STUDY AND TESTING OF DIRECT CONTACT HEAT EXCHANGERS FOR GEOTHERMAL BRINES

FINAL REPORT

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TEST UNIT - BOILER AND PREHEATER

TEST UNIT - OVERALL VIEW

TEST UNIT - BOILER AND PREHEATER

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STUDY AND TESTING OF DIRECT CONTACT
HEAT EXCHANGERS FOR GEOTHERMAL BRINES

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ABSTRACT

The object of the work reported herein was to assess the technical and economic feasibility of preheating and evaporating a secondary fluid via direct contact with hot geothermal brine. The work covered a period of 12 months and included the design, construction and testing of a unit which heats and vaporizes 10 gpm of isobutane by direct contact with 325°F brine. The analytical and experimental efforts explored design and economic characteristics, including anticipated problem areas such as working fluid loss in the brine, production of a stable dispersion of the working fluid in brine, fluids separation, axial mixing and carry-over of water vapor with the working fluid.

Preliminary studies resulted in the selection of isobutane as the working fluid for tests primarily because of the favorable amount of net work produced per pound of geothermal brine and the low amount and cost of working fluid lost in the heat exchange process. The Elgin Spray Tower concept was selected for the preheater and boiler.

The test apparatus includes a separate boiler and a separate preheater, each 6" diameter by 6' high. Brine enters the top of each vessel and leaves the bottom. Isobutane enters the bottom of the preheater through a distributor plate to produce .15 inch diameter drops.

The experimental unit operated with no major problems and demonstrated its hydraulic and thermal capabilities. Volumetric heat transfer coefficients obtained ranged up to 4000 BTU/hr - °F - Ft³. Boiling heat transfer coefficients of as high as 17,000 BTU/hr - °F - Ft³ were obtained with a design value of 10,000 BTU/hr - °F - Ft³. Amount of isobutane in a 21% NaCl solution leaving the preheater was less than 40 ppm.

It is recommended that testing of the experimental unit be continued with different heights and dispersion systems; and that it be automated for continuous operation on geothermal brine at East Mesa. Then, a larger "pilot plant" should be designed incorporating the optimum design features. Concurrently with this, detailed economic optimization studies of the entire power cycle utilizing direct contact heat exchangers should be accomplished.

SECTION 1

SUMMARY AND INTRODUCTION

1.1 SUMMARY

The purpose of this development program is to assess the technical and economic feasibility of utilizing direct contact heat exchangers in conjunction with a binary cycle to extract energy from highly saline geothermal brines. The recently completed 12 month program included the design, construction and testing of an experimental system which continuously heats and vaporizes 10 gpm of isobutane by direct contact heat exchange with 325°F simulated brine. The analytical and experimental efforts explored a broad range of design and economic characteristics of direct contact heat exchangers including anticipated problem areas such as working fluid loss in the brine, production of a stable dispersion of isobutane in brine, fluids separation, axial mixing in the preheater, and carry-over of water vapor with the isobutane.

Preliminary Investigations

Preliminary studies conducted prior to the experimental activities established a simple thermodynamic model for analyzing the relative merits of various working fluids and heat exchanger concepts. These studies resulted in the selection of isobutane as the working fluid primarily because of the favorable amount of net work produced per pound of hot geothermal brine and the low amount and cost of working fluid lost in the direct contact heat exchange process. Evaluation of various concepts for the direct contact preheater and boiler resulted in the selection of an Elgin Spray Tower for the preheater. For the boiler, a design was developed that injects the isobutane into the bottom of a vertical cylindrical vessel through tangential nozzles. The isobutane begins vaporization immediately and the droplets and bubbles spiral upward until they disengage from the free surface of the brine near the mid-point of the vessel. Hot brine is sprayed into the top of the vessel and exits at the bottom.

Test Apparatus

The experimental unit consisted of the direct contact preheater and boiler, conventional surface condenser, hot well, brine circulating pump, conventional steam heater (to reheat the recirculated brine), and an isobutane circulating pump. The fluids were circulated counter-currently through the two direct contact heat exchangers in a continuous fashion thereby simulating a binary cycle, with the exception of a turbine. At design conditions, the brine enters the boiler at 325°F and is cooled by the vaporizing isobutane to a temperature of 225°F. From the boiler, the brine flows to the top of the preheater. The isobutane enters the boiler at 220°F (saturated) and the vapor leaves with between 0 and 25°F of superheat. The isobutane vapor is then throttled and condensed in the standard surface heat exchanger. The liquid isobutane is pumped to the bottom of the preheater where it is dispersed in the brine as small droplets (.15 inch diam.) and rises through the brine, leaving the top of the preheater at 220°F.

Operational Tests and Results

The experimental unit operated with no major problems and demonstrated its hydraulic and thermal capabilities. The volumetric heat transfer coefficients obtained in the preheater varied predominantly with the volumetric isobutane holdup and ranged up to 4000 BTU/Hr-°F-Ft.³ At the design hourly rate capacity ratio, the "size" of the preheater was measured as high as 3 heat transfer units, which is approximately defined by ΔT isobutane/ ΔT log mean. The boiling heat transfer coefficients were measured to be as high as 17,000 BTU/Hr-°F-Ft.³ The boiler

also demonstrated its capacity to preheat isobutane rather efficiently. Sub-cooled isobutane was preheated at UA values as high as 5000 BTU/HR-°F.

For test operations conducted with fresh water, the isobutane content of the water leaving the preheater was determined to be essentially at the saturation level. For normal test conditions these equilibrium levels ranged between 240 and 300 ppm. The concentration of isobutane in 21% saline solutions was too low to be measured by the small stripping column, and spectographic analysis was not successful because of the sampling problems associated with the high temperature and pressure of the brine. The saturated isobutane content for the highly saline brines was calculated to be less than 40 ppm, however. The water vapor content in the isobutane vapor leaving the boiler was measured at 1 to 2% of the isobutane flow, and only 10% of that amount is expected to condense as the isobutane flows through the turbine.

Cost Estimate

A conceptual design and cost estimate was prepared for a direct contact heat exchange system sized for a 50 MW power plant. The design of the heat exchangers was based on the configuration and performance of the experimental heat exchangers, and is quite conservative. The conceptual design incorporates ten parallel heat exchangers each one comprising two preheater stages and a boiler. Each of the ten units is a single 12 foot diameter tower, having the preheater stages and boiler arranged vertically. The heat exchanger system also requires a brine booster pump to increase the brine pressure up to the total vapor pressure of the brine and isobutane in the boiler. The total installed cost of the heat exchangers, pumps, and an additional isobutane storage tank was estimated to be \$6,268,430, or \$125/KW.

The operating expense due to isobutane losses was based on a loss rate of 40 ppm in the effluent brine and amounts to \$273,493 per year. At an average annual fixed charge of 13%, the total annual costs for the system are \$1,088,400. An optimistic appraisal of possible technical improvements to the conceptual design indicate that these costs would be reduced in the future by as much as 33%.

Conclusions

The analytical and experimental program just completed has probed several aspects of direct contact heat exchange between geothermal brine and isobutane. The overall conclusion is that no major obstacles to the development of a commercial system were uncovered. Furthermore, the prospects for the early development of a commercial system appear favorable, particularly for highly saline brines. However, a cautious approach should be taken in the development of the system until the economic comparisons with conventional or other developmental heat exchange systems can be made more accurately. The improved accuracy in comparisons will require that additional trials with all types of heat exchangers be conducted with the more perverse types of geothermal brines in order to document their performance and operating costs under these conditions.

The conclusions reached as a result of the analytical and experimental program just completed are set forth in detail with discussion in Section 5. The overall conclusion is that there are no major technical obstacles to the development of a commercial direct contact heat exchange system.

A summary of our main conclusions are as follows:

- (a) Continuous and stable operation of the direct contact preheater and boiler can be easily accomplished.
- (b) The preheater of the test unit (6" diameter x 6' high) transferred heat from a 21% brine solution to isobutane at a rate of 3500 BTU/Hr-°F-Ft³. The heat exchange effectiveness was 80% which, for the design hourly heat capacity ratio, corresponds to 2.7 heat transfer units.
- (c) The boiler of the test unit transferred heat at rates of up to 17,000 BTU/Hr-°F-Ft³ based on an effective volume of .4 ft³. The boiler also could preheat the isobutane quite efficiently.
- (d) The isobutane droplet size and distribution during test operation was uniform. They did not exhibit any premature coalescence, either with each other or with the sides of the vessel. Droplet size average .15" diameter.
- (e) Isobutane as the working fluid is easy to work with and observations disclosed no undesirable characteristics such as sludge formation, foaming and chemical reactions within the heat exchangers.
- (f) A more comprehensive investigation of the amount of water vapor leaving the boiler with the isobutane vapor is required. Measured amounts during test operation ranged from 0.2 to 1.9% of the isobutane vapor. The effect of this water vapor on the system design and performance should be investigated.
- (g) The method of measuring the dissolved isobutane in the brine effluent from the experimental unit was unsatisfactory. A different method must be developed to determine low concentrations (20-40 ppm) in high salinity brines.
- (h) Capital costs of a heat exchanger system using direct contact units are comparable to using surface exchangers with all carbon steel material. This comparison does not take into account anticipated improvements in performance of the direct contact heat exchangers.
- (i) The effects of backmixing and resultant direct contact heat exchanger performance due to variations in tower diameter and height is unclear. Additional heat transfer data of a more localized nature will have to be determined to relate these variables.
- (j) There are several means for improving the performance of preheaters. The heat transfer units may be increased 20% by increasing the hold-up of isobutane. The throughput of fluids could be increased by 20%.
- (k) The capital and operating costs of a full-sized heat exchange system could be reduced based on accomplishing the improvement as concluded by (j) above.

