2 through B5 existed uniformly when the overall heat balance is within acceptable limits. From Figure 4-3 it is observed that due to the very nature of the hydrocarbon fluid, its temperature enthalpy plot is not a straight line but has curvature. Exchangers B2 through B5 are in this curved region thereby experiencing very small temperature differences between the brine and hydrocarbon streams. Due to this small temperature difference, a small error in the recording of these temperatures can cause a large error in the results. The results obtained from these heat exchangers are not reliable and therefore are not used in determining the performance of these exchangers.

Performance Analysis

Performance of the overall system was determined by analysing the performance of each heat exchanger. This performance was determined by comparing the measured overall heat transfer coefficients with the values obtained from operation of a computer program using advanced heat transfer methods. The measured values of the overall pressure drops were also compared with the values predicted by this computer program.

The PFR propriety computer program, SAT-1, uses a stream analysis procedure for the shell side analysis as proposed by Tinker⁽¹⁾ and further discussed by other researchers. The tube-side correlation used predicted the brine side heat transfer coefficient as discussed in Section 5 and based on work by Lawrence and Sherwood. This program was supplied the physical property data including the density, thermal conductivity, specific heat, and viscosity, and the process variables-temperatures in and out, flow rates and pressure levels. From this data, the correlations predicted pressure drops and heat transfer coefficients for both the shell and tube sides, the overall coefficients, and heat duties.

The validity of the physical and thermodynamic properties of test fluids were first verified. The fluids used were brine (heat source) and isobutane (test fluid). Due to the low salinity of brine, the properties of water were used for these tests and did not require further verification. The properties of isobutane, especially in the super-critical region required validation of the enthalpy values of isobutane used in the process calculations.

(1) Tinker, T., Proceedings of the General Discussions on Heat Transfer, Inst. of Mech. Engrs., London and AM Soc of Mech. Engrs., 1951 and 1958.

Figure 4-3. Temperature versus enthalpy isobutane (Run 110, 11/21/79).

The results indicate that the values recorded in the property data bank are within +7% or better of those obtained from back calculation. A wide scatter in the percentage difference after run no. 200 for both exchangers corresponds to the time period where a tube to tube sheet leak was detected in exchanger B1 thereby producing unreliable results.

Evaluation of Hydrocarbon Properties. The physical and thermodynamic properties of hydrocarbons used for calculations are derived from the work of K. Starling⁽¹⁾. The values of enthalpy, entropy, specific volume and specific heat (Cp) were recorded as a function of temperature and pressure. The main property of interest for the working fluid is its enthalpy. The enthalpy of isobutane recorded at various temperatures and pressures was checked by comparing it with the values reported by C. F. Braun⁽²⁾ from test results. The recorded values of enthalpy were checked against the values obtained by back calculation from the brine heat duty.

Both pressures and temperatures are required to determine the enthalpies, but since the pressure gauges between the heat exchangers were inoperable only the enthalpies for B1 and B6 could be accurately determined. These enthalpies are significant as the hydrocarbon is in liquid phase at the exit of B1 and is superheated at the exit of B6. Table 4-2 and Figures 4-4 and 4-5 summarize the results of this comparison. Figure 4-4 corresponds to exchanger B1 and 4-5 to B6.

Overall Heat Transfer Coefficient. The overall heat transfer coefficient has been plotted as a function of run number (time) for both heat exchangers B1 and B6. This plot is shown as Figure 4-6. It should be noted that the heat transfer coefficient shows a decreasing trend from run number 129 through run number 146 and again from run number 200 through 268. In both cases, this decrease is due to the decreased tubeside and shell side flows in the heat exchanger units according to the Test Plan as the overall heat exchanger duty was decreased to about 80% and 60% for the run series shown. Reference to Figure 3-2 shows the relationship of decreased flow rate and heat transfer coefficient.

- (1) K. Starling, Fluid Thermodynamic Properties of Light Petrochemicals, Gulf Publishing, 1973.
- (2) Experimental Enthalpies for a Mixture of 80 MOL Percent Isobutane in Isopentane; EPRI Report ER1034.

Table 4-2

EXIT ENTHALPY, BTU/LB

Run No.	B1			B6		
	Back Calculated From Brine Duty	From Data Bank	% Diff.	Back Calculated From Brine Duty	From Data Bank	% Diff.
51	51.5	51.1	0.78	191.2	179.5	6.52
52	53.6	53.7	-0.18	196.9	190.4	3.41
53	52.6	51.3	2.5	200.4	181.6	10.35
56	51.0	52.0	-1.9	193.2	188.2	2.65
57	51.8	52.2	-0.76	192.7	189.9	1.5
58	51.2	52.0	-1.54	192.7	188.8	2.1
59	52.9	52.6	0.57	193.3	190.5	1.5
60	54.2	52.9	5.1	200.5	189.0	6.1
61	54.9	53.7	2.23	202.4	190.8	6.1
62	54.6	53.4	2.24	200.5	190.9	5.03
63	55.1	53.9	2.22	199.1	190.8	4.35
83	60.6	58.1	4.3	198.5	198.5	0.0
84	59.4	58.1	2.24	198.4	199.3	-0.45
85	59.6	57.7	3.3	201.3	203.0	-0.84
86	60.7	58.7	3.4	201.1	200.5	0.3
87	60.9	57.5	5.9	199.9	200.8	-0.45
88	60.8	57.7	5.4	200.2	201.2	-0.6
92	59.1	56.8	4.05	187.4	197.8	-5.3
94	56.8	55.4	2.53	189.5	219.5	-13.66
96	58.2	54.9	6.01	196.1	194.1	1.03
97	55.9	55.3	1.08	195.2	194.0	0.62
98	56.5	53.7	5.21	198.7	196.8	0.96
99	56.2	55.8	0.72	197.9	195.8	1.1
110	58.5	56.9	2.8	195.2	190.3	2.6
112	58.6	57.1	2.6	192.2	191.4	0.42
113	57.8	57.0	1.4	194.6	188.6	3.23
114	61.9	57.8	5.4	194.4	189.6	2.53
115	62.5	56.9	9.8	195.2	190.9	2.25
116	61.8	58.5	5.6	195.1	190.6	2.4

Table 4-2 (continued)

Run No.	B1			B6		
	Back Calculated From Brine Duty	From Data Bank	% Diff.	Back Calculated From Brine Duty	From Data Bank	% Diff.
129	54.3	56.4	-3.7	180.4	183.7	-1.8
130	55.4	56.7	-2.3	183.2	185.8	-1.4
131	54.5	57.0	-4.4	182.7	185.4	-1.4
132	55.0	55.7	-1.3	181.8	184.3	-1.4
133	55.3	57.0	-3.0	182.2	184.7	-1.3
134	55.2	56.2	-1.8	183.3	184.1	-0.43
135	55.3	57.2	-3.3	182.0	185.4	-1.8
136	52.9	57.7	-6.6	176.1	188.5	-6.6
184	60.4	58.7	2.9	194.3	188.6	3.02
185	61.6	60.4	2.0	200.3	193.0	3.8
186	62.4	60.3	3.5	201.0	193.2	4.04
187	62.8	60.9	3.1	202.2	193.6	4.4
188	62.3	60.7	2.6	201.7	193.3	4.35
189	62.5	60.6	3.1	201.7	193.3	4.35
190	61.5	61.0	0.8	194.6	191.6	1.57
191	61.5	61.0	0.8	194.5	191.8	1.4
192	63.4	62.4	1.6	207.9	195.7	6.23
193	65.1	63.9	1.88	213.4	196.2	8.7
194	65.3	63.9	2.2	214.0	198.8	7.65
195	65.5	63.6	3.0	214.2	198.8	7.75
196	65.8	63.4	3.3	214.1	198.8	7.7
197	65.9	63.9	3.1	214.6	198.8	7.95
198	66.4	63.3	4.9	214.6	197.9	8.4
200	61.7	69.7	-11.5	195.2	204.1	-4.4
201	63.8	71.3	-10.5	198.3	203.2	-2.4
202	63.8	71.3	-10.5	198.3	212.1	-6.5
203	64.7	69.9	-7.4	198.8	202.3	-1.73
204	63.8	71.3	-9.1	205.8	212.5	-3.15
205	64.5	71.5	-9.8	199.3	209.9	-5.05
206	64.0	71.3	-10.2	197.8	213.3	-7.3
216	74.7	66.5	12.33	181.8	184.4	-1.4

Table 4-2 (continued)

Run No.	B1			B6		
	Back Calculated From Brine Duty	From Data Bank	% Diff.	Back Calculated From Brine Duty	From Data Bank	% Diff.
217	76.0	67.2	13.1	184.5	186.1	-0.86
218	76.3	67.7	12.7	189.0	185.4	1.94
219	81.7	67.7	20.68	214.5	188.8	12.61
220	81.9	67.8	20.7	226.7	189.1	19.88
221	98.8	68.4	44.44	179.1	186.4	-3.19
222	98.5	68.6	43.6	178.9	186.7	-4.2
223	98.9	68.4	44.6	179.5	186.2	-3.6
224	68.7	69.9	-1.7	216.9	190.3	13.98
225	76.1	69.2	9.9	211.5	190.8	10.85
226	75.5	69.2	9.1	207.6	191.0	8.7
227	80.2	70.2	14.25	220.6	190.2	15.72
228	80.6	70.6	14.16	207.5	190.8	8.75
229	73.5	68.4	7.5	173.8	190.6	-8.81
236	80.2	73.0	9.86	177.4	189.4	-6.34
237	78.1	73.0	7.0	174.8	191.0	-8.5
238	77.0	72.7	6.0	174.3	190.5	-8.5
239	80.3	73.3	9.5	171.8	192.1	-10.6
240	80.8	73.6	9.8	173.1	193.7	-10.64
241	80.9	74.4	8.74	173.3	193.7	-10.53
242	81.4	74.2	9.7	171.8	192.5	-10.75
243	81.7	74.2	10.11	171.8	192.5	-10.75
247	78.7	72.6	5.6	159.7	202.4	-21.1
248	76.6	73.4	4.4	153.3	201.0	-23.7
249	77.4	72.6	6.6	158.8	201.4	-21.1
250	79.7	72.6	9.8	161.3	201.7	-20.3
251	77.6	73.0	6.3	157.0	202.7	-22.55
252	78.3	73.8	6.1	156.0	203.0	-33.15
253	85.6	76.4	12.04	222.7	208.7	6.71
254	77.4	76.4	1.31	208.6	208.7	-0.05
255	79.3	77.5	2.32	212.6	210.1	1.2

Table 4-2 (continued)

Run No.	B1			B6		
	Back Calculated From Brine Duty	From Data Bank	% Diff.	Back Calculated From Brine Duty	From Data Bank	% Diff.
256	80.3	77.7	3.35	212.4	210.5	0.9
258	79.7	78.2	1.9	211.5	210.6	0.43
259	81.6	78.3	4.2	211.7	210.6	0.52
260	79.8	78.2	2.05	212.1	210.6	0.71
262	78.1	71.6	9.1	189.7	191.7	-1.0
263	76.4	71.3	7.15	185.4	192.2	-3.43
264	75.9	70.9	7.05	168.4	204.3	-17.57
265	76.7	71.6	7.12	157.6	206.6	-23.72
266	75.9	71.6	6.01	164.0	206.9	-20.73
267	78.5	71.8	9.3	159.3	207.3	-23.15
268	75.9	71.6	6.01	165.1	207.4	-20.4

Table 4-3

COMPARISON OF HEAT TRANSFER COEFFICIENTS MEASURED VS. CALCULATED

Measured values are those that were obtained from the results of data reduction.

Calculated values are the values predicted by PFR SAT-1 computer program for the process conditions.

<u>Run #</u>	<u>Exch.</u>	<u>Measured Values</u> <u>Btu/Hr Ft² °F</u>	<u>Calculated Values</u> <u>Btu/Hr Ft² °F</u>	<u>% Diff.</u>
56	B1	309.13	345.1	-10.42
	B6	550.74	527.25	4.46
57	B1	312.59	344.6	-9.3
	B6	566.58	526.7	7.6
83	B1	357.76	346.3	3.31
	B6	514.83	530.2	-2.9
84	B1	350.87	345.8	1.47
	B6	486.81	529.5	-6.2
85	B1	345.24	346.0	-.22
	B6	527.61	529.8	-.41
53	B1	319.91	349.0	-8.34
	B6	578.62	535.05	8.14

Figure 4-4. Enthalpy difference, exchanger B1.

Figure 4-5. Enthalpy difference, exchanger B6.

The overall heat transfer coefficients for exchangers B1 and B6 were compared with the values obtained from the PFR SAT-1 computer program. Table 4-3 shows the results of this comparison.

The percentage difference between the measured and predicted values indicate that these units provided consistent results and that no changes are required to the generalized correlations used in the computer program.

The overall measured heat transfer coefficients of the heaters have been determined to an accuracy which varies depending on the factors arising from data uncertainty considerations. The smaller the temperature driving force between the hydrocarbon working fluids and the geothermal brine, the less accurate the results will be. Run number 110 illustrates the potential combined error inherent in the data. Table 4-4 summarizes the measured and calculated results of this run.

Table 4-4
SUMMARY OF RUN NO. 110

Heat Exchanger Unit No.	Q Brine Duty M-Btu/hr	% Difference in Duty from Hydrocarbon	LMTD (°F)	Measured Overall U (Btu/hr ft ² °F)	% Uncertainty Error		
					Q	LMTD	U
B1	3.37	-4.2	33.53	348.5	5.8	4.4	10.6
B2	2.20	-9.2	19.0	401.0	9.0	7.7	17.9
B3	1.55	-17.8	12.72	422.4	12.8	11.2	25.9
B4	1.24	-25.9	10.59	404.1	16.1	13.4	31.2
B5	2.14	-32.4	15.51	478.2	9.3	9.9	21.9
B6	5.05	-8.5	33.12	528.0	4.0	4.6	10.0

From the above table it is clear that the data from units B1 and B6 will be the most accurate. All the heat exchanger units have the same tube and shell side configurations except for B6. B6 has the same number of tubes but has double segmental baffles while the others have single segmental baffles. Since these units were in series, the mass flow rate was identical in each unit, but the velocity of the hydrocarbon increased unit to unit as it was heated. The main difference in the performance of these units is that as the hydrocarbon is heated, the temperature versus enthalpy profile bows (see Figure 4-3) nearly meeting the

nearly straight profile of the brine. The minimum temperature "pinch" for this run was 10.2°F occurring between units B2 and B3. There were other runs with pinches of less than five degrees Fahrenheit. This low pinch point temperature difference makes the accuracy of the data for B2 to B5 quite low. This type of problem is inherent in the working fluid and the precision of the instrumentation and is not due to the manner in which the operating data was taken,

By reviewing Table 4-4, it can be seen that the overall heat transfer coefficient varied from 348.5 to 528.0 Btu/hr ft² °F. These values are consistent with values which were obtained with the PRF shell and tube program, SAT-1, which is summarized in Table 4-5.