Recommendations

The studies and testing, as reported herein, have shown that there are numerous possibilities for improving the performance and economics of the system. Also, there are certain design criteria which should be verified so that different size units can be designed. The recommended development of work is aimed at the following three objectives:

- (a) Demonstrate the ability of the test system to operate reliably on actual geothermal brine in a field application.
- (b) Improve the performance and reduce the costs of the system.
- (c) Determine better design criteria for varying size and parameters, and to optimize the overall system.

The specific tasks to accomplish these objectives are summarized below:

- (a) Modify existing test unit so that it can be operated continuously 24 hours per day with only "supervisory" attendance.
- (b) Install a duplicate preheater in series with the existing preheater to enable the system to simulate a complete binary cycle and operate over the full brine temperature range.
- (c) Improve the method of measuring the water vapor carry-over from the boiler so that long term data can be obtained during field testing.
- (d) Add to the existing test apparatus a stripping column, or degassifier, that is sized for the entire brine discharge from the test unit.
- (e) Operate the existing preheater with increased hold-up and achieve a "close-packed" mode of hold-up.
- (f) Test the variation in preheater heat exchanger effectiveness by varying the height of this unit.
- (g) Test the system with the preheater and boiler combined into one integral unit, one above the other.
- (h) Test the boiler with different nozzles and distributors. At least two different types of atomizing nozzles should be tested.
- (i) Modify the boiler so that the brine is introduced and distributed below the free surface level of the brine.
- (j) Test mixtures of working fluids in the experimental test unit. The existing equipment is well suited for testing mixtures of binary fluids and no major modifications would be necessary to do this.
- (k) Install additional temperature instrumentation in the preheater sufficient to make a detailed temperature profile of the isobutane and brine.
- (l) Apply the additional test and development data, as determined by the above recommendations, to the refinement of proposed commercial systems. In These, the heat exchanger system should not be considered alone but the overall reservoir/power plant/heat rejection and reinjection system should be evaluated.

- (m) Perform further detailed optimization studies of overall power cycle systems considering boiler pressure, plant efficiency, heat exchange vessel wall thickness, number of vessels, vessel diameter, other working fluids, effect of water vapor, and overall operating costs.
- (n) Make detailed and thorough studies on the technical and economic aspects of heat rejection from the overall power plant cycle. The size and costs of condensers and cooling towers should be determined. Direct contact condensation of isobutane on water followed by separation and cooling of the water in a cooling tower should be included in a heat rejection study.
- (o) Design a small, complete power plant about 20 times larger (100 gpm brine) than the present test unit. This pilot plant should include the complete binary power cycle with direct contact heat exchangers. A turbine generator with a net electric output of 100 KW minimum should be included.

1.2 INTRODUCTION

This research was conducted in response to the Energy Research and Development Administration's desire to develop the Direct Contact Heat Exchanger/Binary Cycle process for the extraction of energy from geothermal brines.

Recent studies of means to economically extract energy from liquid dominated geothermal fields have focused on utilizing a secondary fluid in a closed Rankine cycle for power generation. Much theoretical work has been done involving different cycle variations and the many potential advantages of such a cycle for geothermal power generation have been discussed and documented. The major problem associated with the binary cycle is the need for heat exchangers to preheat and evaporate the working fluid with heat extracted from the geothermal brine. Unless carbon steel can be used exclusively, surface type heat exchangers for this duty will be expensive due to low heat transfer rates, low ΔT 's, and the costly materials required. Scale formation and corrosion with many of the geothermal brines will cause severe operating and maintenance problems for conventional tubular heat exchangers.

A possible means of reducing the cost of the heat exchangers and solving problems related to scaling and corrosion is the use of direct contact heat exchangers.

The overall purpose of the recent study and testing work reported herein was to assess the technical and economic feasibility of utilizing direct contact heat exchangers for geothermal applications, and to provide an impartial analysis identifying any engineering problems requiring particular attention in future R&D activities.

To accomplish the overall purpose, the following main objectives were specified:

- .Selection of the most promising working fluid and development of methodology for fluid comparison studies.
- .Establishment of basic direct contact heat exchanger concepts and preliminary design procedures.
- .Design, fabrication and assembly of test apparatus.
- .Operate test apparatus to provide design information for the use of Direct Contact heat exchangers applied to geothermal brine.
- .Development of conceptual design of full size system and estimate its cost.
- .Recommend future R&D activities.

SECTION 2 PRELIMINARY STUDIES

2.1

DESIGN PHILOSOPHY

The preliminary investigations were aimed at creating a complete picture of the binary power cycle and the brine to isobutane heat exchange process. Analysis of the entire system enabled us to select the best heat exchange concept and to determine its impact on the entire binary cycle.

The process of establishing a basis for the subsequent conceptual design and test program required making several design choices. To some extent these decisions were intuitive choices based on our experience with previous direct heat exchange investigations, the design of desalination plants and geothermal mineral recovery systems. The key decisions centered on the following topics:

- a. Working Fluid
- b. Thermodynamic Cycle: Subcritical (saturated and superheated) or supercritical
- c. Brine Type
- d. Type and Configuration of Direct Contact Heat Exchanger
- e. Materials of Construction

The detailed description of these decision making activities constitutes the bulk of this Section.

2.2

SELECTION OF WORKING FLUID

A working fluid was sought which would be suitable for a commercially sized binary power cycle utilizing direct contact heat exchangers. The selected fluid would also be used in the experimental system to be built during this development program. As a result of a comparative study, Isobutane was selected as the most promising fluid.

The optimal working fluid is one that, when utilized in the direct contact binary cycle, minimizes the busbar cost of power. To achieve this, the sum of the following costs must be minimized:

1. Brine supply (pounds brine per KW out)
2. Working fluid losses (solubility in brine)
3. Heat Exchanger(s) (operating pressure, temperature, and H.T. rate)
4. Pumping Power (flow quantity and heads)
5. Pumps - (flow quantities and heads)

To determine the minimum for items 3-5, above, the type of heat exchangers and system design must be known. These are quite different between a closed loop binary cycle and open loop (direct contact) binary cycle. Initially the factors effecting these items and their relationship to other costs were not definitely established. Therefore, it was decided to select a working fluid that would minimize the costs of items 1 and 2 only. Subsequent analysis can be made incorporating items 3-5 once the working concept and cost factors are established.

The first criteria required calculating the effectiveness of each fluid at various boiler temperatures. This procedure has been duplicated by numerous other investigators and there appears to be agreement that, from a thermodynamic standpoint, isobutane maximizes the gross work out with a wide range of brine temperatures.

The second requirement, minimum cost of working fluid loss, was also met by isobutane. In our preliminary analysis, the cost of isobutane loss was estimated by considering the brine flow rate per KWH, the equilibrium solubility of isobutane in brine, and the unit cost of isobutane. Based on a system effectiveness of 15 BTU net work per pound of brine, there are 227.5 lbs/hr. of brine required per KW electricity. The solubility of isobutane at equilibrium with a 21% NaCl brine at 137°F and 100 psia partial pressure was determined from published data³³ and is approximately 30 ppm. The unit cost of isobutane is \$.07/lb. The cost of the isobutane loss due to solubility in terms of \$/KWH is then $227.5 \times 30 \times 10^{-6} \times .07 = \$.0005/\text{KWH}$.

The solubility of the isobutane in fresh water at the same brine pressure and temperature (100 psia, 137°F) is 258 ppm, or nearly 10 times higher than in 21% NaCl brine. The "salting out" effect of the NaCl in the brine reduces the isobutane solubility tenfold, but the exact solubility in real geothermal brines has not been determined. The presence of salts other than NaCl in the brine may cause the solubility to vary slightly.

In addition to solubility losses, there remains the possibility that amounts of isobutane will be carried out with the rejected brine in the form of droplets or a very fine emulsion. Again, isobutane is excellent in this regard because of the ease with which dispersions can be broken. Isobutane is very easily separated from water due to its very low density.

The cost of commercial isobutane is 7.4¢/lb or about 33¢/gallon in small quantities. We purchased 160 gallons at this price for our test unit from a major oil company located in the southwest. According to the supplier, the commercial grades of isobutane normally contain 97% isobutane, 2% N-butane and the balance methane and propane.

Another desirable characteristic of a working fluid for direct contact heat exchange is the ability to coalesce with itself in a manner that can be predicted and controlled by the designer of the heat exchanger. Isobutane exhibits this characteristic very well. The isobutane droplets will not coalesce against steel in contact with water, but will coalesce against plastics. The ability to control the location of self coalescence permits the use of devices which continuously disperse the fluid into the brine and separate it for removal in one vessel.

Several experimental direct contact heat exchangers tested in connection with desalination processes were plagued by the formation of sludge at the oil-water interface. It is difficult to predict the long term performance of isobutane in this respect and only long term testing can establish the sludge forming potential of isobutane/brine systems. The testing conducted during this program with fresh water and simulated brine did not encounter any sludge problems.

The thermodynamic properties of isobutane used throughout this report were obtained from the tabulated data and curves found in the OSW Handbook.³³

The general scope of work for this program, as originally suggested by ERDA, specified a system which would "vaporize and superheat" the working fluid by contact with a brine between 300°F and 350°F. This original direction implies a subcritical cycle and is one with which we have never disagreed. It is our belief that for the development of geothermal brine utilization a supercritical cycle introduces an unwanted element of uncertainty and is inappropriate for a pioneering effort such as ours. For high effectiveness, an isobutane supercritical cycle requires a relatively high brine temperature and assurance that the temperature of the brine will not decline below a certain minimum over the lifetime of the brine supply. In view of this, and the moderate brine temperature range specified, the super critical cycle was not considered further.

The question of investigating a saturated, as opposed to a superheat, type of cycle was not considered a critical choice in the design of the test unit since the degree of superheat can be varied with minor design or operating changes to the boiler. The conceptual design on which the test unit was based did, however, incorporate provisions for superheating. There are several factors to be considered with regard to the amount of superheat desired for a direct contact cycle. Without regeneration a very small gain in thermal efficiency can be obtained with high superheat because the paths of isentropic expansion from high superheat conditions have a slightly steeper slope with respect to enthalpy. For example, if the turbine inlet and outlet pressures are fixed at 400 psia and 80 psia respectively, the isentropic expansion at 60°F superheat will produce approximately 29% more work per pound of isobutane but will require about 27% more heat input and the theoretical thermal efficiency of the cycle is increased from 15.36 to 15.64%; hardly worth the effort. With a regenerator however, the superheat cycle efficiency could theoretically be increased to 20%, but from a practical standpoint perhaps only to 16 or 17%. Unfortunately, the cycle effectiveness, which is defined as BTU net work per pound of brine, will not increase by using superheat and regeneration unless a hotter brine is available. In this manner, the additional heat input to the superheater can be accomplished without increasing the brine flow and lowering the effectiveness.

The brine temperature seems to be the source of some confusion in comparing thermodynamic cycles. To compare cycle effectiveness (BTU/LB Brine) the cycles being compared must assume the same brine inlet temperature as well as brine specific heat (a function of salinity).

An analysis of the cycle thermodynamics was made for several candidate working fluids at various boiling temperatures in order to select the best working fluid with respect to cycle efficiency. The same method of cycle analysis is used later in the report in discussing test results and the conceptual design. The cycle analysis assumed a brine temperature of 325°F, a solids content of 21% by weight, and the use of direct contact heat exchangers. The analytical expressions used for the cycle studies along with additional assumptions and input to the model are given on the following pages. The state points and terminology are shown on the Temperature-Enthalpy diagram, Figure 2-1, and a list of symbols is given on the following page.

SYMBOLS FOR CYCLE ANALYSIS

C_{pb}	=	Specific heat brine (BTU/lb-°F)
C_{pf}	=	Specific heat working fluid (BTU/lb-°F)
E	=	System effectiveness, (BTU/lb of brine)
Δh_{cw}	=	Enthalpy difference of cooling water (BTU/lb)
Δh_{fcond}	=	Heat rejected to condenser (BTU/lb of working fluid)
$h_1 - h_6$	=	Latent heat of vaporization at T_{sat} (BTU/lb)
Δh_t	=	Isentropic turbine work (BTU/lb)
M_b	=	Brine Flow rate (lb/hr)
M_f	=	Working fluid flow rate (Lb/hr)
M_f/M_b	=	Ratio of working fluid flow rate to brine flow rate (Lb/Lb)
N_t	=	Turbine efficiency
N_p	=	Pump efficiency
P_{cond}	=	Condenser pressure (psia)
P_1	=	Boiler Pressure (psia)
P_{net}	=	Net turbine power (BTU/hr)
T_{sat}	=	Working fluid saturation temperature (°F)
T_{B1}	=	Temperature of inlet brine (°F)
T_{B3}	=	Temperature of rejected brine (°F)
T_{B5}	=	Temperature of brine at boiler outlet (°F)
TDS	=	Total dissolved solids taken as wt % NaCl for evaluating brine properties
T_{f3}	=	Temperature of sub-cooled working fluid at condenser outlet (°F)
ΔT_1	=	Temperature difference between inlet brine temperature, T_{B1} , and T_{sat} , (°F)
ΔT_p	=	Pinch point temperature, $T_{B6} = T_{sat}$, (°F)
ΔT_{ph}	=	$T_{sat} - T_{f3}$ (°F), Working fluid preheater
T_{sat}	=	Saturation Temperature of working fluid in boiler (°F)

v_b	=	Specific volume of brine at (ft ³ /lb)
v_f	=	Specific volume of working fluid (ft ³ /lb)
w_{pb}	=	Brine pump work (BTU/lb) of working fluid)
w_{pf}	=	Feed pump work (BTU/lb) of working fluid)
w_{pcw}	=	Cooling water pump work (BTU/lb of working fluid)
w_t	=	Net turbine work

(a) Mass flow ratio = $M_f/M_b = (h_{b1} - h_{b3}) \div (h_{f1} - h_{f3})$

The mass flow ratio of working fluid to brine is calculated by dividing the total change in enthalpy of the brine by the total change in enthalpy of the working fluid measured from the hot well to the boiler outlet. The units are pounds working fluid per pound brine.

(b) Gross turbine work = $w_t = (h_{f1} - h_{f2})$

The gross turbine work per pound of working fluid is calculated by the isentropic difference in enthalpy between the turbine inlet and the state corresponding to the condenser pressure of 83 psia. For all of the cycles studied it was assumed that the vapor state at the boiler exit lies on the saturated vapor portion of the working fluid "dome" at the boiling temperatures of interest, so that the cycles considered operate with essentially saturated vapor.

(c) Boiler pressure (psia) = $P_1 = P_{sat} + P_{wv}$

The boiler pressure was calculated by assuming that the water vapor was in thermal equilibrium with the working fluid vapor leaving the boiler ($T_{wv} = T_{f1}$). The water vapor pressure corresponding to the working fluid temperature was then added to the saturation pressure of the working fluid.

(d) Feed pump work = $w_{pf} = (P_1 - 83) v_{f3} \frac{144}{778}$

The boiler feed pump work was calculated by assuming that for all cases the condensing temperature was 110°F and condenser pressure 83 psia. The condensate is assumed to be subcooled 5°F, to 105°F and the feed pump efficiency is 75%. The fraction 144/778 converts the pump work to units of BTU/lb of working fluid.

(e) Brine pump work = $w_{pb} = (P_1 - 120) v_{b1} \cdot \frac{144}{778}$

The brine pump work was calculated by assuming that the brine was delivered to the brine pump at a constant pressure of 120 psia, a constant temperature of 325°F and a total dissolved solids content of 210,000 ppm. The term M_f/M_b converts the brine pump work into units of BTU per pound of working fluid.

$$(f) \text{ Mass flow ratio} = M_{CW}/M_f = \frac{(h_{f2} - h_{f3})}{(h_{CW2} - h_{CW3})}$$

The mass flow ratio of cooling water to working fluid is calculated from tabulated values of enthalpy for the working fluid and cooling water at the condenser and points. There is no regeneration assumed in the cycle and the small amount of desuperheating required is accomplished in the condenser. Sub-Cooling in the condenser is assumed to be a constant 5°F so that the working fluid at the hotwell is 105°F and 83 psia.

$$(g) \frac{\text{Cooling Water}}{\text{Pump Work}} = W_{PCW} = 78.3(v_{CW}/N_p)(M_{CW}/M_f) \cdot \frac{144}{778}$$

The cooling water pump work assumes the use of a direct contact condenser and a dry cooling tower. The working fluid is condensed by direct contact with cooling water and the cooling water is then circulated through a dry cooling tower after the liquid working fluid is separated from the cooling water. The 78.3 represents the cooling water pumping total head required (psi). The term (M_{CW}/M_f) converts the cooling water pump work into units of BTUs per pound of working fluid. The incorporation of this type of heat rejection system into the model results in cooling water pump work about 8 times greater than with a conventional surface condenser using cooling water.

$$(h) \text{ Effectiveness} = E = (W_t - W_{pf} - W_{pb} - W_{PCW})(M_f/M_b)$$

The "effectiveness" of the brine/working fluid cycle is the net work per pounds of geothermal brine in BTUs per pound of brine. In comparing various cycle studies the specific heat of the brine must be specified and adjustments made if they are different. It can be seen from equation (a) that if a "brine" with an average specific heat of 1.0 is compared to a brine having an average specific heat of .817, the flow ratios change and the effectiveness comparison will not be meaningful. An effectiveness based on a brine specific heat of 1.0 and will be 22% higher than the analysis using a brine specific heat of .817, all other variables being the same.

(i) Water Vapor Effect

The formulas given above and figure 2-1 do not incorporate the effect of water vapor that is produced in the boiler due to vaporization of a small portion of the geothermal fluid. The amount of water vapor as a fraction of the working fluid is proportional to the partial pressure of the water vapor divided by the total pressure. This is approximately 1.7% by weight. The effects of this water vapor on determining cycle state points and system effectiveness is discussed in Appendix D.