Table 4-5
COMPARISON OF HEAT TRANSFER COEFFICIENTS - RUN 110

<u>Unit Number</u>	<u>U-Measured</u>	<u>U-Calculated</u>	<u>% Difference</u>
B1	348.5	390.0	+11.9
B2	401.0	397.0	-1.0
B3	422.4	449.1	6.3
B4	404.1	406.7	0.6
B5	528.0	460.0	-12.9
B6	528.0	460.0	-12.9

The low percentage difference from units B1 to B6 indicates that if competent organizations specializing in the design of heat exchangers are given the basic process, thermodynamic, and transport data, the heaters and condensers could be designed to better than 15% accuracy for East Mesa and other geothermal sites with brines which have similar, minimal fouling characteristics.

Overall Pressure Drop. A comparison between the pressure drop measured across the six brine to hydrocarbon heat exchangers and those pressure drops calculated by the computer program. SAT-1 was made and is summarized in Table 4-6. A difference of up to 20% between these two sets of values is found from this

Table 4-6
PRESSURE DROP COMPARISON

<u>Run #</u>	<u>Measured</u>	<u>Calculated</u>	<u>% Diff.</u>
53	95.0	86.5	9.83
56	90.0	81.7	10.16
57	90.0	81.2	10.84
83	85.0	76.57	11.01
84	90.0	80.52	11.77
85	93.0	83.5	11.4
110	109.0	89.6	+21.65
129	110.0	136.77	-19.57
132	105.0	132.0	-20.45
184	100.0	86.34	15.82
188	80.0	68.7	16.45
DESIGN	50.0	84.4	-40.76

comparison. Therefore, heat exchanger design for similar geothermal brine to hydrocarbon service should include allowance for pressure drops 20% greater than expected.

Statistical Data Analysis

The value of an experimental result depends on the consistency and reliability of the data analysed. To determine this consistency and reliability, data taken over the test period was divided into several groups corresponding to similar process conditions. Root Mean Square methods were used to determine the deviation of these data from their mean value. The main data points that were tested were the hydrocarbon flow rate and the hydrocarbon temperature and pressure at the exit of exchanger B6. Table 4-7 lists the average percentage deviation of these data from their mean values for the period during the heat exchanger performance tests.

From the results of statistical data analysis it is observed that the error in the hydrocarbon flow rate for the first three groups is unacceptably high compared to other groups. This difference in the flow rates affects the overall results, principally the heat transfer coefficients. From Figure 4-6 it is seen that a maximum scatter of approximately 17% is observed in the values of overall heat

transfer coefficient for exchanger B1 and a 10% scatter for exchanger B6 for runs number 51 through 63. From run number 110 onwards, the deviation in the results is very small. This corresponds to the results of the statistical analysis done and as shown on Table 4-7.

The results of statistical data analysis for run number 200 onward show that the data reported during this time period was very consistent. However, as previously noted, a large percentage difference existed between the heat duties, and a decreasing overall heat transfer coefficient was observed for these runs.

This indicates that even though the measurements were consistent, they were consistently wrong. The heat exchanger performance as indicated by these runs is thus suspect due to instrumentation and instrument system inaccuracies.

HEAT EXCHANGER PERFORMANCE, MIXED HYDROCARBON WORKING FLUIDS

The purpose of these tests was to determine the performance of the six brine to hydrocarbon heat exchangers using mixtures of hydrocarbon fluids. These tests were performed January 14 through 16, 1980. Three different hydrocarbon mixtures were used. Table 4-8 summarizes the compositions and the critical temperatures and pressures of these mixtures. It also includes the properties of an ideal 80-20 mixture of isobutane and isopentane.

Table 4-7
HYDROCARBON MIXTURES, PERCENT MOLE FRACTION

<u>Mixture No.</u>	<u>Propane</u>	<u>Isobutane</u>	<u>Isopentane</u>	<u>T_c °F</u>	<u>P_c PSIA</u>
1	4.8	95.2	0.0	272.6	538.5
2	4.3	85.8	9.9	276.4	519.3
3	3.8	76.5	19.7	280.2	500.34
80-20	0.0	80.0	20.0	282.75	496.50

Figure 4-6. Heat transfer coefficient versus run number (time).

Table 4-8

STATISTICAL DATA ANALYSIS

(percentage deviation)

Run #	Brine Flow	Brine Temp IN	I-C4 Flow	I-C4 Temp OUT	I-C4 Press. OUT
51-63	2.141	0.358	30.881	1.099	0.916
83-88	0.000	0.396	34.899	0.395	0.864
92-99	0.000	1.255	33.572	3.827	0.807
110-116	0.408	0.324	0.287	0.535	1.030
129-136	0.564	0.299	2.509	0.648	0.516
184-198	0.675	1.591	0.000	1.005	0.905
200-206	0.000	0.622	1.214	1.695	0.000
216-223	0.000	0.048	0.000	0.316	0.756
224-230	0.000	1.118	5.121	0.103	0.337
236-243	0.000	0.106	0.000	0.196	0.948
247-252	0.000	0.245	1.026	0.215	0.622
253-259	0.000	0.308	0.000	2.268	0.338
262-268	0.000	0.412	0.507	2.428	0.480

A comparison of the critical temperature and pressure of mixture No. 3 with the 80-20 mixture shows the effects of propane to be negligible. Therefore the data scans taken with mixture No. 3 were processed using the physical properties of an 80-20 mixture. The other mixtures were not computer processed at this time due to the time and expense required to obtain and implement computer listings of their physical properties. A total of five data scans were taken with mixture number 3 and two of these data scans were not computer processed due to errors in the data. Corrections to the other data scans were made and the results obtained from hand calculators are included here.

Heat and Mass Balance

The overall heat and mass balances for the mixture data that was processed are within an acceptable accuracy of $\pm 7\%$; however, due to the lack of a sufficient amount of data no further analysis on these balances can be done.

Performance Analysis

From the plots of temperature vs. enthalpy of the hydrocarbon mixtures, Figures 4-7 through 4-11, the data appears to be in the supercritical region. It was found that the temperature points on the brine side are not all in a straight line. This implies that some of the intermediate brine temperature measurements for the six heat exchangers appear to be unreliable. The results for exchanger B1, where both the inlet and outlet temperatures of brine fall on the overall straight temperature vs. enthalpy line for brine, appear to be the only results that can be analyzed. An overall heat transfer coefficient for exchanger B1 is calculated to be approximately $280 \text{ Btu/hr-ft}^2\text{-}^\circ\text{F}$. The calculated heat transfer coefficients for B1 are reported on Table 4-9.

This value of the heat transfer coefficient for heat exchanger B1 is 20% less than what was obtained with isobutane for similar shell side velocities. This performance decrease is anticipated for these hydrocarbon fluid mixtures.

Heat transfer coefficients for exchanger B6 were calculated for these runs after making a temperature correction for brine at the exit of exchanger B6 (shown by a X (cross) on Figures 4-7 through 4-11. These calculations yielded heat transfer coefficients of approximately $220 \text{ Btu/hr-ft}^2\text{-}^\circ\text{F}$. Additional data is required to enable accurate determination of these heat exchange coefficients

Table 4-9
HYDROCARBON MIXTURE

<u>Run #</u>	<u>B1 Heat Exchanger U, Btu/hr-ft²-°F</u>
AA1	295.53
AA2	303.67
BB1	276.37
BB2	277.64
DD1	263.5

Avg = 283.4 Btu/hr-ft²-°F

Figure 4-7. Temperature versus enthalpy (Run No. AA1).

Figure 4-8. Temperature versus enthalpy (Run No. AA2).

Figure 4-9. Temperature versus enthalpy (Run No. BB1).

Figure 4-10. Temperature versus enthalpy (Run No. BB2).

Figure 4-11. Temperature versus enthalpy (Run No. DD1).

CONDENSING COEFFICIENT EVALUATION

Operations

While the test plan originally included condensing coefficient evaluation tests, basic design and operations with the condenser inhibited determination of condensing coefficients. However, a value for the overall heat transfer condensing coefficient can be calculated by hand principally based on data taken during the time period including runs number 51 through 63.

Since this test period was directed toward evaluation of the performance of the brine to hydrocarbon heat exchangers no particular attention was paid to the cooling water side. Operations sought to maintain complete condensation of the hydrocarbon while keeping the condensing pressure sufficiently high in order to maintain the hydrocarbon circulation pump suction side pressure within test specification limits. Heat balances between the hydrocarbon and cooling water streams varied between 2% and 20% for these runs.

Performance Analysis

Overall heat transfer coefficient determinations for the condenser were not performed by computer. The mean temperature driving force in the condenser can be quite complex since the hydrocarbon may simultaneously undergo desuperheating, condensing and cooling, as shown on Figures 4-12 and 4-13. There are two basic methods used by industry to determine condenser design or performance when these three heat transfer regimes may be present. One method uses a weighted mean temperature driving force. The second method uses a stepwise and iterative procedure. Both methods are further complicated by the possibility of wet wall condensation.

The desuperheating portion of the process may undergo "wet-wall" condensation in which the wall temperature is lower than the dew point of the hydrocarbon and therefore the vapor condenses even though the bulk stream gas temperature is above the dew point. When this occurs, the temperature driving force for the condensation process is the difference between the saturation temperature of the hydrocarbon and the local water temperature.

Further complications are also introduced by cooling water side fouling. the condenser was purposely fouled during operations because its design was too oversized to allow the desired operating conditions. This fouling was of a magnitude where in some cases it accounted for more than 50% of the resistance to

heat transfer. Computer data reduction techniques are quite complex because it is necessary to determine heat transfer wall temperatures at all the important points in the condenser. To avoid complicating the computer analysis with an iterative procedure, the condenser data was reduced by hand.

Figure 4-12.
Desuperheating -
condensing -
subcooling.

Figure 4-13.
Wet-wall
condensing -
subcooling.

An overall condensing heat transfer coefficient for isobutane was calculated by hand using data taken during runs 51 through 63. To do this the saturation temperature of isobutane for the measured condenser pressure was used to determine the log mean temperature difference. The value of the overall heat transfer coefficient was estimated to be in the neighborhood of $250 \text{ Btu/hr-ft}^2\text{-}^\circ\text{F}$ for the average conditions of 195 psia and 175°F .

In the case of the hydrocarbon mixtures approximating the $80/20$ mixture, similar techniques were used and an overall heat transfer coefficient of $180 \text{ Btu/hr-ft}^2\text{-}^\circ\text{F}$ is obtained.

A more detailed analysis of these results is not possible due to the lack of sufficient and reliable condenser data.

Section 5

DATA PROCESSING SYSTEM

DATA PROCESSING SYSTEM DESCRIPTION

The data processing system consisted of the following units:

- a. minicomputer
- b. data sheets, and
- c. computer programs.

The mini computer is a PDP 11/34 computer supplied by Digital Equipment Corporation. This PDP 11/34 includes a central processing unit, two disk drives, a manual entry I/O terminal, and a RT-II Fortran operating system. The design of this system is shown in Figure 5-1 Figure 5-2 schematically shows the computer programs used with the Data Processing System.

The Data Processing System was divided into the following steps:

- o Manual checking of data sheets
- o Inputting the data in the computer
- o Printout of data
- o Storage of data
- o Data processing
- o Printout of final results.

Manual Checking of the Data Sheets

Before the data was entered, it was checked for any inconsistency or human error by the PFR resident engineer. Any inconsistencies were reported to the Plant Manager for immediate correction or his direction. These inconsistencies included missing or inconsistent values of temperature, pressure, or flow rate readings.

Inputting of Data into the Computer

After the data was checked and found satisfactory by the PFR resident engineer, it was logged into the computer using program INDATA. Program INDATA was designed to receive the data directly from the data sheets as described in Section 7. No changes to or conversions of the data were required.

Figure 5-1. Data processing system.

Figure 5-2. Computer program schematic.

Checking of Data Using Computer Programs

Once a complete set of data was logged into the computer, subroutine CHECK, checked for any missing data, inconsistencies, or human error that might occur during the input of data. This subroutine followed the process streams in their flowwise directions and checked for gradual and consistent increase or decrease of the temperatures and pressures at each data station. For example, if the temperature of hydrocarbon at a station was reported higher than the brine temperature at the same station, an error message would be printed out. If the temperature of hydrocarbon in an exchanger was lower than its temperature in the preceding exchanger, an error message would be printed out. These and other checks were included to assure the credibility of the data before further processing was done.

Printout of Data

After a complete check was made, the data was stored on the disks for further processing. A copy of the data (up to three copies, obtained by using multiple copy carbon papers) was printed out. The printout is shown as Figure 5-3. The format of the printed data was kept identical to that of data logging sheets to enable rapid visual verification of the data. In the case of an error in the original data, the printout included an error message, coded to that particular data value.

Storage of Data

The system was designed to protect the data against unforeseen circumstances. The data was stored on two separate disks and a hard copy obtained from the printer for permanent record. Together with the original data sheet, three different sources of data existed in the event of the loss of one source.