2.4

BRINE PROPERTIES

One of the ground rules established for this study was the Temperature range of the brine. The brine was specified as a "saturated solution at a temperature of 300°F to 350°F, with a prograde solubility dependence". This is, of course, a very generalized description of only one geothermal brine type that may be a candidate for application of direct contact heat exchangers in conjunction with a binary cycle.

For our binary cycle model we have used a brine temperature of 325°F. This temperature is taken as the inlet condition at the direct contact boiler and not at the well head or down the hole. The temperature of 325 F, while admittedly arbitrary, could apply to a situation where the brine is initially flashed down to produce steam for a steam power cycle or other steam process. Similarly, if down hole pumps are not available, the wells may rely on flashing a small part of the brine to induce flow. In either case, the brine would then be available for providing heat to a binary cycle. This situation would apply only for the hotter brines, say 350°F to 450°F and for these the brine concentration would be increased and the available temperature decreased. This might best apply to the brines in the Salton Sea area where, for example, a brine at 434°F and 360 psia could be flashed down to 325°F and 120 psia for initial steam production and still have considerable use for energy extraction in a binary cycle.

For determining the thermodynamic properties of brine, we have assumed a solution of 210,000 ppm NaCl (21% by weight). This was believed to be representative of some of the highly saline brines and all of the properties for this type of brine are widely published. A brine having a solids content of 210,000 ppm TDS was also prepared for use with the experimental heat exchangers, but it contained a variety of salts to simulate a geothermal brine. This is described in more detail in the discussion of experiments.

Throughout this report enthalpy and specific heat values for the brine were obtained from the OSW Handbook. The specific heat at 21% NaCl happens to be very nearly constant at .817 BTU/lb °F whereas for fresh water the specific heat varies from 1.0 to 1.05 in the temperature range of interest.

2.5 HEAT EXCHANGER CONCEPTS

2.5.1 General

After isobutane was selected for the working fluid and the brine classification was tentatively established, our efforts turned to examining the various concepts for direct contact heat exchange.

Considering the entire brine to isobutane heat exchange process, the subcritical process comprises three modes of heat transfer: liquid-liquid preheating, boiling and liquid-gas superheating. It is interesting to examine the relative importance of each mode by noting the percent of the total head load being transferred in each mode. These are as follows:

- a. Brine to liquid isobutane: 51%
- b. Brine to boiling isobutane: 49%
- c. Brine to isobutane vapor 0%

These percentages are for the conceptual design which has a boiling temperature of 240°F and does not incorporate superheating, the third mode of heat transfer. The percentage for superheating could be made as high as 23% in an extreme case and the other two modes reduced proportionately. Varying the boiling temperature will also change the

relative percentages for preheating and boiling. For example, a cycle with a boiling temperature of 220°F and no superheat transfers 57% of the total heat load in the boiler as opposed to 49% for the 240°F case. It should be noted that heat transfer coefficients for boiling are generally several times higher than liquid-liquid heat transfer and that liquid-gas coefficients such as found in the superheater are the lowest of the three modes. From these considerations, it was apparent that the liquid-liquid preheater would require more heat transfer volume than the boiler and would represent the highest cost item in the system. The research efforts were proportioned accordingly.

To efficiently transfer the heat from the brine to the isobutane by direct contact in preheating requires satisfying the following criteria:

- a. Provide intimate mixing of the brine and isobutane in order to develop the highest specific interfacial area possible. (Heat transfer surface per unit volume).
- b. The heat transfer fluids should approach counter-current flow as much as possible.
- c. The local heat transfer coefficients should be as high as possible to minimize equipment volume.

For the boiling mode, counter-current flow is not necessary for good efficiency.

2.5.2 Co-Current or Mixing Exchangers

Various concepts for direct contact heat exchange can be classified as counter-current or co-current. The pipe mixer and the free surface tray shown in Figure 2-2 are both co-current devices. There are other variations of co-current devices such as the agitated vessels (mixers) and venturi type mixers, all of them suffering the inefficiencies of a co-current or mixing type heat exchanger. Several methods of concatenating co-current devices in such a way as to approach a gross counter-current effect have been proposed. All of these, however, require the liquids to be separated after each stage. In view of the enormous amounts of liquid that will be flowing through the system such complications as staged co-current units and separators quickly become uneconomical.

The free surface type of exchanger is the most unlikely method because it fails two of the criteria. First, it is a co-current device and secondly it has a very low interfacial area per unit volume. Assuming that a multitude of sloped rectangular trays can be tightly stacked in a horizontal cylindrical vessel, the best specific area practical would be six square feet per cubic foot of volume of vessel. This assumes that the brine and isobutane are in contact only between the two films or layers, the isobutane floating on the brine. The specific area could be increased by the use of turbulence promoters or fingers that intersect the two layers and cause mixing, but the developed interfacial area would still be relatively low.

The co-current pipe mixer is somewhat more promising because a dispersion of small isobutane droplets can be produced in the brine.

FIGURE 2-2
CO-CURRENT CONTACTORS

FREE SURFACE CONTACTOR - TRAY TYPE
INTERFACIAL AREA = 6 SQ. FT./CU.FT.

PIPE MIXER
INTERFACIAL AREA = $\frac{6\phi}{dp}$ (230 sq. ft/cu ft) possible

The interfacial area developed in a dispersion of droplets is calculated by the relationship

$$A \text{ (sq.ft)} = \frac{6\phi}{d_p}$$

Where ϕ is the volumetric fraction of isobutane in the pipe and d_p is the droplet diameter in feet. With a hold up of 40% and droplet diameter of .125 inch the specific area is 230 sq.ft./cu.ft. While this is quite an improvement over the free surface type of contactor the criteria of counter-flow is not met. The local heat transfer coefficients will be high at the point where droplets are formed and mixed with the brine. This is caused by the turbulence within the droplet as well as the relative motion of the droplet through the brine. After initial mixing however, the turbulence decays and the fluids begin to separate. The droplet Reynolds numbers decrease unless sufficient velocity or mechanical devices are used to maintain turbulence. At 100% mixing efficiency, the mixing devices can approach a heat exchange effectiveness of 50%. To carry out the liquid-liquid preheat five theoretical stages of liquid-liquid preheat/mixers would be required in a typical binary cycle application of six to seven units having efficiencies of 70 to 80%. To attain an overall counter-current effect, each mixing unit would require an isobutane-brine separator and a booster pump for the isobutane between each stage in addition to the brine booster pump required for all direct contact systems.

While a preheating system utilizing pipe mixers does not appear promising, a pipe mixing boiler is relatively attractive. Co-current flow for pure boiling is, of course, an acceptable flow arrangement from a thermal efficiency standpoint. The boiling would be initiated by introducing the isobutane and brine concurrently in one end of the pipe where it is mixed and vaporization initiated. The vaporization would accelerate the two phase mixture along the pipe producing considerable turbulence and good heat transfer. The only disadvantage noted for this method of boiling is the requirement of a liquid-gas separator at the end of the pipe-boiler. This could be accomplished by connecting the end of the pipe tangentially to a cylindrical vessel in the manner of cyclone separators. This concept was incorporated into our test plan and the experimental boiler was built with a tangential inlet so that a pipe-boiler could be attached at a later date. In this way, the boiler which we were to test first would, in effect, become the cyclone separator for the pipe boiler. Once testing was under way, however, it was noted that the boiler being tested was performing extremely well so, in accordance with our earlier decision, time and effort was concentrated on improving the liquid-liquid preheater and the pipe boiler was not tested.

2.5.3 Counter-Flow Exchangers

The choice of isobutane suggested that whatever concept was chosen should take advantage of isobutane's low density and attending facility for separation from water.

Two types of counter-flow exchangers were considered in the selection process, both of which were originally developed for liquid-liquid extraction (mass transfer operations) rather than heat transfer.

Spray Towers and Tray Columns are both classified as continuous contact (as opposed to stagewise) counter-flow devices. They are characterized by an overall vertical counterflow of the contacted liquids being maintained by the density difference of the two fluids. The lighter fluid (isobutane in this case) is continuously introduced at the bottom of the column as a dispersion of droplets and rises to the top due to buoyancy forces, while the heavier liquid, brine, flows downward. The Spray Tower and the Tray Columns differ in that the Tray Columns contain a series of perforated trays which increase the efficiency of the mass or heat transfer per unit of height; whereas the Spray Tower is merely an empty shell. Figure 2-3 illustrates the two concepts.

A third type, the Packed Column, resembles the spray tower except that the interior of the shell is filled with packing to increase efficiency. This device was not considered seriously because it is believed that the packing would become fouled rapidly by the brine. Also, the maximum possible through-put of fluids is about 6 to 8 times less than the spray tower.

The spray tower was selected as the most promising type of direct contact heat exchanger on the basis of its higher flow capacity and possibly less fouling and corrosion. It was estimated that for an isobutane brine system, the spray tower could handle a volumetric flow of three to four times that of a perforated plate tower. In view of the fouling and corrosion potential of the highly saline geothermal brines it also appeared advisable to avoid internal structures within the vertical shell wherever possible. The tray column, with a somewhat higher efficiency per unit height will result in a tower that is shorter than a spray tower of equal overall efficiency. However, since the flow capacity is smaller by a factor of 3 to 4, it will have a column diameter as much as twice that of the spray tower. In the end, one may have two towers of different height and diameter but with equal amounts of steel and construction cost.

While the above description is admittedly over-simplified, it points out the key factors in comparing the two concepts. The heat transfer from brine to isobutane can be effectively accomplished in both the spray tower or the tray column. The spray tower was selected on the merits of simplicity and the belief that the cost of transferring heat in the spray tower will be somewhat less than in a tray column.

2.5.4 The Spray Tower - Background

The spray tower has been the subject of investigations as a heat exchanger for many years. The bulk of this research has been aimed at desalination plants where heat transfer without metallic surfaces would be desirable. The motive for developing the direct contact heat exchanger for desalination parallels that for geothermal brines. Both processes are subject to the effects of hot saline solutions which include problems of scaling, plugging, and corrosion.

There are a variety of published reports dealing with the spray tower type heat exchanger. Many of them are solely concerned with the phenomenon of heat transfer to, or from, a single liquid droplet rising

FIGURE 2-3

COUNTER-CURRENT DIRECT CONTACT HEAT EXCHANGE CONCEPTS

or sinking in a continuous phase of another liquid. Others have operated various types of small experimental heat exchangers involving a multitude of droplets rising through water or brine. We do not intend to rewrite or even extensively review this literature since it has been done several times already.

For review of this literature we refer the reader to Kern^{2,1}, Treyball^{5,2}, FMC^{1,2}, Orcutt,etal^{3,6}, Perry^{3,8}, Pinder^{3,9}. The unanimous conclusion that was drawn by the authors of this report and the most recent reviewers of the subject is that at the present time there is no simple, or even complicated method of designing a spray tower heat exchanger with confidence. This is particularly true of towers over 16 inches in diameter since the effect of back mixing is not well defined. This does not imply that these devices are too complicated or that large ones cannot be built, but it is a reflection on the state-of-the-art. Development of the spray tower heat exchangers by the Office of Saline Water was terminated in 1967. The Freeze Desalination process which employs low temperature boiling of freon in direct contact with sea water is still actively researched by AVCO Corporation^{3,14,19} but this process has little in common with the spray tower or the objectives and methods of this study.

2.6 DESIGN OF EXPERIMENT

2.6.1 Initial Considerations

It was observed that the experimental results from the majority of previous direct contact research had been obtained from very small bench scale units that are operated intermittently. Many of these experimental results are somewhat unusable for the design of large units because of hydraulic, thermal and operational dissimilarities with a large unit. As a result of this observation, and a desire to proceed with the development of the direct contact heat exchanger as quickly as possible, it was decided to build the largest test unit possible with the allotted funds. In addition, a unit was conceived which could operate continuously with the actual fluids selected, isobutane and geothermal brine.

2.6.2 Glass Column

One of the major sacrifices required by the decision to experiment with isobutane stems from the high operating pressures developed by isobutane, which in turn, prevented incorporation of a glass experimental spray column. The use of glass would be highly desirable since it allows visual observation of the inside of the tower. To incorporate a glass column would require the use of a lower pressure working fluid or to operate the unit with isobutane at low temperatures and pressures. The large diameter glass pipes (6" or larger) that were available were restricted to an operating pressure of 50 psi. This meant that the isobutane could not reach a temperature over 90°F. This idea was abandoned as impractical, unsafe and very expensive.

2.6.3 Separate Units

Another important decision in the design of the experiment was to incorporate a separate preheater and boiler. For a large commercial unit it may be desirable to combine the preheater and boiler. For the experiment, however, it was decided to construct two separate vessels so that the performance of each could be distinguished separately and also to allow the operation of one without the other. It was felt that it was more important at this stage of development to understand the individual phenomena taking place in the boiler and preheater separately than to confuse the issue with instabilities resulting from a combined unit.

2.6.4 Continuous Operation

The requirements of continuous operation required several decisions. It was originally planned to incorporate a direct contact "brine" heater that would heat the simulated geothermal brine without utilizing a surface heater. The idea was that the direct contact brine heater (not to be confused with the experimental isobutane preheater or boiler) would not become scaled or clogged by the brine. It would also produce a brine circuit that would not be "closed" and would have the advantage of not accumulating corrosion products, or other undesirable contaminants. The heating steam would be injected directly into the brine stream and an equal amount of water would be continuously blown down. To maintain the desired chemical composition of the simulated brine, fresh chemicals would be continuously added. This last requirement led to the abandonment of the concept. Two reasons made continuous chemical feed impractical. Since steam would be diluting the brine, the chemicals would have to be added without further addition of water, and an in-line dissolver would be required. The system would be operating under 315 psia pressure and all equipment heavily built. Therefore, the salt dissolver, being in line, and required to hold large quantities of salt, would be very expensive. A system utilizing pre-mixed geothermal brine in a closed loop was consequently utilized.

2.6.5 Materials of Construction

With regard to materials of construction it was felt that for the direct contact system to be significantly more economical than surface heat exchangers, carbon steel should be utilized wherever possible. This decision was based on the belief that the geothermal brine feed would be essentially oxygen free and that contact with corrosive non-condensable gases would not occur in the heat exchanger vessels. Only long term experience with geothermal brines will confirm the effectiveness of carbon steel, however.

2.6.6 Process Flow Diagram

Preliminary process calculations were carried out concurrently with conceptual studies of the experimental heat exchangers to determine the largest affordable test system and one which would simulate the actual process flow conditions of a full sized heat exchange system. The resulting flow sheet for the experimental system is shown in Figure 2-4. The decision to use a boiling temperature of 220°F was influenced by the pressure developed by the isobutane. The cost of some components increased significantly with pressure. It was also noted that

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the thermodynamic efficiency of a system operating at 220°F was not very much less than at 250°F, and that very little would be learned by increasing the operating temperature of the experimental unit above 220°F.

It was estimated that a single preheater, as envisioned for the test, would not be capable of heating the isobutane 110°F as required in a commercial unit with a 220°F boiling temperature. The process flow sheet for the experimental system indicates two modes of operation. This reflects the desire to test the experimental preheater at both the high and low range of temperatures expected in a power plant application. It would have been impractical to operate the boiler during low temperature testing of the preheater without imposing an unrealistic temperature profile on the boiler. Therefore, the Mode I flow sheet for the test unit shows the boiler being bypassed during low temperature operation of the preheater and the isobutane being sent to a cyclone separator. In actuality, the cyclone separator was never incorporated in the test circuit. Instead, the isobutane from the preheater was throttled and sent directly to the condenser as a two phase mixture. Figure 2-5 presents the two modes of test operations in terms of the temperature-enthalpy diagram for isobutane. The dashed lines represent the brine and isobutane temperature profiles of the binary cycle that was simulated.

2.7 EXPERIMENTAL SPRAY TOWER PREHEATER

2.7.1 Approach

Design of the experimental heat exchangers commenced as soon as the process flow sheet was established and heat and mass balances were known for the experimental conditions. The design of the spray tower was approached in the following sequence:

- a. Estimate the tower diameter based on the flow requirements (10 gpm isobutane 5 gpm brine).
- b. Estimate the tower height based on heat transfer requirements.

2.7.2 Tower Diameter

To estimate the tower diameter requires the knowledge of the dispersed (isobutane) and continuous (brine) phase flow rates that will cause "flooding" of the tower. For each ratio of flow rates there is a maximum throughput for a given tower diameter. The best known method for estimating the flow rates at flooding is that of Minard and Johnson. Flooding, as defined by Minard and Johnson, is that condition in the spray tower where the interface between the zone of freely rising drops and the tightly packed bed of drops would move downward until droplets were rejected out of the bottom outlet with the brine. In this case, the mass of droplets would move downward because the flow and consequent pressure drop of the brine would be overcoming the terminal rise velocities of the isobutane.

The relationship between superficial velocities at flooding is:

$$V_c^{.5} = \left[47 N_c^{-.07} \cdot \Delta SG^{.14} \cdot SG_c^{-.5} \right] - \left[1.8 D_p^{+.056} \cdot N_c^{-.075} \cdot (SG_d/SG_c)^{.5} \cdot V_d^{.5} \right]$$

Where:

V_c = continuous phase (brine) superficial velocity Ft³/hr-ft²

V_d = dispersed phase (isobutane) superficial velocity

ΔSG = specific gravity difference, brine-isobutane

N_c = continuous phase viscosity, centipoise

D_p = drop diameter, inches

SG_c = continuous phase specific gravity

SG_d = dispersed phase specific gravity

The above empirical expression requires a knowledge of drop diameter in addition to the fluid properties. The expression can be evaluated for an assumed drop diameter and $V_d^{1/2}$ vs. $V_c^{1/2}$ can be plotted as shown in figure 2-6. An alternate form which substitutes an orifice coefficient for the droplet diameter is also given by Minard & Johnson.

Both expressions were evaluated for isobutane and brine at various temperature conditions according to the process flow sheets for Modes I and II. Figure 2-6 shows the flood lines for droplet diameters of .15 inch and the flood lines using the alternate expression with an orifice coefficient for an orifice diameter of .062 inch.