Data Reduction

After the data had been stored on the disk, it was available for immediate or future processing. Data reduction was achieved through the main program called PSEMT. This program controls other subroutines as described in Figure 5-2. The main data values required for processing are the temperature, pressure, and flow measurements for each process stream together with the physical parameters of the equipment involved. The hydrocarbon physical properties were based on "Fluid Thermodynamic Properties for Light Petroleum Systems," Gulf Publishing Company,

TIME START : 05:00
TIME COMPL : 05:20
OPERATOR : STAN

PSEMT DATA LOGGING SHEET
NO HOURS FROM START OF SERIE : 11
TEST FLUID : I-C4

RUN NO: 134/SERIE: 2
RUN DATE : 11-24-79
CURRENT DATE : 24-NOV-79

	THERMOMETER THERMOWELL			TEMPERATURE INDICATOR			PRESSURE INDICATOR			DIFFERENTIAL PRESSURE INDICATOR		
	HIGH	LOW	%	HIGH	LOW	%	HIGH	LOW	%	HIGH	LOW	%
HEATERS												
TW202	191.0	0.0	0.00	0.0	0.0	0.00	0.0	0.0	0.00	DPI101	6.7	0.0 0.00
TW102	214.6	0.0	0.00	0.0	0.0	0.00	0.0	0.0	0.00	DPI201	11.0	0.0 0.00
TW204	241.0	0.0	0.00	0.0	0.0	0.00	PI204A	606.5	0.0 0.00	DPI102	6.7	0.0 0.00
TW104	250.0	0.0	0.00	0.0	0.0	0.00	PI204B	568.2	0.0 0.00	DPI202	15.4	0.0 0.00
TW206	263.2	0.0	0.00	0.0	0.0	0.00	PI204C	494.8	0.0 0.00	DPI103	6.7	0.0 0.00
TW106	287.0	0.0	0.00	0.0	0.0	0.00	PI204D	0.0	0.0 0.00	DPI203	15.4	0.0 0.00
TW207	282.8	0.0	0.00	TI207	290.0	0.0 0.00	PI207	500.0	0.0 0.00	DPI104	6.7	0.0 0.00
TW107	342.0	0.0	0.00	TI107	345.0	0.0 0.00	PI107	260.0	0.0 0.00	DPI204	17.6	0.0 0.00
TW205	253.0	0.0	0.00	0.0	0.0	0.00	PI108A	0.0	0.0 0.00	DPI105	6.7	0.0 0.00
TW105	265.0	0.0	0.00	0.0	0.0	0.00	PI108B	0.0	0.0 0.00	DPI205	22.0	0.0 0.00
TW203	221.0	0.0	0.00	0.0	0.0	0.00	PI108C	0.0	0.0 0.00	DPI106	6.7	0.0 0.00
TW103	237.0	0.0	0.00	0.0	0.0	0.00	0.0	0.0	0.00	DPI206	28.6	0.0 0.00
TW201	143.0	0.0	0.00	TI201	140.0	0.0 0.00	PI201	610.0	0.0 0.00	0.0	0.0	0.00
TW101	183.4	0.0	0.00	TI101	175.0	0.0 0.00	PI101	220.0	0.0 0.00	0.0	0.0	0.00
DESUPERHEATER, CONDENSER AND SUBCOOLER												
ALL FLOW RATES BELOW												
TW212	140.0	0.0	0.00	TI212	141.0	0.0 0.00	PI212	200.0	0.0 0.00	M 201	385.0	0.0 0.00
TW302	90.2	0.0	0.00	TI302	90.0	0.0 0.00	PI302	27.0	0.0 0.00	M 101	202.0	0.0 0.00
TW305	90.8	0.0	0.00	TI305	90.0	0.0 0.00	PI305	19.4	0.0 0.00	M 302	341.0	0.0 0.00
TW307	91.2	0.0	0.00	TI307	90.0	0.0 0.00	PI307	10.0	0.0 0.00	M 305	402.0	0.0 0.00
TW306	121.0	0.0	0.00	TI306	120.0	0.0 0.00	PI306	17.0	0.0 0.00	M 307	152.0	0.0 0.00
TW210A	180.0	0.0	0.00	0.0	0.0	0.00	0.0	0.0	0.00	M 304	411.0	0.0 0.00
TW304	91.0	0.0	0.00	0.0	0.0	0.00	PI304	21.0	0.0 0.00	M 301	904.0	0.0 0.00
TW209	204.8	0.0	0.00	TI209	210.0	0.0 0.00	PI209	210.0	0.0 0.00	0.0	0.0	0.00
TW208	244.0	0.0	0.00	TI208	235.0	0.0 0.00	PI208	220.0	0.0 0.00	0.0	0.0	0.00
TW210B	180.0	0.0	0.00	0.0	0.0	0.00	0.0	0.0	0.00	0.0	0.0	0.00
TW308	146.5	0.0	0.00	TI308	136.0	0.0 0.00	PI308	8.0	0.0 0.00	0.0	0.0	0.00
TW309	127.4	0.0	0.00	0.0	0.0	0.00	0.0	0.0	0.00	0.0	0.0	0.00
TW303	129.0	0.0	0.00	TI303	126.0	0.0 0.00	PI303	22.5	0.0 0.00	0.0	0.0	0.00
TW211	180.2	0.0	0.00	TI211	180.0	0.0 0.00	0.0	0.0	0.00	0.0	0.0	0.00
TW301	92.0	0.0	0.00	TI301	87.0	0.0 0.00	PI301	29.0	0.0 0.00	0.0	0.0	0.00
TW210	180.0	0.0	0.00	TI210	180.0	0.0 0.00	PI211	201.0	0.0 0.00	0.0	0.0	0.00
** BRINE SOURCE AND REINJECTION WELLS **												
WELL NO.	1	2	3	4	5	REINJECTION						
FLOW RATE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TEMPERATURE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
PRESSURE AT PUMP	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0

Figure 5-3 Printout of Data

1973. Due to the relatively low salinity of the brine, a geothermal model of the physical properties of water, designed by PFR and using those properties as listed by Keenan, et al., was used to determine the required properties of brine for processing.

Physical Properties of Hydrocarbons

A computerized listing of the physical properties for Isobutane, Propane and Isopentane, was obtained through the Lawrence Berkeley Laboratory, University of California. These properties included enthalpy, entropy, specific heat and specific volume as a function of both temperature and pressure. A permanent file was created in order to store these properties inside the computer. The format used for storage is described below.

The property file had eight columns, each containing one particular set of numbers:

<u>COLUMN</u>	<u>CONTENTS</u>
1	1 for isobutane or propane 0.8 for 80-20 mole percent isobutane/isopentane mixture
2	Temperature in degrees Fahrenheit 80 - 360° in increments of 5° for isobutane and the 80 - 20 mixture 50 - 245° in increments of 5° for propane
3	Pressure in psia 70 - 695 psia in increments of 5 psia for isobutane and the 80 - 20 mixture 100 - 690 psia in increments of 10 psia for propane
4	Specific volume in Su. ft./lbm
5	Enthalpy in Btu/lbm
6	Entropy Btu/lbm°F
7	Specific heat Btu/lbm°F
8	1: for outside the dome 2: for the vapor side of saturation line 3: for liquid side of saturation line

The properties of saturated liquid and vapor hydrocarbons were recorded separately following this same format.

Subroutine FLUPRO was used to read the required physical properties at given temperature and pressures. If required, linear interpolation between the recorded values was done.

The physical properties established at given temperatures and pressures were used in other subroutines to calculate the basic heat and mass balances and other appropriate quantities as described later.

Printout of Results

The results of the data reduction were printed out using program OUTPUT. There were two main printouts.

1. Page 1 of the printout (Figure 5-4) consisted of two parts, part one contained the data used in obtaining the results and part two contains the results of the data reduction. These results included heat and mass balances for individual units together with calculated values of LMTD, and the overall heat transfer coefficient for Exchangers B1 through B7. Results of the uncertainty analysis for Exchangers B1 through B6 were also included in this part. At the bottom of this page are the results of overall heat and mass balances for both the cold and hot sides of the primary heat exchangers as well as the hydrocarbon coolers.
2. Page 2 of the printout (Figure 5-5) contained the detailed information for each of the heat exchanger units involved. This included temperature, pressures, pressure drops, velocity, mass flow rates, and heat transfer coefficients, for both the tube side and shell side. Also state point information for each fluid was included.

Any unusual discrepancies in the results were printed out before the complete set of results was printed. These discrepancies included unacceptable behavior in any one or more than one heat exchanger. For example, if the vapor quality of the hydrocarbon condensing stream exiting the condenser is greater than zero, an error message is printed. This message also included the conditions necessary for total condensation, allowing changes to the process conditions to achieve complete condensation.

DATA REDUCTION MODELS

Heat and Mass Balance

One of the main objectives was to achieve heat balances between the cold and hot streams to $\pm 3\%$ without consideration of any bias or fixed type of error. Being a closed loop the mass of hydrocarbon was assumed to remain constant for a measured

TIME START : 7:00
TIME COMPL. : 7:40
OPERATOR : MARTINEL

** P8ENT TESTS COMPUTER OUTPUT **
NO. HOURS START OF SERIES : 0
TEST FLUID : I-C4

RUN / SERIES : 116/ 0
RUN DATE : 11-22-79
CURRENT DATE : 11-FEB-80

** REDUCED DATA **

PAGE 1

HEATERS			HEATERS			HEATERS			DESUPERHEATER, CONDENSER AND SUBCOOLER					
TEMPERATURES (DEG F)			PRESSURES (PSIA)			PRESSURE DROPS (PSI)			TEMPERATURES (DEG F)			PRESSURES (PSIA)		
BRINE		X	BRINE		X	BRINE		X	HYDROCARBON		X	HYDROCARBON		X
TW107	340.3	0.0	PI107	249.7	0.0	DPI104	5.8	0.0	TW208	248.0	0.0	PI208	229.7	0.0
TW104	289.0	0.0		0.0	0.0	DPI105	5.8	0.0	TW209	210.5	0.0	PI209	219.7	0.0
TW105	268.0	0.0		0.0	0.0	DPI104	5.8	0.0	TW210A	177.1	0.0		0.0	0.0
TW104	255.0	0.0		0.0	0.0	DPI103	5.8	0.0	TW210B	178.0	0.0		0.0	0.0
TW103	240.0	0.0		0.0	0.0	DPI102	5.8	0.0	TW211	177.8	0.0	PI211	209.7	0.0
TW102	220.0	0.0		0.0	0.0	DPI101	5.8	0.0	TW212	134.0	0.0	PI212	209.7	0.0
TW101	184.0	0.0	PI101	234.7	0.0		0.0	0.0		0.0	0.0		0.0	0.0
HYDROCARBON			HYDROCARBON			HYDROCARBON			COOLING WATER			COOLING WATER		
		X			X			X			X			X
TW201	140.0	0.0	PI201	425.0	0.0	DPI201	10.5	0.0	TW309	124.4	0.0		0.0	0.0
TW202	194.3	0.0		0.0	0.0	DPI202	14.7	0.0	TW308	154.0	0.0	PI308	21.7	0.0
TW203	224.0	0.0		0.0	0.0	DPI203	14.7	0.0	TW307	92.0	0.0	PI307	23.2	0.0
TW204	244.8	0.0	PI204A	414.8	0.0	DPI204	14.8	0.0	TW306	121.0	0.0	PI306	23.7	0.0
TW205	257.2	0.0	PI204B	585.1	0.0	DPI205	21.0	0.0	TW305	94.0	0.0	PI305	29.2	0.0
TW206	244.0	0.0	PI204C	519.5	0.0	DPI206	27.3	0.0	TW304	91.0	0.0	PI304	32.7	0.0
	0.0	0.0	PI204D	525.7	0.0		0.0	0.0	TW303	128.4	0.0	PI303	245.7	0.0
TW207	293.0	0.0	PI207	520.0	0.0		0.0	0.0	TW302	91.1	0.0	PI302	41.2	0.0
	0.0	0.0		0.0	0.0		0.0	0.0	TW301	91.7	0.0	PI301	41.7	0.0
FLOW METERS (LB/HR)			WATER FLOWS											
BRINE:														
M101	97073.5	0.0	M301	470094.7	0.0	M304	239044.3	0.0	M307	40750.1	0.0			
M201	89444.7	0.0	M302	170330.0	0.0	M305	284090.7	0.0	M310**	119704.7	0.0			

** HEAT DUTIES (IN BTU/HR) AND PERCENTAGE OF DIFFERENCE OF HYDROCARBON FROM WATER/BRINE **

UNIT	BRINE DUTY	HYDROCARBON DUTY	% DIFF.	QUALITY	LMTD	OVERALL U	% UNCERTAINTY ERRORS		
							BR DUTY	LMTD	OVERALL U
B1	3513674.3	3368877.0	-4.1	0.00000	34.03	357.71	8.8	10.2	19.0
B2	1940569.8	1845027.2	-4.9	0.00000	19.24	352.67	13.3	17.7	31.0
B3	1475160.1	1349549.3	-8.5	0.00000	12.17	420.11	16.6	30.2	46.7
B4	1282106.4	1027415.9	-19.8	0.00000	10.45	417.14	18.4	37.9	54.6
B5	2078905.0	1426870.7	-31.4	0.00000	14.14	444.32	12.8	33.8	44.6
B6	5125105.5	4726447.5	-7.8	1.00000	33.70	526.89	7.1	15.9	23.1
TOTAL	15435522.0	13744429.0	-10.8						
FEED PUMP ENERGY		4737821.0							
WATER DUTY									
B7	3740522.5	2794728.2	-25.7	1.00000	97.97	135.17	*****	*****	*****
B8	7137815.5	8090029.0	13.3	0.00000	*****	*****	*****	*****	*****
B10	6340438.5	4737821.0	-25.3	0.00000	*****	*****	*****	*****	*****
TOTAL	17238976.0	15622578.0	-9.4						

** OVERALL HEAT DUTIES (MBTU/HR) **

HIGH PRESSURE SIDE			HEAT REJECTION SIDE		
HEAT DELIVERED BY BRINE	:	15.417	HEAT DELIVERED BY HYDROCARBON	:	15.434
HEAT RECEIVED BY HYDROCARBON	:	15.037	HEAT RECEIVED BY COOLING WATER	:	14.280
X DIFF :		-2.5	X DIFF :		4.1