Also shown in figure 2-6 are the limits imposed on V_d and V_c by the pumps selected for the test unit. The sloped line extending from the origin represents a fixed ratio of isobutane flow to brine flow that is equal to 2.0 (10 gpm/5 gpm). The intersection of this fixed flow ratio with the estimated flooding lines indicates the superficial velocities of isobutane and brine that will produce flooding at that flow ratio. Flooding, as mentioned earlier, is the limiting condition, not the desired operating condition. The tower should be sized so that the desired flow can be passed without danger of flooding as defined by Minard & Johnson. As a result the tower should be somewhat larger than indicated by the flooded limits.

The alternate form of the Minard and Johnson equation using the nozzle coefficient indicates a tower of 8.6 inch diameter would flood at considerably lower flow rates than desired. On the other hand, the flooding line estimated by assuming a droplet diameter of .15 inch indicated a 6.2 inch diameter column would flood at the design flow rates. It was decided to use a six inch diameter, schedule 40 carbon steel pipe for the midsection of the spray tower and attempt to produce a drop diameter of .15 inch. The isobutane distributor, discussed in detail in section 3, was designed to produce a droplet diameter of .15 inch (1.8mm). The largest error in the Minard & Johnson correlation is apparently introduced by the droplet diameter function. By reviewing their experiment results, it can be seen that the influence of droplet

FIGURE 2-6
SUPERFICIAL VELOCITIES OF
ISOBUTANE & WATER AT FLOODING

LIMITING SUPERFICIAL VELOCITIES

BASED ON TOWER CROSS SECTIONAL

AREA. ($\text{FT}^3/\text{HR} \cdot \text{FT}^2$)

V_d = DISPERSED PHASE (ISOBUTANE)

V_c = CONTINUOUS PHASE (WATER)

diameter appears to be difficult to pinpoint accurately. Drop diameters that will be produced for a given liquid-liquid system are very difficult to estimate accurately without prior experiments with a given nozzle configuration.

2.7.3 Distributor

Having decided on an approximate drop diameter, an isobutane distributor must be designed to produce that size drop within the constraints imposed by column diameter, orifice spacing, flow rate and cost. The type of distributor generally employed in spray columns consists of a flat plate containing a large number of orifices, variously formed by drilling, punching or separate inserts. The 0.15 inch drop diameter assumed above was arrived at after preliminary calculations and a review of other spray column investigations indicated that drops of that size might reasonably be expected from an orifice diameter of 0.060 inch. In view of machinability and clogging susceptibility 0.060 inch was felt to be a practical minimum orifice diameter.

The number of orifices required for a given diameter and design flow rate is determined by the orifice velocity. In general, three drop formation regimes are encountered as nozzle velocity increases from very low values up to a critical atomization threshold⁵². At low velocities, drops are formed slowly and individually at the orifice opening. As velocity increases, drops are formed by the break-up of a jet of liquid that extends into the continuous phase. Ultimately, at higher velocity, the jet disappears and atomization to small, non-uniform drops occurs at the orifice itself. By a relationship given in reference (4) this critical velocity was estimated at 3.3 ft/sec. for an isobutane-brine system with an .060 inch orifice diameter. For drop formation by jet break-up, Christiansen and Hixon⁹ found that the maximum interfacial area occurs at a particular orifice velocity for each system. Their results are expressed in more usable form in reference (52) from which an optimum velocity for this system of 1.6 ft/sec. was calculated. The corresponding drop diameter is about 0.11 inch. However, this velocity required too many orifices for the available cross sectional area of the tower. Furthermore, in this experimental unit, the isobutane flow would often be at less than the design rate. Therefore it was decided to use a velocity of 3.0 ft/sec. at an isobutane flow of 10 gpm. This is close to the upper limit of the jet break-up regime and the drop diameter was expected to be somewhat larger than the 0.11 inch minimum. The orifice configuration and pattern is shown in Figure 3-2. An interfacial tension of 58 dynes/cm calculated from the "Fowkes' relationship" for saturated hydrocarbons and water has been used in the above calculations.

2.7.4 Tower Height

The tower height is determined from the heat transfer requirements of the system and knowledge of the thermal performance of the spray tower. Unfortunately, methods of estimating heat transfer coefficients for spray towers are unconfirmed and many are in conflict with each other. To quote Kern²¹ who has reviewed most of the previous investigations "the seawater heat transfer data which are available resulted from sincerely conceived investigations but the results of each investigation are at best comparable within themselves and not from investigation to investigation". The height of the column was eventually sized

by selecting a volumetric heat transfer coefficient of 5000 BTU/HR-°F-Ft³. By referring back to Figure 2-5, the heat load and log mean temperature difference for Mode II can be determined as 35,775 BTU/HR and 6°F respectively. The required volume, then, is 35,775/(5000 x 6) or 1.2 cubic feet. For a diameter of 6 inches the height required is six feet. During test operations, a coefficient of only 3500 was attained and as a result, the log mean temperature difference had to be increased slightly in order to maintain the same heat load indicated on the process flow sheet.

2.8 EXPERIMENTAL BOILER

2.8.1 Heat Transfer Design

Heat transfer data on direct contact boiling is even more scarce than liquid-liquid heating. As a result of reviewing all available literature, the boiler was tentatively sized by simply assuming a volumetric heat transfer coefficient of 15,000 BTU/hr-°F-ft³, and applying it with the overall LMTD and heat load, as specified by the flow sheet. This method of sizing a boiler is shamefully unscientific, but a more credible method was not found. From Mode II flow sheet (figure 2-4) an LMTD of 32.8°F and boiler heat load of 230,000 BTU/hr was used to arrive at a boiler volume of .46 cubic feet.

2.8.2 Sizing

A diameter of six inches was assumed, again quite arbitrarily, so that a height of 28 inches was established to attain .46 cubic feet for the effective boiling volume. To this was added 15 inches for the superheating zone, 6 inches for the demisters, 4 inches for the brine inlet nozzle, and 19 inches below the isobutane inlet for brine surge volume and outlet cone, making the total height 72 inches. See figure 3-3 for a schematic of the boiler and drawing no. 051-2204 in the Appendix "C" for details of the boiler. Thermally and hydraulically, the boiler is essentially an agitated mixing chamber. The agitation is created by the injection of isobutane directly into the brine. Further agitation is created by the rapid vaporization and rising of the isobutane droplets. The isobutane is injected by means of two tangential nozzles which penetrate the boiler wall near the bottom of the vessel.

2.8.3 Isobutane Distribution

The tangential jets were used in an effort to impart a spinning motion to the fluids in the boiler that would improve mixing and provide a definite circulation pattern. Consider the inside of the boiler where the brine is rotating about the vertical axis of the boiler and isobutane droplets are introduced tangentially at the periphery of the boiler. Because of centrifugal forces the brine tends to flow outward to the walls of the boiler displacing the vaporizing isobutane droplets that are beginning to rise upward due to buoyant forces. The net effect is to move the isobutane away from the wall towards the center in a spiral motion and consequently mix the isobutane with fresh incoming brine. The boiler has proved quite successful and coefficients of 17,000 BTU/HR-°F-Ft³ were attained.

SECTION 3

EXPERIMENTAL STUDIES

3.1 TEST UNIT DESCRIPTION

3.1.1 General Description

The test unit provides for the continuous circulation, contact and separation of two immiscible fluids, brine and isobutane. Heat is added to the system in a brine heater, transferred to the isobutane in two direct contact heat exchangers, the preheater and boiler, then rejected in a water cooled condenser. Other major components are the separator, hot well, surge tank, and two circulating pumps. The general relationship of these components is illustrated in the process flow sheet, Figure 2-4. (in previous section), and the P&I diagram, Drawing No.051-6101, in Appendix C.

Isobutane liquid is pumped from the hotwell to the preheater where it is heated to the boiler saturation temperature by counter current contact with brine. The isobutane flows, still a liquid, to a separate boiler due to a pressure differential between the two vessels. In the boiler, vaporization and superheating of the isobutane are accomplished by contact with the brine. The mixture of isobutane and water vapor is then throttled to its condensing pressure through a globe-needle valve in lieu of expansion in a turbine. The vapor is condensed and the isobutane returns to the hot well to repeat the cycle. Water condensate is periodically drained off from the condenser and hot well.

Hot brine from the brine heater enters the direct contact boiler where heat is extracted from it to vaporize and superheat the isobutane. The brine is then pumped to the preheater where additional heat is removed by contact with the isobutane. The cooled brine leaving the preheater enters the separator where any small, entrained isobutane drops are coalesced and settled out. Brine flow through the separator, brine heater and return to the boiler is provided by the pressure differential between the preheater and boiler.

3.1.2 Vessel Design

All test vessels (except the two surface type heat exchangers) were fabricated of carbon steel to Section VIII, Division 1 of the ASME Boiler and Pressure Vessel Code. Plate, standard weight pipe (sch.40) and welding fittings were used for the shells. Nozzles were either schedule 80 pipe with 300# flanges or 3000# screwed couplings. Construction details are shown on the drawings in Appendix C. Overpressure protection was provided by frangible carbon rupture discs or, where appropriate, by small safety relief valves. The brine heater was constructed of carbon steel tubing, swagelok and screwed fittings, while the condenser was a standard off-the-shelf item with brass shell and tubes. Piping was predominantly carbon steel hydraulic tubing to facilitate bending and minimize the number of joints. Swagelok type fittings were used extensively to allow frequent disassembly and modification of the piping system. Hydraulic control of the test unit was accomplished with manually operated globe-needle valves. The electrical installation was done in accordance with the National Electrical Code for Class I, Division 2 locations.

3.1.3

Preheater Design

The preheater is a liquid-liquid spray column having an overall height of about ten feet and distance between distributors of six feet. The straight section of the column is five feet in length with an inside diameter of six inches. At both ends a conical section with an included angle of 20 degrees flares to a ten inch diameter. The transition at the bottom results in a gradually decreasing downward velocity of the brine and the enlarged diameter at the bottom allows sufficient area in the annulus between the isobutane distributor and the wall to reduce the brine velocity below its value in the straight section. At the top of the column the ten inch diameter provides a large area for coalescence of the isobutane drops as they reach the interface. The column is shown schematically in Figure 3-1 and in detail by drawing 051-2202, Appendix C.

Isobutane is introduced near the bottom of the column by a plate containing 390 drilled holes 0.060 inch in diameter. Type 316 stainless steel, 1/4 inch thick was used for the plate. The holes were drilled in concentric circular patterns as shown in Figure 3-2. On the bottom face of the plate the edges of the holes were beveled while the upper face of the plate was ground to ensure sharp edges. After machining, the surface of the plate was slightly oxidized to make it more oleophobic (less wettable by isobutane) by pickling in concentrated nitric acid.

Brine is introduced through a perforated distribution ring near the top of the column. The purpose of this distributor is to minimize turbulence caused by the injection of the brine. Immediately above the brine distributor is a plastic screen at which the isobutane droplets coalesce. This screen, being oleophilic, is preferentially wetted by the isobutane and will exclude brine from leaving with the isobutane. The isobutane-brine interface level is maintained at this screen during normal operation.

The construction of the column incorporates four pairs of flanges to permit internal access and facilitate testing of modifications. Various connections are provided on the shell of the column for vents, drains and instrumentation including three thermowells, one pressure gage connection, two manometer taps, four high pressure windows (or "bullseyes") for viewing the dispersion, and three pairs of gage glass connections.

3.1.4

Boiler Design

The direct contact boiler, shown in Figure 3-3, is a cylindrical vessel six feet tall with an inside diameter of six inches. Brine is introduced near the top through a spray nozzle and exits at the bottom. Isobutane liquid from the preheater enters through two tangential nozzles in the lower part of the vessel and is vaporized (and slightly superheated) by contact with the brine. The vapor outlet is at the top of the boiler. For construction details, see Drawing 051-2204 in Appendix C.

A full cone, narrow angle spray nozzle (Sprayco #2136) is used to distribute the brine near the top of the vessel. This is a swirl type nozzle which produces relatively large droplets rather than a fine mist to discourage liquid brine carryover in the vapor stream leaving

DIRECT CONTACT PREHEATER
LIQUID - LIQUID HEAT EXCHANGER

FIGURE 3-1

FIGURE 3-2
ISOBUTANE DISPERSION PLATE ASSEMBLY

ORIFICE LOCATIONS	
CIRCLE ϕ	ARC
1"	30°
1½"	20°
2"	15°
2½"	12°
3"	10°
3½"	9°
4"	7°30'
4½"	6°20'
5"	6°
5½"	5°18'

FIGURE 3-3
DIRECT CONTACT BOILER

the boiler. Before leaving the boiler, the vapor must pass through a knit stainless steel wire mesh demister pad six inches deep which minimizes carryover of either liquid isobutane or brine.

Situated in the brine outlet at the bottom of the boiler are a vortex breaker and a perforated steel plate. The vortex breaker straightens the spinning brine flow while the perforated plate immediately below excludes random drops of unvaporized isobutane from the brine outlet. These devices prevent entrainment of isobutane in the brine stream and subsequent emulsification in the brine pump.

Another nozzle is located tangentially about midway up the vessel. Its main purpose is to allow testing of a concurrent pipeflow boiler. In that configuration, the main body of the boiler would serve mostly as a cyclone separator for the vapor and liquids. This nozzle may also be used as an alternate means of introducing brine into the boiler.

The boiler is flanged at top and bottom for access and modification. Instrumentation connections include four thermowells, one pressure gage tap and one pair of gage glass connections.

3.1.5 Separator

Brine leaving the preheater enters the separator which is a settling tank with a coalescer at the inlet end. The vessel is seven feet long and eight inches in diameter, providing a retention time of 3.7 minutes for the brine. It is installed with its longitudinal axis tilted up one degree from the horizontal toward the outlet end. The isobutane brine interface level in the separator is visible in a gage glass.

The coalescer, which fills the entire cross section of the separator at the inlet end, consists of a knit mesh of stainless steel wire and glass fiber. Its function is to entrap and hold the small isobutane drops entrained in the brine stream so that they can coalesce with other small drops and form drops large enough to settle out. Although it is usual to employ a fiber that is preferentially wet by the dispersed phase, stainless steel and glass are hydrophilic rather than oleophilic. In this case, the materials were chosen for durability and heat resistance during fabrication since the vessel was closed by welding rather than by flanging.

Besides recovering small quantities of dispersed liquid isobutane (as opposed to dissolved isobutane) from the brine discharge stream during normal operating conditions, the separator also provides volume to cope with the large amounts of isobutane entrained in the brine during upset conditions. Isobutane liquid collected is periodically drained off to the hotwell which is always at a lower pressure than the separator when the test unit is in operation.

3.1.6 Hotwell

The hot well serves as a receiver for condensed isobutane from which the isobutane pump takes suction and provides some residence time to improve pump suction conditions by further subcooling the condensate.

As test unit operating conditions change, the amounts of isobutane in the preheater, boiler and separator vary. The hotwell provides the inventory to cope with these operating changes as well as minor losses through leakage. Instrumentation on the hot well includes temperature and pressure indication and two gage glasses.

3.1.7 Condenser

Isobutane vapor from the boiler, after throttling, is condensed on the shell side of a conventional shell-and-tube heat exchanger by sea water flowing through a single pass of tubes. The unit is installed vertically with vapor entering at the top and condensate exiting at the bottom; sea water flows counter to the isobutane. The condensate outlet and a vent line are connected to the hot well in such a way that a liquid level is maintained in the bottom of the condenser to sub cool the condensate. Heat transfer area is 120 sq. ft.

3.1.8 Brine Heater

Saturated steam at 120 psig is used to heat the brine before it enters the direct contact boiler. The brine heater is a double pipe heat exchanger with brine flowing inside the central tube and steam flowing in the opposite direction through the annular space between the tubes. This arrangement was selected because it is inexpensive and, having no tube sheet, it is suited to high pressures. The brine heater is simply a hot water heater and has no counterpart in a system using actual geothermal fluid.

3.1.9 Brine Surge Tank

This vessel is located off-line and provides brine make-up capacity to maintain the brine level in the boiler by adding or subtracting brine. This is accomplished by controlling the pressure in the tank with nitrogen gas. The surge tank is also connected to the suction side of the brine pump and acts as an accumulator to reduce pressure pulsation.

3.1.10 Pumps

Brine is circulated by a two cylinder John Bean plunger pump (Model B-0410C) with a rated capacity of 5 GPM. A gas filled accumulator on the discharge side is used to reduce pressure pulsation. Flow rate is varied by a valve on a bypass line from the discharge to suction side of the pump. A relief valve is provided on the discharge to prevent overpressure damage to the pump or test unit.

The isobutane pump (John Bean Model F-0410C) rated at 10 GPM, is similar except that it has four cylinders to reduce flow pulsation in the suction line. Pumping requirements for the isobutane are especially demanding because of the combination of low available net positive suction head and high discharge head in a single pump. Liquid isobutane also has extremely low density and viscosity, placing further restrictions on the type of pump which can be used.

3.1.11 Instrumentation

Temperatures are measured with mercury-in-glass thermometers, bimetal thermometers and iron constantan thermocouples. Specified accuracies

are as follows: mercury-in-glass, within $1/2$ °F; bimetal, $1/2\%$ of full scale; and type J thermocouples ± 2 F. Thermocouples are backed up by thermometers for local readout. Remote read-out is by a single pen recorder/indicator. Since this recorder is not explosion proof or intrinsically safe, it is located in a building about 50 feet from the test unit. A rotary switch at the test unit is used for channel selection so only one pair of wires is routed to the recorder. Three calibrated rotameters with accuracies of $\pm 1\%$ of full scale measure brine, isobutane and steam flows. These are of armored design with magnetic indicators. Pressures in the vessels are indicated by gages with a specified accuracy of $\pm 1/2\%$ of full scale. Differential pressure across the preheater is measured with a manometer for determination of dispersed phase hold-up. Level indication is by high pressure gage glasses.

As a safety precaution an explosive gas monitor was used while tests were being conducted. This unit, which is portable and intrinsically safe, was placed down-wind of the test unit and alarms when the isobutane gas concentration reaches 20 % of the lower explosive limit.