Figure 5-4. Computer Output, Page 1

TIME START : 7100
TIME COMPL. : 7140
OPERATOR : MARTINEL

** PSEMT TESTS COMPUTER OUTPUT **
NO. HOURS START OF SERIES : 0
TEST FLUID : I-C4

RUN / SERIES : 116/ 0
RUN DATE : 11-22-79
CURRENT DATE : 11-FEB-80

PAGE 2										
EXCHANGERS :	B1	B2	B3	B4	B5	B6	B7	B8	B10	
** OVERALL PARAMETERS **										
DUTY (BTU/HR)	3513674.3	1960569.8	1475160.1	1282106.4	2078905.0	5125105.5	3760522.5	7137815.5	6340638.5	
LMTD (DEG.F) (3)	34.034	19.261	12.164	10.649	16.139	33.702	97.970	*****	*****	
OVERALL U (3)	357.7	352.7	420.1	417.1	446.3	526.9	135.2	*****	*****	
FOULING FACTOR (3)	0.000000	0.000000	0.000000	0.000000	0.000000	0.000000	0.000000	*****	*****	
** TUBE SIDE PARAMETERS ** (BRINE / WATER)										
FLOW RATE (LB/HR)	97073.5	97073.5	97073.5	97073.5	97073.5	97073.5	60750.1	286090.7	170330.0	
MASS VELOCITY (LB/HR-FT2)	746788.1	746788.1	746788.1	746788.1	746788.1	746788.1	475013.2	1007607.2	443944.0	
INLET VELOCITY (FT/SEC)	3.452	3.493	3.321	3.345	3.577	3.653	2.141	4.525	1.994	
INLET DENSITY (LB/FT3)	60.097	59.396	58.917	58.512	57.993	56.791	61.641	61.856	61.838	
INLET TEMPERATURE (DEG.F)	184.0	220.0	240.0	255.0	268.0	289.0	92.0	96.0	91.1	
EXIT TEMPERATURE (DEG.F)	220.0	240.0	255.0	268.0	289.0	340.3	154.0	121.0	128.4	
PRESSURE DROP (PSI)	5.830	5.830	5.830	5.830	5.830	5.830	1.500	5.500	-204.500	
HEAT TRANSFER COEF.	1228.3	1279.3	1308.5	1330.3	1355.2	1401.8	769.0	1293.3	709.0	
** SHELL SIDE PARAMETERS ** (HYDROCARBON) - FLOW RATE (LB/HR) 89444.7										
MASS VELOCITY (LB/HR-FT2)	634499.6	634499.6	634499.6	634499.6	634499.6	317249.8	660610.6	12873.7	457996.3	
INLET VELOCITY (FT/SEC)	5.438	6.018	6.601	7.167	7.814	4.323	84.955	1.593	5.952	
EXIT VELOCITY (FT/SEC)	6.018	6.601	7.167	7.814	8.445	12.337	81.730	0.167	3.985	
INLET DENSITY (LB/FT3)	32.413	29.287	26.700	24.591	22.554	20.387	2.160	2.245	21.373	
EXIT DENSITY (LB/FT3)	29.287	26.700	24.591	22.554	20.387	7.143	2.245	21.373	31.925	
INLET TEMPERATURE (DEG.F)	140.0	194.3	226.0	244.5	257.2	264.0	248.0	210.5	177.8	
EXIT TEMPERATURE (DEG.F)	194.3	226.0	244.5	257.2	264.0	293.0	210.5	177.4	134.0	
SAT. PRESS. AT INLET TEMP.	127.3	242.1	336.2	401.8	451.3	487.4	415.1	287.5	201.4	
SAT. PRESS. AT EXIT TEMP.	242.1	336.2	401.8	451.3	487.4	0.0	287.5	200.9	120.9	
INLET PRESSURE (PSIA)	639.7	629.2	614.5	599.8	583.0	562.0	229.7	219.7	209.7	
EXIT PRESSURE (PSIA)	629.2	614.5	599.8	583.0	562.0	534.7	219.7	209.7	209.7	
SAT. TEMP. AT INLET PRESS.	0.0	0.0	0.0	0.0	0.0	0.0	189.5	185.8	181.3	
SAT. TEMP. AT EXIT PRESS.	0.0	0.0	0.0	0.0	0.0	0.0	185.5	181.3	181.3	
INLET QUALITY	0.000000	0.000000	0.000000	0.000000	0.000000	0.000000	1.000000	1.000000	0.000000	
EXIT QUALITY	0.000000	0.000000	0.000000	0.000000	0.000000	1.000000	1.000000	0.000000	0.000000	
INLET ENTHALPY (BTU/LB) (1)	22.5	61.8	83.7	100.2	114.5	137.8	194.7	152.6	72.8	
ENTHALPY BASED ON BRINE/H2O(2)	61.8	83.7	100.2	114.5	137.8	195.1	152.6	72.8	2.0	
ENTHALPY BASED ON DATA FILE(2)	58.5	82.6	98.8	111.7	123.2	190.6	174.2	51.6	19.9	
EXIT ENTHALPY (BTU/LB)	60.1	82.6	98.8	111.7	130.5	190.6	163.4	62.2	19.9	
PRESSURE DROP (PSI)	10.500	14.700	14.700	16.800	21.000	27.300	10.000	10.000	0.000	
HEAT TRANSFER COEF.	611.2	583.0	780.6	761.5	852.4	1162.5	177.1	*****	*****	
** BRINE SOURCE AND REINJECTION WELLS **										
WELL NO.	1	2	3	4	5	REINJECTION				
FLOW RATE	0.0	0.0	0.0	0.0	0.0	0.0				
TEMPERATURE	0.0	0.0	0.0	0.0	0.0	0.0				
PRESSURE AT PUMP	0.0	0.0	0.0	0.0	0.0	0.0				

NOTES :

- (1) FOR EXCH. B2-B6, B8-B10 : INLET ENTHALPY IS DETERMINED FROM EXIT ENTHALPY BASED ON BRINE/WATER OF UPSTREAM EXCH.
- (2) EXIT ENTHALPIES FOR EXCH. B1-B10
- (3) LMTD, U AND FOULING FACTOR FOR EXCH. B8 AND B10 ARE NOT OUTPUT BECAUSE OF NON-LINEARITY OF TEMP. PROFILE
- (4) DENSITY IS BASED ON VALUE FROM DATA FILE

Figure 5-5 Computer Output, Page 2

flow rate. The same assumption applies to the brine stream. The cooling water stream was divided into three streams, one for each cooler; the flow rate for each was measured separately.

Heat duties for each of the three processes streams, brine, hydrocarbon and cooling water were obtained as follows:

Brine or cooling water:

$$Q = \dot{m} c_p \Delta T$$

where $Q = \text{Btu/hr}$
 $\dot{m} = \text{lbm/hr of brine or water}$
 $c_p = \text{sp. heat in Btu/lbm}^\circ\text{F}$
 $\Delta T = (T_{\text{out}} - T_{\text{in}}), \text{ Temperature in degrees Farenheit}$

Hydrocarbon:

$$Q = \dot{m} \Delta h$$

where $Q = \text{Btu/hr}$
 $\dot{m} = \text{lbm/hr of hydrocarbon}$
 $\Delta h = (H_{\text{out}} - H_{\text{in}}) \text{ in Btu/lbm}$

Values of enthalpies were obtained from the property data bank for the given conditions of temperature and pressure.

Heat balances were done for each individual heat exchanger using the heat rejected by the hot streams and that received by the cold streams; from these a percentage difference between the heat duties of the two streams was obtained.

$$\text{Percentage Difference} = \frac{Q_c - Q_h}{Q_h} \times 100$$

where $Q_c = \text{heat duty of cold stream, Btu/hr}$

$Q_h = \text{heat duty of hot stream, Btu/hr}$

Similarly, separate overall balances were done for both the geothermal brine to hydrocarbon heat exchangers as well as the heat rejection side, using the overall inlet and outlet conditions of the streams involved.

Heat Transfer Coefficient

The following equation was used to determine the overall heat transfer coefficients in the heaters and the desuperheater. The coefficients in the

condenser and subcooler were not reduced by computer due to the lack of knowledge of the LMTD.

$$Q = UA \text{ LMTD}$$

where Q = heat exchanger duty (Brine/Water)
 U = overall heat transfer coefficient
 A = Heat transfer difference
 LMTD = Log mean temperature difference

This relationship yields: $U = (A \cdot \text{LMTD})$

This overall heat transfer coefficient can also be formulated in terms of thermal resistances taken in series:

$$U = 1/((1/h_{io}) + (1/h_o) + r_{\text{foul}} + r_{\text{wall}})$$

where h_{io} = tube side film coefficient based on the outside
 tube surface area

h_o = shell side film coefficient

r_{wall} = thermal resistance of tube wall

r_{foul} = thermal resistance from tube side fouling

The shell side fouling term was neglected based on the assumption that the hydrocarbon fluids are nonfouling in nature.

The shell side coefficient, h_o , was determined by

$$\frac{1}{h_o} = \frac{1}{U} - \frac{1}{h_{io}} - r_w - r_f$$

The tube side coefficient was calculated from a semi-empirical Nusselt type correlation specially derived for water streams⁽¹⁾. The accuracy of this correlation has been found to be better than 7.5%.

$$Nu = 0.057 (Re)^{.7} (Pr)^{.5}$$

where Nu = Nusselt Number
 Re = Reynolds Number
 Pr = Prandtl Number

(1) Lawrence and Sherwood, I and EC, Vol. 23, 1931.

Pressure Drop

The pressure drops for the heater units were measured using differential pressure gauges. Also the inlet and outlet pressures for both brine and hydrocarbon were measured at the ends of the heat exchanger field. With the failure of the differential gauges, a method recommended by the plant manager was used to estimate the pressure drop of each heater from the overall measured pressure drop.

This method consisted of calculating the average density of hydrocarbon in each exchanger and then apportioning the overall pressure drop in the ratio of the inverse square of the average density. The differential pressure gauges also failed on the brine side and the overall pressure drop was distributed equally over the six exchangers. The overall shell-side (hydrocarbon) pressure drop measured was approximately twice the value used in design.

Condensing Coefficient

The overall heat transfer coefficient for the condenser was not determined by computer analysis. The mean temperature driving force for the condenser is quite complex because the hydrocarbon underwent desuperheating, condensing and subcooling. There are two basic methods used by industry to determine the design or performance of this type of unit. One method uses a weighted mean temperature driving force, the other method uses a stepwise procedure. Both methods are complicated by the possibility of wet wall condensation.

For the purposes of these tests a simple procedure was used to hand calculate the overall condensing coefficient. This procedure assumes negligible desuperheating and subcooling. The log mean temperature is calculated by taking the saturation temperature of hydrocarbon at the measured condenser pressure and the two terminal temperatures of cooling water. The following schematic explains this method:

$$LMTD = \frac{TDH - TDC}{\ln \frac{TDH}{TDC}}$$

and
$$U = \frac{Q}{(LMTD)A}$$

Where Q = Heat received by water
A = Surface area

Uncertainty Analysis

The accuracy of a measured value is the degree of agreement between this number and the number that represents the true value of that quality; it may be expressed in either absolute or relative terms. In case of a quantity whose value is determined using mathematical correlations from some measured or known variables, this accuracy is a strong function of the accuracy of the variables used.

Uncertainty analysis on the Power System Equipment Module Test results was performed to determine the random errors in the calculated quantities propagated by the known inaccuracies of the measured data.

Uncertainty considerations proved to be critical in reviewing the data. The uncertainties resulting from determination of heat duties, overall heat transfer coefficients and the log mean temperature differences are of principal interest. The estimated errors include considerations of only random errors.

Heat Exchanger Duty. Heat exchanger duty can be determined from the following equation:

$$Q = Wc_p(T_{out} - T_{in}) \text{ or } Q = Wc_p\Delta T$$

Where W = Flow rate, LB/HR

c_p = Specific heat, BTU/LB °F

ΔT = Temperature difference, °F

Since the results of the test operations were based on the brine duty, the uncertainty analysis was performed only for the brine duty.

With logarithmic differentiation of the above equation and some algebraic manipulation, the expression for the error, called Root-Sum Square method, is expressed as:

$$\frac{\delta Q}{Q} = \left[\left(\frac{\delta W}{W} \right)^2 + 4 \left(\frac{T}{\Delta T} \right)^2 \left(\frac{\delta T}{T} \right)^2 \right]^{\frac{1}{2}}$$

Where δW - Error in the measurement of flow by an orifice meter

δT = Error in the measurement of temperature by mercury bulb thermometer

T = Average temperature

ΔT = Heat-exchanger stream-wise temperature difference

Fixed values for δW and δT were used, these values also include allowances for human error.

$$\delta W = \pm 2.5 \text{ GPM}$$

$$\delta T = \pm 1^\circ \text{F}$$

Log Mean Temperature Difference. In similar fashion, the error function for LMTD was found to be

$$\frac{\delta \text{LMTD}}{\text{LMTD}} = 2 \left[\left(\frac{\delta T}{T} \right)^2 \left(\frac{T^2 (\ln^a/b + b/a - 1)^2 + (\ln^a/b + 1 - a/b)^2}{((a-b) \ln^a/b)^2} \right) \right]^{\frac{1}{2}}$$

Where $T = \frac{\sum T}{n}$ average of four temperatures

$$a = \text{TDH} = T_{\text{in}} (\text{Brine}) - T_{\text{out}} (\text{Hydrocarbon})$$

$$b = \text{TDC} = T_{\text{out}} (\text{Brine}) - T_{\text{in}} (\text{Hydrocarbon})$$

$$\delta T = \pm 1^\circ \text{F}$$

Overall Heat Transfer Coefficient. Since the overall heat transfer coefficient is a function of heat duty and LMTD, its uncertainty can easily be computed by taking the sum of the uncertainties of heat duty and LMTD as follows:

$$\frac{\delta U}{U} = \frac{\delta Q}{Q} + \frac{\delta LMTD}{LMTD}$$

The above equation produces the maximum possible error that can be anticipated in the determination of U. The nominal uncertainty errors in U can be found by using the root sum square method as explained below:

$$\frac{\delta U}{U} = \left[\left(\frac{\delta Q}{Q} \right)^2 + \left(\frac{\delta LMTD}{LMTD} \right)^2 \right]^{\frac{1}{2}}$$

Results of uncertainty analysis on U obtained by using the two methods mentioned above are reported with the other results in Table 5-1.

Conclusions. Uncertainty Analysis was performed only for brine to hydrocarbon heat exchangers B1 and B6. Due to the supercritical cycle and the nature of the hydrocarbon itself the heat duties in exchangers B2 through B5 behaved rather non-linearly and the LMTD's for these exchangers were unacceptably low, thereby eliminating the value of any uncertainty analysis.

The table below summarizes the uncertainty results expected for exchangers B1 and B6.

<u>Exchanger</u>	<u>Heat Duty</u>	<u>% Uncertainty LMTD</u>	<u>Overall 'U'</u>
B1	±6.0	±5.0	±7.0
B6	±4.0	±6.0	±10.0

Prior to the beginning of the tests it was estimated that the uncertainties associated with the heat duty and the overall heat transfer coefficient would be ±1.5% and ±8.5%. Due to the loss of certain accurate instruments these uncertainties could not be attained. However, with reference to the operating conditions, the results of the uncertainty analysis are within the normal, experimental accuracy expected for industrial sized units.

Table 5-1

HEAT TRANSFER COEFFICIENT UNCERTAINTIES

Run Number	Brine Flow LB/HR	INLET Brine Temp °F	I-C4 Flow LB/HR	Temp °F	Press PSIG	Q		LMTD		Nominal		U Maximum Possible		% Diff. Between Duties
						B1	B6	B1	B6	B1	B6	B1	B6	
52	92331.7	344.2	85630.9	300.4	560	5.4	3.5	4.0	5.3	6.7	6.4	9.6	11.1	-3.4
53	92543.6	343.0	84974.0	291.3	550	5.6	3.6	4.1	5.1	6.9	6.2	9.9	10.9	-4.9
56	87672.9	345.2	84838.4	296.5	550	5.5	3.5	4.2	4.9	6.9	6.0	10.0	10.3	-2.6
57	87672.9	344.2	84572.4	298.0	550	5.4	3.6	4.1	5.0	6.8	6.2	9.9	10.7	-1.4
58	87672.9	345.0	84971.4	299.0	550	5.5	3.5	4.2	4.9	6.9	6.0	9.9	10.5	-2.1
59	87642.9	345.0	84739.4	299.0	552	5.4	3.6	4.1	4.8	6.7	6.0	9.9	10.3	-1.4
60	90083.6	346.0	84338.9	299.2	560	5.4	3.5	4.2	5.0	6.8	6.1	10.0	10.5	-2.9
61	90077.4	347.6	84295.0	300.8	560	5.3	3.5	4.3	4.9	6.8	6.0	10.1	10.4	-2.9
62	90083.6	345.6	84183.0	299.0	550	5.4	3.4	4.3	5.0	6.9	6.1	10.0	10.7	-5.0
63	90089.8	344.6	84166.0	300.8	560	5.2	3.5	4.2	5.1	6.7	6.2	10.0	10.8	-4.3
83	89965.1	346.8	84521.8	308.4	560	4.8	4.0	4.1	4.6	6.3	6.1	9.5	10.1	0.2
84	89952.4	346.4	84234.5	309.2	560	5.0	4.1	4.5	6.5	6.1	9.8	9.9	0.7	0.7
85	89952.5	350.0	84373.7	312.0	552	4.9	3.9	4.0	4.6	6.3	6.0	9.4	10.0	1.0
86	89952.5	348.6	84199.8	310.4	560	4.9	4.0	4.2	4.6	6.4	6.1	9.7	10.0	-0.2
87	89946.2	346.8	83872.7	309.5	552	4.9	4.1	4.0	4.3	6.3	6.0	9.5	9.6	6.7
88	89946.2	347.4	84001.6	309.8	550	4.9	3.9	4.1	4.6	6.4	6.0	9.5	10.0	0.8
92	89958.8	338.2	84619.9	306	550	4.9	4.6	3.9	4.9	6.3	6.7	9.1	10.9	3.2
94	89990.2	340.4	86047.4	336	560	5.2	4.2	4.2	4.3	6.7	13.7	9.9	26.3	8.8
96	89958.8	346.5	86278.6	303	555	5.2	3.8	4.1	4.6	6.6	5.0	9.7	9.9	-1.0
97	89946.2	346.0	86074.5	303.8	560	5.5	3.8	4.2	4.7	6.9	6.0	10.1	10.2	-0.6
98	89952.5	348.0	85921.3	305.2	552	5.6	3.6	4.1	4.7	6.9	5.9	9.9	10.1	-1.0
99	89952.5	348.8	86337.8	305.6	560	5.5	3.7	4.3	4.7	6.9	6.0	10.2	10.0	-1.1
110	98114.4	337.8	89054.7	291.0	511	5.8	4.0	4.4	4.6	7.3	6.1	10.6	10.0	-2.7
112	97132.6	337.4	89269.3	292.0	510	5.8	4.0	4.4	4.7	7.3	6.2	10.5	10.2	-0.3
113	97129.2	339.6	89269.3	289.0	510	5.9	3.9	4.4	4.4	7.4	5.8	10.6	9.7	-3.4
114	97108.4	339.2	89159.8	292.0	520	5.4	3.6	4.2	5.5	6.8	6.6	10.0	11.5	-2.6

Table 5-1 (continued)

Run Number	Brine Flow LB/HR	INLET Brine Temp °F	I-C4 Flow LB/HR	Temp °F	Press PSIG	Q		LMTD		Nominal		U Maximum Possible		% Diff. Between Duties
						B1	B6	B1	B6	B1	B6	B1	B6	
115	97108.4	339.2	89043.1	293.3	520	5.4	3.7	4.1	5.5	6.8	6.6	9.8	11.5	-2.4
116	97073.5	340.3	89446.7	293.0	520	5.6	3.9	4.3	4.5	7.0	5.9	10.3	9.8	-2.5
129	97108.4	340.8	99304.0	282.4	500	6.4	3.7	4.7	4.1	7.9	5.5	11.4	9.0	2.2
130	97125.7	341.0	98190.7	285.4	505	6.3	3.7	4.8	4.2	7.9	5.6	11.5	9.1	1.7
131	97108.4	341.4	98083.8	285.0	505	6.3	3.6	4.8	4.2	7.9	5.5	11.4	9.2	1.8
132	98294.0	341.0	99846.0	283.0	500	6.5	3.7	4.6	4.1	7.9	5.5	11.2	9.0	1.7
133	97087.5	342.0	99270.6	283.4	500	6.4	3.6	4.8	4.1	8.0	5.4	11.6	9.0	1.7
134	98065.4	342.0	99590.8	282.8	500	6.4	3.6	4.7	4.1	7.9	5.4	11.4	8.9	0.7
135	97080.5	342.0	99467.6	284.0	500	6.4	3.6	4.9	4.1	8.1	5.4	11.7	8.9	2.3
136	98044.3	344.0	105654.1	288.0	5.5	6.5	3.5	4.9	4.1	8.1	5.4	11.8	9.0	4.2
184	91977.1	340.0	82752.9	302.6	580	6.5	4.1	4.1	6.4	7.6	7.6	10.5	13.5	-3.3
185	91949.7	345.2	82582.5	307.2	585	6.4	4.1	4.2	5.5	7.6	6.8	10.6	11.7	-4.1
186	91949.7	345.3	82506.1	307.4	585	6.3	4.1	4.2	5.5	7.5	6.8	10.5	11.8	-4.4
187	91949.7	346.4	82518.8	308.6	590	6.2	4.0	4.2	5.5	7.5	6.8	10.5	11.7	-4.8
188	91935.9	346.4	82518.8	308.4	590	6.3	4.0	4.2	5.5	7.6	6.8	10.5	11.7	-4.7
189	91935.9	346.2	82493.3	308.4	590	6.3	4.0	4.2	5.4	7.6	6.7	10.9	11.6	-4.7
190	91922.1	340.0	82417.0	304.2	575	6.6	4.3	4.3	6.1	7.8	7.5	10.9	12.9	-1.7
191	91911.7	340.2	82404.3	304.4	575	6.6	4.3	4.3	6.0	7.8	7.4	10.9	12.8	-1.5
192	93117.6	349.4	82328.2	309.6	585	6.4	3.8	4.4	5.4	7.7	6.6	10.9	11.5	-3.3
193	93096.5	354.0	82214.2	311.0	590	6.2	4.8	4.5	6.4	7.6	8.0	11.1	13.6	-4.5
194	93068.4	354.0	82004.2	312.0	580	6.4	3.6	4.5	5.1	7.8	6.2	11.0	10.8	-4.0
195	93075.4	354.0	82004.2	312.0	580	6.3	3.6	4.5	5.0	7.7	6.2	11.0	10.6	-4.0
196	93082.5	354.0	82057.8	312.0	580	6.3	3.6	4.5	5.0	7.7	6.2	10.9	10.6	-4.0
197	93082.5	354.0	81977.3	312.0	580	6.3	3.6	4.5	5.1	7.7	6.2	11.1	10.8	-4.1
198	93054.3	354.0	81843.5	311.0	580	6.3	3.6	4.4	4.9	7.7	6.1	10.8	10.4	-4.4
200	87877.9	354.0	81603.7	320.0	595	7.1	4.1	4.4	5.0	8.3	6.5	11.3	10.8	5.6
201	87877.9	352.1	79044.5	319.0	595	6.9	4.2	4.5	5.1	8.2	6.6	11.4	11.1	3.0
202	87877.9	352.1	79053.6	329.2	595	6.9	4.2	4.5	6.0	8.2	7.3	11.4	12.8	4.3
203	87919.8	351.2	79011.4	318.0	595	6.7	4.2	4.4	5.2	8.0	6.7	11.1	11.2	2.2

Table 5-1 (continued)

Run Number	Brine Flow LB/HR	INLET Brine Temp °F	I-C4 Flow LB/HR	Temp °F	Press PSIG	Q		LMTD		Nominal		U Maximum Possible		% Diff. Between Duties
						B1	B6	B1	B6	B1	B6	B1	B6	
204	87905.8	357.8	79044.5	329.6	595	6.7	3.7	4.5	5.6	8.1	6.7	11.3	11.8	3.9
205	87877.9	353.0	79044.5	326.5	595	6.8	4.2	4.5	5.6	8.1	7.0	11.2	12.0	3.2
206	87863.9	352.4	79093.1	330.6	595	6.8	4.3	4.4	5.6	8.1	7.0	11.2	12.0	3.2
216	80876.5	342.0	56418.3	300.2	585	6.2	4.1	4.3	5.4	7.5	6.8	10.6	11.5	-25.8
217	80876.5	342.0	56291.2	301.6	585	6.2	4.1	4.4	5.4	7.6	6.8	10.8	11.5	-25.6
218	80851.5	342.0	56217.4	302.8	580	6.2	4.0	4.4	5.7	7.6	7.0	10.8	12.1	-25.9
219	80963.1	342.0	56199.0	302.8	580	5.6	4.1	4.6	5.5	7.2	6.8	10.7	11.7	-17.6
220	80957.0	342.4	56180.6	303.0	580	5.6	4.0	4.6	5.5	7.2	6.8	10.7	11.7	-17.6
221	81322.4	342.2	56272.7	302.8	590	4.2	4.1	5.2	5.3	6.7	6.7	11.4	11.4	-32.8
222	81299.1	343.4	56199.0	303.0	590	4.3	4.1	5.3	5.3	6.8	6.7	11.4	11.4	-32.7
223	81328.2	342.0	56291.2	302.6	590	4.2	4.2	5.3	5.3	6.7	6.7	11.4	11.3	-32.9
224	94008.3	344.0	73591.4	301.2	565	7.0	3.8	4.9	4.9	8.5	6.2	12.0	10.6	-7.1
225	93978.8	344.09	72442.4	300.8	560	6.8	3.8	4.9	4.9	8.4	6.2	11.9	10.6	-12.5
226	93934.2	344.0	72033.6	301.0	560	7.4	3.8	5.1	4.9	8.9	6.2	12.6	10.6	-14.4
227	93934.2	343.0	64601.0	300.2	560	7.4	3.9	5.4	4.9	9.1	6.3	13.0	10.6	-20.3
228	93971.4	344.0	64948.7	300.8	560	6.9	3.9	5.2	4.9	8.6	6.3	12.5	10.5	-22.7
229	93971.4	334.0	72179.9	300.6	560	7.5	4.8	5.1	5.5	9.1	7.3	12.7	10.5	-23.6
230	93971.4	345.0	72304.9	300.6	560	7.5	3.8	5.1	4.9	9.1	6.2	12.7	10.5	-23.6
236	92585.5	344.1	67071.0	303.6	582	7.1	4.1	5.0	5.1	8.7	6.5	12.2	11.1	-25.6
237	92566.8	344.2	67335.3	304.6	580	7.2	4.1	4.8	5.0	8.6	6.5	12.0	10.9	-24.6
238	92566.8	344.0	67359.4	304.2	580	7.4	4.1	4.9	5.0	8.8	6.5	12.2	10.9	-24.6
239	92521.7	343.0	66853.5	305.2	578	7.3	4.3	4.9	5.0	8.8	6.6	12.2	10.9	-25.6
240	92581.7	344.0	67142.9	305.2	570	6.9	4.2	4.9	4.9	8.5	6.4	12.0	10.7	-25.2
241	92581.7	344.0	67118.9	305.2	570	6.9	4.2	5.1	4.9	8.6	6.4	12.3	10.7	-25.3
242	92506.6	344.0	66673.1	304.2	570	7.2	4.3	5.0	4.8	8.7	6.4	12.4	10.5	-26.5
243	92506.6	344.0	66673.1	304.2	570	7.2	4.3	5.0	4.8	8.7	6.4	12.3	10.5	-26.5
247	86722.3	333.0	64202.8	312.6	560	6.3	5.4	4.8	6.8	7.9	8.7	11.4	14.6	-18.0
248	86701.6	330.2	64202.8	311.0	560	6.6	5.8	5.0	7.0	8.3	9.1	12.0	15.2	-18.0
249	86722.3	331.8	62943.9	311.5	560	6.6	5.4	5.0	6.9	8.3	8.7	12.0	14.7	-19.4

Table 5-1 (continued)

Run Number	Brine Flow LB/HR	INLET Brine Temp °F	I-C4 Flow LB/HR	Temp °F	Press PSIG	Q		LMTD		Nominal		U Maximum Possible		% Diff. Between Duties
						B1	B6	B1	B6	B1	B6	B1	B6	
250	86776.9	330.8	62943.9	311.0	555	6.2	5.6	5.1	6.9	8.0	8.8	11.9	15.0	-19.0
251	86708.5	331.0	62943.9	312.2	555	6.6	5.7	5.0	7.0	8.3	9.0	11.9	15.2	-18.7
252	86674.1	331.0	62806.4	312.0	552	6.6	5.8	5.0	7.0	8.3	9.1	11.9	15.1	-18.8
253	92529.2	332.5	63451.3	320.1	560	6.4	8.4	4.8	9.2	8.0	12.4	11.6	20.3	-3.7
254	92506.6	330.0	64543.8	320.0	560	6.6	9.8	4.5	10.5	8.0	14.4	11.1	23.1	0.1
255	92506.6	332.8	64566.0	321.8	560	6.3	9.0	4.4	9.8	7.7	13.3	10.9	21.6	-1.3
256	92506.6	332.6	64543.8	322.2	560	6.2	9.4	4.4	10.0	7.6	13.7	10.8	22.2	-1.0
257	92788.5	332.2	64477.4	302.4	560	5.0	9.9	4.7	5.8	6.8	11.5	10.6	15.2	-20.4
258	92354.1	332.6	64455.3	322.4	560	6.3	9.8	4.4	10.2	7.7	14.1	10.9	22.6	-1.4
259	92461.3	331.2	64160.6	321.6	555	6.3	10.5	4.5	10.7	7.7	15.0	11.0	23.8	-0.6
260	92446.1	331.8	64138.6	321.6	555	6.6	10.3	4.6	10.3	8.0	14.5	11.3	23.0	-0.8
262	72297.1	338.0	52686.2	301.0	557	7.2	3.7	5.7	6.5	9.2	7.5	13.5	13.6	-19.6
263	72257.1	338.0	52745.8	300.2	550	7.4	3.7	5.4	6.5	9.2	7.5	13.2	13.6	-18.1
264	72319.8	340.0	53011.7	313.2	550	6.5	4.3	4.8	5.6	8.1	7.0	11.6	12.0	-20.5
265	72314.1	338.0	53498.3	316.0	550	6.3	4.9	4.8	5.9	7.9	7.6	11.5	12.8	-20.7
266	72314.1	341.0	53066.2	316.4	550	6.5	4.5	4.9	5.7	8.1	7.3	11.7	12.2	-21.1
267	72285.7	340.0	52991.8	316.8	550	6.2	4.8	4.5	5.8	7.6	7.5	11.0	12.5	-21.4
268	72314.1	341.0	53066.2	317.0	550	6.5	4.4	4.9	5.8	8.1	7.3	11.7	12.4	-20.7

Blank

5-20

Section 6

SITE AND FACILITY DESCRIPTION

SITE LOCATION

The site is located approximately 9 miles east of Holtville and 6 miles north of the U.S. - Mexico border. The exact location is S $1\frac{1}{2}$ NW $\frac{1}{4}$ SW $\frac{1}{4}$ Section 7 T16S R17W San Bernardino Baseline & Meridian. This overall location is shown on Figure 6-1. The site characteristics are those of desert land from which all brush has been removed. The site slope increases two feet in every hundred to the east. The soil has enough cohesive properties to stand when first excavated for foundations. The soil is alkaline. Corrosion on underground piping without external protection is from moderate to severe.

The environment is typical Southern California low desert with temperatures ranging from typical 110°F summer days to below freezing at night in the winter. The air is clean with no smog; however, winds blow especially in the spring months sometimes making construction work impossible due to blowing sand. The humidity normally is quite low, however in the late summer, tropical storms in the Gulf of California cause the area to be humid. The design dry bulb and mean coincident wet bulb in the summer are 112°F and 71°F, respectively. The design wet bulb temperature was chosen at 81°F.

The Binary Module Test Site was placed in the southwest corner of the fenced Magma Electric Company East Mesa Geothermal Plant site. This location is shown on Figure 6-2.

The geothermal fluid characteristics are as follows:

Bottom Hole Temperature	365 - 385°F
Total Dissolved Solids (TDS)	10,000 RPM
Non-condensable gas content	approx. 0.03 percent by weight

Figure 6-1 Test Site Location

FACILITY OVERVIEW

Brine Supply and Return System

The Brine was supplied to the Binary Module Test by means of a 3 inch line, run underground and insulated for a distance of 350'.

This line originates immediately down stream of Magma Electric Company's Sand Filter on the main brine supply to their facility. A 3 inch, 300# flanged gate valve was the tie-in point.

The return line was also 3 inches and was run parallel and adjacent to the supply line. The return line was uninsulated.

The return line terminates through a 3 inch - 300# flanged gate valve on Magma Electric Company's discharge main brine line to their injection wells.

Two supply points exist, one just down stream from Magma Electric's brine booster pump, the other up stream of this pump and immediately downstream of Magma Electric's Sand Filter. Likewise, two brine return points exist, one just up stream of Magma Electric's brine injection pumps, the other into Magma Electric's Plant brine by-pass. This Brine supply and return system allowed test operations independent of Magma Electric Company's operations. The run of these supply and return lines were as shown on Figure 6-3.

Service Water System

The Service Water System came from Magma Electric's deep cooling pond through a 2 and 3-inch supply line connected to Magma Electric's pressurized water supply to their well pads. This water was used for cooling tower makeup, fire system, and general utility purposes. The Service water system is shown on Figure 6-4.

Normal water pressure ranged from 58 psig to 38 psig; however, approximately 7 feet of static head was available enabling about 50 gpm of makeup requirements in the event Magma Electric's pressurized water system was not operating.

The water supply system initially received water with 973 ppm Total Dissolved Solids (TDS). The composition of this water is shown on Table 3-8.

Since no Magma Plant operations occurred during test operations, the water used during their operations maintained substantially the same composition.

NUMBER	DATE	REFERENCE DRAWINGS

1 ISSUED FOR APPROVAL EAST MESA		3-79
P.S.E.M.T.		
SURVEY & PLOT PLAN		
EAST MESA, CALIFORNIA		L.B.L. / EPRI
COLLEY engineers & constructors, inc.		

Figure 6-3 Geothermal Brine Supply and Return System

Figure 6-4 Service Water System

Blowdown from the PSEMT cooling water loop was delivered to Magma Electric Company's plant blowdown line to the Warren Drain.

Site Electrical System

Electrical power was delivered via a 3" underground conduct with 3,500 MCM wires from Magma's 350 KVA, 480 volt-3 phase-outside switch gear panel. The feed ran to the Binary Module Test switch gear metering station and to the 350 AMP main breaker. This panel location is shown on Figure 6-5.

From the main breaker it was fed through a gutter to seven combination starters which controlled the various motors ranging in size from 3 to 125 horsepower.

A 15 KVA 480-120/240 volt transformer was also fed from the gutter. This transformer supplied lighting and control power as well as utility outlets.

The main switchgear was located outside the operating and restricted area and was classified for General Purpose enclosures. Seal off fittings were used on all conducts leaving the panel.

Due to the presence of hydrocarbons, the operating area was classified Class I Group D Division II. All motors and fittings within this operating area were explosion proof. Detail of the Electrical Supply System is shown on Figure 6-6.

Damage Control Systems

A possible fire hazard exists should a hydrocarbon leak occur from the hydrocarbon circulating pump seal, valve packings, or flange gaskets. These areas were all located in close proximity to this circulating pump and its motor as shown on Figure 6-7.

The damage control system was predicated on the use of the cooling water circulating pump drawing water from the cooling tower basin, and in the event this pump is inoperative, an internal combustion engine driven pump was installed to supply sufficient pressure to spray heads while drawing water from the water supply system. This arrangement is shown on Figure 6-4.

As shown on Figure 6-8, the spray heads are directed in the area of the hydrocarbon pump seal, flanged connections near the pump, and on the uninsulated areas of pipe below the condenser.

Figure 6-5 Electric Panel Location

Figure 6-7 Potential Hydrocarbon Leak Locations

Should a fire occur, these heads should extinguish the fire and cool the uninsulated piping so excessive hydrocarbon pressures would not build up.

As part of the Damage Control System, a hydrocarbon relief system was provided. The basis for this relief system was to relieve any excessive pressure build-up caused by continuous fire exposure.

Process Fluid Supply, Loading, and Unloading

The process fluids, isobutane and isopentane were shipped in 55 gallon containers. The isobutane was shipped in cylinders which had a valve in the vapor space and another valve on an eduction tube which ran to the bottom of the cylinder. The isopentane was shipped in standard 55 gallon drums.

A Corken, positive displacement vane type pump was used to fill the system. This pump has a capacity of 5 gallons per minute. This pump in conjunction with a Neptune LPG meter was installed on a skid with all necessary manifolding so the isobutane or isopentane containers could be connected to the suction hose and pumped into the system. When filling isobutane, the return vapor hose also was connected to the cylinder to prevent creation of a vacuum in the cylinder.

After loading several cylinders, it became apparent the isobutane would expand to a gas and spin the meter giving false readings. A beam scale was then procured and all hydrocarbon fluid additions or withdrawals from the system were made by weight.

Emergency Shutdown System

The purpose of the emergency shutdown system was to terminate the supply of hot brine and provide the maximum cooling sink to the hydrocarbon loop in the event of equipment malfunction or operator decision. This system is shown schematically in Figure 6-9. The following signals would actuate the Emergency Shutdown System:

1. Interruption of power to the cooling water circulation pump.
2. General loss of electrical power.
3. High heat exchanger hydrocarbon outlet pressure (greater than 650 psig).
4. Low cooling water flow (less than 100 gpm).
5. High cooling water supply temperature (more than 120°F).
6. Low fluid level in the hydrocarbon accumulator (less than 10%).

Figure 6-8 Fire System Spray Layout

- PA-201 PRESSURE ALARM
OPENS ON HIGH PRESSURE
ISOBUTANE
- FA-301 FLOW ALARM
OPENS ON NO COOLING
WATER FLOW
- TA-301 TEMPERATURE ALARM
OPENS ON INCREASING TEMP.
OF COOLING WATER
- LLSD LOW LEVEL SHUT DOWN
OPENS ON LOW LEVEL
IN ACCUMULATOR

△				SCALE	P.S.E.M.T.	
△				DRAWN R.B.	ESD SYSTEM SCHEMATIC	
△				CND	HEBER, CALIFORNIA	
△				APP'D	LBL/EPRI	
REV	DATE	DESCRIPTION	CHK'D	DATE 9-26-78	JOB NO. 778	
COLLEY engineers & constructors, inc.						

Figure 6-9 Emergency Shutdown
System Schematic

Disruption of any of these signals actuated solenoid valves on the air supply to two ball valves, one located at the brine inlet and the other located at the brine exit the brine to hydrocarbon heat exchangers. The actuation of these solenoid valves releases the air pressure holding these valves open. These two ball valves then close terminating the flow of brine through these heat exchangers. The air system also bleeds down. This decrease in air pressure then allows the flow and pressure control valves in the hydrocarbon and cooling water loops to open, providing for maximum circulation of both hydrocarbon and cooling water within their respective loops.

PROCESS SYSTEM DESCRIPTION

The Power System Equipment Module Test Process Systems were designed by Colley Engineers and Constructors, Inc., of Gardena, California. The process systems are shown on the following drawings:

<u>Figure</u>	<u>Drawing Number</u>	<u>Title</u>
6-10	C-778-A-1	Process Flow Diagram
6-11	-A-2	Process & Instrumentation Design
6-12	-L-1	Piping Plan and Elevations
6-13	-L-3	Piping Elevations
6-14	-L-4	Piping Sections & Details
6-15	-L-5	Piping Flexibility Details
6-16	-N-1	Piping Insulation Details
6-17	-N-3	Piping Insulation
6-18	-Q-1	Foundation Details

Brine System

The brine system was designed to deliver approximately 185 gpm (at 5.5 lb/ft³) of geothermal brine to the primary heat exchanger train at 360°F and return approximately 175 gpm (at 6.25 lb/ft³) at 150°F to Magma Electric Company for disposal. This system was designed for a 300# service rating and included basically 3 inch schedule 40 pipe (A53 grade B or equivalent steel).

The construction was welded in sizes 2" and larger. Flanges were 300# Raised Face and were used for all flanged connections 2" and larger. Standard 1/16" asbestos gaskets were used without any problems.

Figure 6-10 Process Flow Diagram

Figure 6-11 Process and Instrumentation Diagram

Figure 6-12 Piping Plan and Elevations

1 ISSUED FOR APPROVAL EAST MESA		3-79
REV	DESCRIPTION	DATE
<p align="center">P. S. E. M. T. PIPING ELEVATIONS</p>		
EAST MESA, CALIF.		LBL / EPRI
<p align="center">CEC</p>		
COLLEY engineers & constructors, inc.		

Figure 6-13 Piping Elevations

Figure 6-14 Piping Sections and Details

Figure 6-18 Foundation Details

Expansion in the piping was calculated and standard configurations were used to keep pipe stresses within allowable limits. The six primary heat exchangers were constructed with expansion joints in the shell to compensate for the difference in expansion between the exchanger shell and tubes during upset operation modes.

In order to provide isolation, two air operated valves were located at the inlet to and exit from the primary heat exchanger train. These valves close upon activation of the emergency shutdown system or upon a loss of air.

The brine flow was controlled by a Fisher flow control valve (FC101), Series 3 inch 667 ED, located downstream from the brine exit valve. The flow control signal for this valve originated at an orifice plate, whose differential pressure range was 100 inches of water. This was converted by a differential pressure cell with a 3 to 15 psi output signal to a Taylor flow controller whose signal modulated the valve.

A pressure relief valve (PSV 6) was located at the brine inlet to the primary heat exchangers. This valve was set to relieve at 412 psig and served to protect the brine side of the primary heat exchange train in the unlikely event of a shell-side to tube-side leak. The release point for this relief valve was located at ground level.

Each piping configuration between exchangers contained a vent, drain, pressure sensing point and a thermowell to accomodate the desired instrumentation.

Hydrocarbon System

The hydrocarbon system was designed to circulate approximately 95,150 lb/hr of isobutane. This system was designed for 635 psig operating pressure.

This system was constructed mostly of 6" and 4" schedule 40 pipe with all joints welded. Weld end valves were used. All connections to the heat exchangers, desuperheater, condenser, accumulator, and subcooler were welded. All small piping was either steel screwed and backwelded or socketwelded.

All flanged connections were 600# raised face with the exception of pump, the hydrocarbon circulating flanges of which were 300# raised face.

Spiral wound asbestos "Flexatallic" gaskets were used and worked very satisfactorily.

Thermal expansion was accommodated in the piping by use of standard piping configurations. The primary heat exchangers were mounted on shoes allowing expansion in two, orthogonal horizontal directions.

The vertical expansion in the hydrocarbon condensing train was accommodated by anchoring the condenser and supporting the desuperheater above on springs. When the system was cold the spring supports were compressed. As the system expanded vertically above the condenser, the preloaded springs relaxed while maintaining support of the desuperheater and piping weight. The accumulator and subcooler were similarly mounted below the condenser on preloaded spring supports. During operation and at elevated temperatures, the downward expansion increased the springs loading maintaining the desired support.

Hydrocarbon flow control was accomplished by use of an orifice plate with a differential pressure cell. The 3 to 15 psi dp cell outlet signal was received by a Taylor flow controller which modulated a bypass control valve allowing a bypass stream to return to the accumulator. This method of control, which is different than the normal method of controlling the flow of a main stream, allowed the flow through the hydrocarbon circulation pump to remain constant. This eliminated the possibility of cavitation at the pump suction during conditions of low test flows.

In normal operation, the majority of the fluid entered the primary heat exchanger train. Upon leaving the primary heat exchanger train, at about 300°F and 585 psig, the fluid was throttled to a pressure of about 235 psig. After this throttling, the fluid entered the desuperheater and exited therefrom saturated at about 200°F. A condenser condensed the fluid at 200°F, which condensed fluid flowed to the accumulator and thence to the subcooler and the hydrocarbon circulating pump.

The primary heat exchanger pressure was controlled by a throttle valve (PV201) the controller for which sensed its upstream pressure. A system of rupture discs was installed around this throttle valve to assure continued access of the primary (hot) side to the cooling train. These discs were set to rupture at 635 and 685 psig respectively.

Hydrocarbon fluid make-up to the hydrocarbon system flowed into the accumulator; fluid let down from the system came from the hydrocarbon feed pump discharge.

Vents, thermowells, pressure connections, cleaning and sample connections were included in the hydrocarbon system at the inlet and outlet of each exchanger, on each side of the throttle valve, and where otherwise required. These various locations can be found on Figure 6-11.

Cooling Water System

The cooling water system was designed to provide a low pressure supply of cooling water to the desuperheater, condenser, and the subcooler. The heated water was cooled by evaporation in a wet, two-cell cooling tower.

The cooling water circulation pump took suction from the cooling tower basin and discharged to a manifold supplying water to the cold water inlets of each of the hydrocarbon coolers.

The hot water discharge from each of the coolers was manifolded to return to the cooling towers. A hot water bypass directly to the cooling tower basin was provided to maintain the cooling water supply at an even temperature of about 90°F.

Makeup water was supplied to the cooling water loop at the cooling tower basin. Blowdown from the cooling water loop occurred at the cooling water circulation pump discharge.

The water flow through the subcooler was controlled at its water exit by a control valve whose controller sensed the subcooler hydrocarbon exit temperature. Likewise, the water flow through the desuperheater was controlled at its water outlet by a control valve whose controller sensed the desuperheater hydrocarbon fluid exit temperature. The cold water flow to the condenser recirculation system was controlled by a valve regulating the hot water flow from the condenser recirculation system. This valve's controller was governed by the hydrocarbon condenser pressure.

The condenser recirculation system shown schematically on Figure 6-11 was used to enable specific cooling water temperature ranges and tube velocities to be obtained independent of the operation of the cooling water system. This

recirculation system consisted of a recirculation pump taking suction from the condenser hot water outlet and returning it to the condenser cold water inlet. Additional provision was made to operate the condenser in a single or double pass water (tube) side geometry.

The cooling water supplies to each exchanger were equipped with a turbine meter for measuring the cooling water flow rate. The total water stream from the main circulating pump was similarly equipped.

EQUIPMENT

The major items of equipment for the Power System Equipment Module Test are discussed below. A picture of the fabricated or installed item is provided together with its specification.

Brine to Hydrocarbon Heat Exchangers

The brine to hydrocarbon heat exchangers are six NEN type units as shown in Figure 6-19. Units B1 through B5 are identical except for their shell side connections being either "right" or "left" hand. Unit B6 differed in its internal baffling to allow a lower pressure drop than would occur with the baffling as installed in units B1-B5. The as built specifications for these units are shown below. These units were designed and fabricated by ThermXchanger of Oakland, California.

BRINE TO HYDROCARBON HEAT EXCHANGERS SPECIFICATION

Code: ASME VIII, Div 1

Surface area: 292.17 ft²/shell, 6 shells

Fluid Flow Rates: 95,150 lbm/hr isobutane

86,260 lbm/hr geothermal brine

Design Pressure: 670 psia shell side, 565 psia tube side

Test Pressure: 1005 psia shell side, 850 psia tube side

Design Temperature: 360°F shell and tube sides

Note: an expansion joint on the shell side was provided
to allow for a 360°F tube side, 70°F shell side
condition.

Tubes: 62 carbon steel, 3/4 inch O.D. BWG 16

24 foot long, Pitch: 15/16 upright 60° triangle

Shell I.D.: 8-3/4 inches

Baffles: units B1-B5, single segmental, 12 inch spacing, 45% cut

unit B6, double segmental, 12 inch spacing, 40% cut

Shell side connections: 2,6 inch Sch 40 pipe nozzles

Channel side connections: 2,3 inch Sch 40 pipe nozzles

Figure 6-19 Brine to Hydrocarbon Heat Exchangers

Desuperheater

The desuperheater is an NJN configuration heat exchanger shown in Figure 6-20. Its as fabricated specification is shown as below. This unit was fabricated by ThermXchanger, Oakland, California.

DESUPERHEATER SPECIFICATION

Code: ASME VIII, Div. 1

Surface area: 287.38 ft²; Duty: 1,971,000 Btu/hr

Design Flow Rates: 95,150 lbm/hr isobutane; 181,000 lbm/hr water

Temperature in/out: 232°F/199°F 90°F/120°F

Operating pressure: 250 psia 50 psia

Design pressure: 670 psia 50 psia

Test pressure: 1,005 psia 75 psia

Design temperature: 360°F 150°F

Tubes: 61 carbon steel, 3/4 inch O.D. BWG 16; 24 ft. length; Pitch: 1 inch
upright 60° triangle

Shell O.D.: 10-3/4 inches

Baffles: 13 inch spacing, double segmented with a 40% cut

Connections: Shellside; 2, 10 inch inlets, 1, 10 inch outlet in center
Tube side; 2, 6" 150# weldneck flanges.

Figure 6-20 Desuperheater, Condenser, Accumulator, and Subcooler

Condenser

The condenser is a NHN configuration heat exchanger and is shown on Figure 6-20. Its as fabricated specification is shown below. This unit was fabricated by Wiegmann and Rose International Corp., Richmond, California.

CONDENSER SPECIFICATION

Code: ASME VIII

Surface area: 1,278 ft²

Design Flow Rates: 95,150 lbm/hr isobutane; 301,250 lbm/hr water

Temperature in/out:	199°F/198°F;	90°F/120°F
Operating Pressure:	245 psig	35 psig
Number of passes:	1	1 or 2
Design Pressure:	655 psig	35 psig
Test Pressure:	983 psig	53 psig
Design Temperature:	360°F	120°F

Tubes: 332 carbon steel, 3/4" O.D. BWG 14, 20 ft. length

Pitch: 15/16 inch upright 60° triangle

Shell I.D.: 22 inches

No baffling, cross-flow arrangement

Connections: 2, 8 inch pipe nozzle inlets; 1, 8 inch 150" RF flange inlet
2, 8 inch pipe nozzle outlets; 2, 8 inch 150" RF flange outlets

Accumulator

The accumulator is shown on Figure 6-20; its specification shown below. This vessel was manufactured by National Tank and Mfg. Company, Los Angeles, California.

ACCUMULATOR SPECIFICATION

Code: ASME VIII

Design Pressure: 650 psig

Test Pressure: 985 psig

Design Temperature: 198°F

Shell Diameter: 16 inches O.D.

Shell Length: 20 ft.

Shell Material: Carbon Steel

Connections: 2, 8 inch pipe nozzle inlets
1, 6 inch pipe nozzle outlet
3, 2 inch 6000# couplings
2, 1 inch 6000# couplings
1, 3/4 inch 6000# coupling
2, 1/2 inch 6000# couplings

Subcooler

The subcooler is a NEN configuration heat exchanger and is shown on Figure 6-20. Its as fabricated specification is shown below.

SUBCOOLER SPECIFICATION

Code: ASME VIII, Div. 1

Surface area: 862.15 ft²

Design Flow Rate: 95,150 lbm/hr isobutane, 176,400 lbm/hr water

Temperature in/out:	198°F/115°F	90°F/120°F
Operating pressure:	243 psia	50 psia
Design pressure:	670 psia	50 psia
Test pressure:	1,005 psia	75 psia
Design temperature:	360°F	720°F

Tubes: 183 carbon steel, 3/4 inch O.D. BWG 16; 24 ft. long

Pitch: 15/16 inch upright 60° triangle

Shell O.D.: 16-3/4 inches

Baffles: single segmental, 12 inch spacing, 44% cut

Connections: shell side, 2, 6 inch pipe nozzles
channel side, 2, 6 inch 150# R.F. flanges

Throttle Valve

The throttle valve is shown on Figure 6-21. Its specification is shown below.
The Throttle Valve was manufactured by the Fisher Control Company.

THROTTLE VALVE SPECIFICATION

Type: Model 3 inch 657 ED

Fluid: isobutane

Design Flow: 91,150 lbm/hr

Upstream Pressure: 600 psi, range 550 to 650 psia

Upstream Temperature: 300°F

Downstream Pressure: 250 psi

Downstream Temperature: 232°F

Valve Body Material: steel

Valve Body Rating: 600# RF ASA

Valve Trim: Whisper

Valve Trim Material: 416 ss

Controller: Taylor Instrument Company 442-RF-1238 Pneumatic Indicating Controller with Proportional Band and Rate Reset, Automatic to manual switch and internal set point. 0-15 psig air output, 30 psig air supply.

Figure 6-21 Throttle Valve

Brine Return Pump

The brine return pump is shown on Figure 6-22. Its specification is shown below. This pump was manufactured by Ingersol Rand, New Jersey.

BRINE RETURN PUMP SPECIFICATION

Type: Chemline

Pump Size: 2 x 3 x 8

Design Flow: 200 gpm

Design Head: 65 psi TDH, sp. gr. 1.1

Seal: Mechanical

Motor Rating: 20 HP @ 3,450 RPM, 480 VAC, 3 phase

Enclosure: TEFC

Figure 6-22 Brine Return Pump

Hydrocarbon Circulating Pump

The hydrocarbon circulating pump is shown on Figure 6-23 and its pump curve shown on Figure 6-24. Its specification is shown below. This pump was manufactured by Afton Pumps, Inc., Houston, Texas.

HYDROCARBON CIRCULATION PUMP SPECIFICATION

Type: Model MPV

Pump size: 3 x 4 x 8

Number of stages: 20

Design Flow: 362 gpm at 120°F

Design Head: 1,791 ft. TDH; sp. gr.: 0.529, NPSHR: 27 ft. of water

Test Pressure: 1,100 psig

Seal: John Crane Type 8B1

Seal Flush Arrangement: API 610 plan 13

Motor Mfg: General Electric

Motor Rating: 125 HP @ 3,600 RPM, 480 VAC, 3 phase

Enclosure: Explosion Proof

Figure 6-23 Hydrocarbon Circulating Pump

Condenser Recirculation Pump

The condenser recirculation pump is shown on Figure 6-27. Its specification is shown below. This pump was manufactured by Ingersoll Rand, New Jersey.

CONDENSER RECIRCULATION PUMP SPECIFICATION

Type: Model HC

Pump Size: 6 x 5 x 10

Design Flow: 600 gpm at 90 to 200°F

Design Head: 58 ft. TDH; sp. gr. 1.0, NPSHR: 9 ft. of water

Maximum Discharge Pressure: 60 psig

Motor Mfg.: Lincoln

Motor Rating: 20 HP @ 1,750 RPM, 460 VAC, 3 phase

Enclosure: TEFC

Figure 6-27 Condenser Recirculation Pump

Cooling Tower

The cooling tower is a two cell, wet tower as shown on Figure 6-28; its specification is shown below. This tower was manufactured by Baltimore Aircoil Company, Madera, California.

COOLING TOWER SPECIFICATION

Type: Baltimore Aircoil Model CFT (2) 2417B

Performance: 1262 gpm of water from 120°F to 90°F at 81°F entering air wet bulb temperature.

Fan motors: 2 at 15 HP each, 1,750 RPM, 460 VAC, 3 phase

Enclosure: T.E.A.O.

Figure 6-28 Cooling Tower (two cells)

PROCESS AND CONTROL INSTRUMENTATION

Control System Design

Simplicity and ability to withstand the environment were two prime considerations governing the design of the control system. A pneumatic system was chosen for proven dependability and accuracy. Operating personnel who understood pneumatic systems were available whereas electronic technicians were not. This consideration provided an additional incentive to use a pneumatic system.

Two flow control devices were required, one on the brine, the other on the hydrocarbon fluid. In both cases, an orifice plate was used as the primary measuring element. Both orifices were designed to yield a differential pressure of 80 inches of water at the design flow. A differential pressure cell with a 0 to 100 inches of water input signal was used to transmit a 3 to 15 psi signal to the flow controller.

The flow control of the brine was standard, however, the hydrocarbon flow control was accomplished by controlling a bypass stream on the hydrocarbon feed pump discharge. This allowed this pump to operate at the same point on its curve at all times.

Temperature control was required on the desuperheater, subcooler, and on the cooling tower bypass. These instruments all used a 3 to 15 psi output signal to the control valve. Each instrument had a bulb inserted in a thermowell at the sensing point. This bulb was connected to the controller by an armored capillary tubing.

Two pressure instruments were involved. Pressure on the heat exchange train outlet was controlled by a throttle valve. This pressure was monitored by a bourdon tube pressure controller with a 3 to 15 psi outlet to the control valve. This throttle valve had a special whisper trim to keep noise levels within comfortable limits.

The other pressure controller sensed the condenser pressure and was to control this pressure by controlling the flow of cooling water through the condenser. This system was not satisfactory since a change in the cooling water flow rate could cause subcooling to occur in the condenser or condensing to occur in the subcooler with no change in the condenser pressure.

Control Requirements

The control requirements for the Binary Module Test are basically the same as those of a commercial plant. Standard, commercial, off-the-shelf, instruments were chosen. Taylor was the supplier of the pneumatic controls, and Fisher, the supplier of the control valves.

The controllers are listed as follows:

Flow Control.

Brine (FC-101). An orifice plate with a 2.20639" bore produced a differential pressure which was transmitted to a Taylor dp cell, model 303TD, with a 0-100 inches of water range input and a 3 to 15 psi output signal. This output was piped to a Taylor model 442 RF 1237 controller with automatic reset and proportional band. The 3 to 15 psi output from this controller controlled a 3 inch, 300# rated globe body Fisher control valve.

Hydrocarbon (FC-201). An orifice plate with a 2.755850 inch bore in 4" - 600# rated flanges produced a differential pressure which was transmitted to a Taylor dp cell, model 303 TD, with a 0-100 inches of water range input and a 3 to 15 psi output signal. This output was transmitted to a Taylor model 442 RF1237 controller with automatic reset and proportional band. The 3 to 15 psi output from this controlled a standard 1", 600# rated globe body Fisher control valve.

Cooling Water (TC-301, 302 and 303). These temperature controllers were each Taylor model 442-RV-1225. With automatic reset and proportional band which supplied a 0-15 psig air output signal to the respective control valve. The input signal to each controller came from a temperature bulb inserted in a thermowell in their respective streams. The control valves were standard 150 pound globe body Fisher control valves of two, three and six inch sizes.

Cooling Water (PC-301). The condenser pressure was controlled by the flow of cooling water through the condenser.

A Taylor model 442-RF-1238 pressure controller with automatic reset and proportional band used a 3 to 15 psi output signal to modulate a 6", 125# rated globe body Fisher control valve. The pressure range of this controller was 200-300 psig.

Process Instrumentation System

In general the philosophy of the process instrumentation system was to enable measurements to be taken of the operational performance independent of a Data Acquisition System. While the data would not be as complete, it would be sufficient.

During the course of operations, complete reliance was placed in the process instrumentation system, except for the glass bulb thermometers, due to inoperability of the sophisticated flow and pressure instrumentation of the Data Acquisition System.

Differential pressure gauges reading in inches of water were placed across the two orifice plates. Since the digital readout meters of the Data Acquisition System were not operating, these differential pressure gauge readings were used to determine flow rates from the calculated curves for each orifice plate.

For temperature back up, dial thermometers were placed at such points that if the ASTM quality mercury thermometers were broken, temperatures could still be measured and recorded.

Pressure measurements were taken directly from pressure gauges mounted in the system. The pressures recorded substantiated the differential pressure gauges when they were functioning and with the failure of many of the differential pressure gauges due to over-range conditions, the operating pressure gauges became the primary source of data.

Pressure Indicators. All pressure gauges were of the bourdon tube type, bottom mounted with 1/4 inch NPT connections. 4 1/2 inch dials were used. All pressure gauges were manufactured by Ashcroft.

Temperature Instruments. Two types of temperature instruments were used. Dial thermometers either back or bottom mounted were used as required. These are bimetallic thermometers were inserted in thermowells to enable instrument change during plant operations. These were primarily used to set test operating conditions.

Temperature data were gathered from ASTM grade, mercury glass bulb thermometers. These were manufactured by the Brooklyn Thermometer Company.

Differential Pressure Gauges. The differential pressure gauges were bellow gauges, manufactured by the Midwest Instrument Company.

Thermowells. Thermowells were all solid barstock, carbon steel, 3/4 inch male thread, 1/2 inch female thread bored to .385 inches. Length was dependent on the pipe diameter in which the thermowells were to be inserted. The thermowells are catalogue items and were procured from Republic Supply Company.

Pressure Relief Valves. Pressure relief valves were of all steel construction. All were 3/4 inch screwed inlet, 1 inch screwed outlet. All relief valves were manufactured by Consolidated, a division of Dresser Industries.

Level Glasses. Level glasses are manufactured with a forged steel body. Pressure ratings are 1270 psi. The level glasses used were manufactured by the Ernst Company.

Temperature Switch. The temperature switches were bimetallic and constructed with explosion proof electrical components. These were Single Pole Double Throw (S.P.D.T.) with a capacity of 5 amps @ 110 Volts, and were manufactured by DeLaval Turbine.

Pressure Switches. The pressure switches were bourdon tubes with explosion proof electrical components. These were also S.P.D.T. with a capacity of 5 amps @ 110 Volts and were manufactured by DeLaval Turbine Company.

Flow Switches. The flow switches were of the paddle type. These units were also S.P.D.T. with an electrical capacity of .17 amps @ 110 Volts. These units were also manufactured by the DeLaval Turbine Company.

Level Switch. The level switch was a snap acting pneumatic switch. Level change beyond the set point changed the signal output from 3 to 15 psi. A pressure switch sensed the pneumatic change and activated an electrical contact. This switch unit was manufactured by Fisher Controls.

Rupture Discs. The rupture discs and rupture disc holders were manufactured by Continental Disc Corporation. The discs were fabricated of stainless steel with a tested thickness to burst should the predetermined pressure difference across the disc of 400 or 450 psi be exceeded.

Section 7

DATA ACQUISITION SYSTEM

DESCRIPTION

The Data Acquisition System consisted of high precision instrumentation at the data station locations on the Binary Module Test brine, hydrocarbon and cooling water loops. Much of this instrumentation failed or was found unreliable during test operations. The data stations used to monitor system and equipment performance are identified on the Project Piping and Instrumentation Drawing, Figure 6-11. Table 7-1 lists the instrumentation used together with its accuracy and range limitations. Three basic types of variables were monitored by this instrumentation; i.e., flow rate, temperature, and pressure. The data read from this instrumentation was manually entered onto log sheets; an example of which is shown as Figure 7-1.

Thermometers. Mercury, glass bulb thermometers provided accurate temperature indications. These thermometers were manufactured by Brooklyn Thermometer Company.

They had a 60°F temperature range capability enabling readings to within $\pm 0.1^\circ\text{F}$. They were installed within thermowells. The thermowells were partially filled with a silicon oil. Additionally, the thermometers were housed within steel sheaths to provide protection against damage.

Pressure Meters. Pressure measurement was provided by two Ashcroft digigauges. One digigauge provided pressure measurement for the brine system; the other monitored selected pressures on the hydrocarbon system. These digigauges are identified as PI108 and PI204 on Figure 6-11. Excess flow check valves were provided in the tubing connecting digigauge PI204 to its sensing points to terminate loss of hydrocarbon fluid in the event of tubing damage or breakage. These two digigauges never furnished data input during test operations.

OPERATOR:

Pentecost
"C" shift

Run No: AS per RS. INSTRUCTING

Hydrocarbon:

Date: 1-14-80 Time Start: 2:15

Day in Week: MON Time Compl: 7:55

[illegible]

DESUPERHEATER, CONDENSER AND SUBCOOLER

	High	Low		High	Low		High	Low		High	Low
TW212	111		T1202 1/2	111		PI202 1/2	215				
TW302	90		T1302	91		PI302	40.5			M302	62
TW305	90.5		T1305	83		PI305	36.5			M305	322
TW307	90		T1307	81		PI307 3/4	34			M307	46
TW306	146		T1306	142		PI306	36				
TW210a	155										
TW304	81.6		T1304	82		PI304	33				
TW209	200		T1209	185		PI209	235			M304	324
TW208	245.8		T1208	220		PI208	240				
TW210b	186.8					PI200	43				
TW308	140		T1308	160		PI308	34				
TW309	155		T1309	158		PI309	1.5				
TW303	164		T1303	155		PI303	42				
TW211	172		T1211	180							
TW301	91		T1301	81		PI301	38			M301	433
			T1210	178		PI211	250				

B-1 646.5

B-2 629.5

B-3 422.5

6-4-609

6.5 593.5

Bk 576

B-6 DISCHARGE 550

Figure 7-1. Data log sheet.

Pressure measurements were made by Ashcroft pressure gauges, part of the process instrumentation system. Since the fluid properties of the brine and cooling water show little pressure sensitivity in the test operating ranges, the accuracy of these gauges was sufficient for Data Acquisition System purposes.

Differential Pressure. Differential Pressure measurements were obtained by Mid-West Instrument WP gauges, as indicated by the symbol DPI on Figure 6-11. Most of these gauges were damaged due to over range conditions and two of these were used to monitor the orifice plate flow readings for both hydrocarbon and brine.

Flow Meters. Two different types of flow meters were to be included in the Data Acquisition System. A Venturi flow meter manufactured by Flow-Dyne Engineering, Inc., was to be used to monitor the flow rate of brine as it exited from the primary heat exchanger train. This venturi meter had a calibrated accuracy of $\pm 0.5\%$ in the flow range of 104 to 244 gpm. Since its digigauge readout instrument was damaged, this venturi meter did not furnish data during test operations. Rather, brine flow measurement was accomplished by monitoring the flow orifice plate installed as part of the brine flow control system.

Turbine Flow Meters were used to monitor the flows in both the hydrocarbon and cooling water loops. These meters were manufactured by Flow Technology, Inc. The hydrocarbon flow meter was located just upstream of the hydrocarbon entrance to the primary heat exchangers, and is identified as M201 on Figure 6-11. It was calibrated to ± 1 gpm in the indicated range. This flow meter developed bearing problems during test operations and subsequent to November 28, 1979 was not monitored for input to the test data. The flow of hydrocarbon fluids was subsequently and successfully monitored by the orifice plate installed as part of the hydrocarbon flow control system.

Five turbine meters were located at various points in the cooling water loop. These meters are identified by the symbols M301, M304, M305, and M307 as shown on Figure 6-11. These also were calibrated to ± 1 gpm in the indicated range and operated satisfactorily during the test operations.

A summary of these flow meters is shown on Table 7-2, and the orifice plate specifications shown on Table 7-3.

Table 7-1
LIST OF DATA ACQUISITION INSTRUMENTATION

Instrument	Range	Accuracy	Instrument	Range	Accuracy
TW 101	As required	$\pm 0.1^{\circ}\text{F}$	TW 201	As required	$\pm 0.1^{\circ}\text{F}$
102	"	"	202	"	"
103	"	"	203	"	"
104	"	"	204	"	"
105	"	"	205	"	"
106	"	"	206	"	"
107	"	"	207	"	"
PI 101	0-300 psig	± 3 psig	PI 201	0-800 psig	± 0.8 psi
107	0-300 psig	± 3 psig	207	0-800 psig	± 0.8 psi
DPI 101*	0-150 in H_2O	± 1.5 in H_2O	DPI 201(1)	0-10 psi	± 0.1 psi
102*	"	"	202(1)	0-15 psi	± 0.15 psi
103(1)	"	"	203(1)	0-15 psi	± 0.15 psi
104(1)	"	"	204(1)	0-15 psi	± 0.15 psi
105(1)	"	"	205(1)	0-30 psi	± 0.3 psi
106(1)	"	"	206(1)	0-30 psi	± 0.3 psi
M 101	orifice plate	$\pm 0.5\%$	M 201	orifice plate	$\pm 0.5\%$
TW 208	as required	$\pm 0.1^{\circ}\text{F}$	TW 307	as required	$\pm 0.1^{\circ}\text{F}$
209	"	"	308	"	"
210 a,b	"	"	M 307	48-132 gpm	± 1 gpm
211	"	"	PI 307	0-60 psi	± 0.6 psi
212	"	"	308	0-60 psi	± 0.6 psi
PI 208	0-600 psi	± 0.6 psi	TW 306	as required	$\pm 0.1^{\circ}\text{F}$
209	"	"	305	"	"
211	"	"	M 304	268-493 gpm	± 1 gpm
212	"	"	M 305	0-600 gpm	± 1 gpm
			PI 306	0-60 psi	± 0.6 psi
			305	"	"
			304	0-60 psi	± 0.6 psi
			TW 303	as required	$\pm 0.1^{\circ}\text{F}$
			TW 302	"	"
			M 302	11-352 gpm	± 1 gpm
			PI 303	0-60 psi	± 0.6 psi
			302	0-60 psi	± 0.6 psi
			TW 301	as required	$\pm 0.1^{\circ}\text{F}$
			M 301	328-1196 gpm	± 1 gpm
			PI 301	0-60 psi	± 0.6 psi
			TW 309	as required	$\pm 0.1^{\circ}\text{F}$

* Relocated to measure orifice plate pressure differences

(1) Movements stretched during test operations, instruments rendered inoperative

Table 7-2
FLOW METER SUMMARY

SERVICE			PIPELINE			METERS						
Fluid	Pipeline	Data Station No.	Design Flow Range (GPM)	Nominal Line Size & Sched. (in./sch)	Flange Rating (PSI)	Type	Serial No.	Nominal Pipe Size (inch)	Flange Size & Rating (in./psi)	Flow Range (GPM)	Calibr. Accuracy (%)	Model No.
Brine	From Prim. Ht. Exchrs.	M101	42-193	3/40	300	Venturi ¹	14891	3	3/300	104-244	±0.5	VP030470-CF
Brine	From Prim. Ht. Exchrs.	M101		3/40	300	Orifice Plate ³		3	3/300	0-200	±0.5	-
H. Carb.	To Prim. Ht. Exchrs.	M201	76-358	4/80	300	Turbine ²	40082 & 40083	2.5	2.5/300	40-400	±0.1	FT40
H. Carb.	To Prim. Ht. Exchrs.	M201		4/80	600	Orifice Plate ³		4	4/600	0-400	±0.5	-
C. Water	To Desuperheater	M307	48-132	6/40	150	"	240424 & 240425	1.5	1.5/150	20-200	±0.1	FT24
"	To Condenser (cooling twr. water)	M304	268-493	6/40	150	"	48207 & 48208	3	3/150	60-600	±0.1	FT48
"	To Condenser (recirculation)	M305	0-600	6/40	150	"	64076	4	4/150	100-1100	±0.1	FT64
"	To Subcooler	M302	11-352	4/40	150	"	40084 & 40085	2.5	2.5/150	40-400	±0.1	FT40
"	From Cooling Twr.	M301	328-1196	6/40	150	"	64077	4	4/150	100-1100	±0.1	FT64

(1) Manufactured by Flow-Dyne Engineering, Inc., Fort Worth, Texas

(2) Manufactured by Flow Technology, Inc., Phoenix, Arizona

(3) Manufactured by Daniel Industries, Compton, California

Table 7-3
ORIFICE PLATE SPECIFICATIONS

<u>Fluid Metered</u>	<u>FC101 Geothermal Brine</u>	<u>FC201 Isobutane</u>
Maximum Flow at Operating Temperature, gpm	177	363
Operating Temperature, °F	165	120
Operating Pressure, psig	158	635
Operating Viscosity, cp	0.375	0.082
Specific Gravity at operating temperature	0.975	0.530
Pipe Diameter, inches	2.20639	4.026
Orifice Size, inches	3.068	2.75850
Pressure Tap Location	Flange	Flange