3.1.12 Isobutane Stripping Apparatus

A small stripping apparatus shown schematically in Figure 3-13 (page 71) is installed to measure the isobutane content of the brine leaving the preheater and separator. Since the brine loop is closed and the brine is recycled continuously, saturation of the brine with isobutane might be expected (but not necessarily obtained in practice as explained in Section 3.3). If this is the case, the concentration of isobutane in excess of the equilibrium amount for conditions at the sample point represents dispersed isobutane. Brine samples are blown down from full system pressure to about one atmosphere through a spray nozzle installed in a small chamber. Isobutane gas dissolved in the brine at higher pressure is released and any dispersed liquid isobutane is evaporated, causing an increase in chamber pressure. Brine then leaves the chamber through a loop seal. The method differs from other physical methods that have been used to measure gas solubilities (37,41) mainly in the respect that the solvent liquid and the gas are not measured in the same volume. This is an important consideration with slightly soluble gases like isobutane where relatively large volumes of brine contain only small amounts of dissolved gas. Operation of the apparatus and assumptions involved in data reduction are explained in Section 3.3.

3.2 DISCUSSION OF TEST OPERATIONS

3.2.1 General

All testing was conducted at the U.S. Department of the Interior's Office of Water Research and Technology test facility at Wrightsville Beach, North Carolina under an inter-agency agreement with the ERDA. This site was selected because of the existence of appropriate test pads, utilities (including steam) and support facilities such as machine shop, laboratory, etc., and personnel experienced in conducting tests of this nature. Also a geothermal desalting unit was tested at Wrightsville Beach by Envirogenics Systems Co.¹⁰ before being moved to the Bureau of Reclamation's Test Site near El Centro, California (East Mesa) for field testing.

Activities at Wrightsville Beach began in October, 1975 but installation of the test unit was not completed until January, 1976 because of delays in equipment delivery. Testing was conducted from January 29 until May 6.

Tests were arranged in order of increasing complexity, leading up to operation of the complete unit at design conditions using simulated geothermal brine. After initial cold hydraulic testing, heat transfer tests of the preheater only were run using fresh water rather than brine. When sufficient operating experience had been gained with the preheater, the boiler was brought on stream and the two direct contact heat exchangers were tested together. The last step was to substitute simulated geothermal brine for fresh water.

Since the entire temperature range of a commercial cycle could not be duplicated in the test unit, accommodations for a low-temperature preheating-only mode were made. In addition to simulating the low temperature end of the cycle, this simplified type of operation in which the boiler is bypassed is useful during system "de-bugging" and for achieving better control over preheater conditions than when both preheater and boiler are operated in combination.

The isobutane used contained from 0.8 to 1.5% propane and 1.5 to 2.5% n-butane (by weight), the balance being isobutane. This grade is usually referred to as "field run" isobutane and is the least expensive type available. The fresh water used in the test unit was Wrightsville Beach city water having a TDS content of about 300 ppm. Brine composition is discussed later in this section. In the various operating descriptions that follow the term "brine" is generally used in referring to the aqueous phase even if it was actually fresh water.

3.2.2 Test Designations

Experimental runs are designated by an alphabetic prefix followed by two numbers. The letters have the following meanings: A, cold flow; B, preheating only, fresh water; C, combined preheating and boiling, fresh water; D, preheating only, simulated geothermal brine; E, combined preheating and boiling, simulated geothermal brine. The first number is the test number and is followed by a data point number. Data point numbers do not generally follow consecutively. The numbers in between represent transient data points taken to determine when relatively steady-state operation was achieved.

3.2.3 Operating Procedures - Preheating Only

The following description of operating procedure may be more readily understood by reference to the Piping and Instrument Diagram (Drawing 051-6101) in Appendix C. Brine is first slowly heated as it is circulated through the unit. Brine temperature is controlled by steam valve CV-106 and condensate valve CV-100A. Valves CV-100 and CV-123 control brine flow. When the desired temperature is reached, isobutane is introduced to the preheater, E-100, by opening recirculation valve CV-109 and starting the isobutane pump, P-100. After opening CV-108, CV-109 is closed until the desired flow is reached. During

preheater operation the position of the isobutane-brine interface at the top of the preheater must be maintained by a pressure balance between isobutane and brine streams. This balance is held by manipulating either the isobutane outlet valve C-112 or the brine outlet valve CV-100. To cause the position of the interface to rise, CV-112 is opened or CV-100 is closed. Conversely, the interface is lowered by closing CV-112 or opening CV-100. These two valves also control the total pressure in the preheater which must be kept higher than the saturation pressure of the isobutane to prevent it from boiling.

In this mode, valve CV-110 is shut and the heated isobutane liquid leaving the preheater bypasses the boiler and goes directly to the condenser E-102. The brine, however, flows through the boiler, E-101, which acts merely as a brine reservoir in this case. As the fraction of the preheater volume occupied by the isobutane liquid (holdup) varies, so too does the brine level in the boiler since the brine loop is closed. Should it be necessary to add to or subtract from the brine inventory, the surge tank provides this capacity. It is pressurized with nitrogen gas to add make-up brine or vented to allow removal of excess brine.

Each heat transfer test usually lasted for about five hours. During this time, the unit was allowed to reach several steady state points at which data was taken. Since all controls are manual and no supervisory instrumentation is incorporated, the unit required constant attention and was, therefore, operated only during day shifts.

3.2.4 Combined Preheating and Boiling - Operating Procedures

To begin a boiling test, brine is circulated at about 4 gpm and heated slowly until the temperature at the boiler brine inlet approaches 180 F. Cooling water flow is established and the vapor throttling valve CV-111 is opened. The isobutane pump is started and flow is established by opening valves CV-108 and CV-110 to allow isobutane into the preheater and boiler. Isobutane flow is adjusted initially to about 2 gpm and a boiler pressure of 110 psig is established. The steam supply is adjusted as necessary to stabilize test unit temperatures. Since start-ups tended to be somewhat difficult to control, the test unit was always started in the same manner and stabilized at low isobutane temperatures, pressures and flows before proceeding to higher levels.

Pressures in the direct contact heat exchangers must be adjusted to satisfy three requirements at all times. First, the boiler pressure is maintained at whatever the value required for the heat transfer test by adjusting vapor throttling valve CV-111. Second, the pressure balance necessary to maintain a stable isobutane-brine interface in the preheater must be achieved. This is done by manipulating valves CV-110 and CV-100. Third, there must be sufficient pressure differential between the preheater and boiler to permit the required fluid flows but not so much differential that the preheater is over pressured.

The temperature at which the boiling occurs is regulated by the boiler pressure. If the pressure is increased without a corresponding increase in preheater isobutane outlet temperature, additional preheating must occur in the boiler. A decrease in boiler pressure below the saturation pressure corresponding to preheater outlet temperature results in flashing in CV-110, causing loss of control.

Condenser pressure and temperature are controlled mainly by regulating the cooling water flow to the condenser. Vapor leaving the boiler is throttled by valve CV-111 and enters the condenser with considerable superheat. The superheat is removed, the vapor condensed and the condensate subcooled in the condenser. The degree of subcooling depends mainly on the condensate level in the condenser which can be controlled to a limited extent by valve CV-120. This valve is on the vent line connecting the condenser and hot well and influences the pressure differential between these vessels. The line (IB01-38) by which the condensate flows from the condenser slopes up to the hotwell to help maintain a liquid level in the condenser. The water condensed along with the isobutane is drained from a low point in line IB01-38 at frequent intervals and the amount measured.

Control of this test unit differs in two important respects from an actual binary cycle using geothermal fluid. First, the temperature of the brine entering the boiler can be changed at will in the test unit simply by varying steam flow and pressure in the brine heater. This temperature would normally be fixed for geothermal fluid. Second, boiler pressure is regulated with the vapor throttling valve on the test unit. Constant boiler pressure in an actual binary cycle would be maintained by varying brine and isobutane flows according to load.

3.2.5 Operating Experience - Preheater

The first tests conducted after the system was charged with isobutane were aimed at answering three critical questions regarding preheater hydraulic operation: 1) does the dispersion plate produce uniformly sized drops of the expected size; 2) can the position of the brine-isobutane interface at the top of the preheater be readily controlled; and 3) is premature coalescence in the upper part of the column a problem? Uniformly sized droplets about 3mm in diameter are produced over a range of isobutane flows from 2 to 12 gpm. At all brine flows and at isobutane flows up to about 5 gpm, drops follow relatively straight paths as they rise through the brine. Beyond this rate, drop movement becomes more erratic and groups of drops appear to rise with a swirling motion. No tendency for the drops to coalesce in the column proper was observed although they seemed to coalesce readily enough at the interface at the top of the preheater.

The position of the interface does not seem to be influenced by the coalescing screen, i.e., there is no apparent tendency for the interface to stabilize at the screen. Instead, it is sensitive only to hydraulic pressure and is easily controlled by either the isobutane or brine outlet valves. The screen does seem to effectively exclude random brine droplets from the isobutane outlet during normal operation. During operation a scum or "rag" accumulates at the brine-isobutane interface. The scum consists, in this case, of finely divided corrosion products but would likely also contain mineral precipitates in a column operating on actual geothermal fluid.

The brine distributor at the top of the preheater was initially a simple open pipe. This arrangement was used for most of the heat transfer tests but was later modified to create a smoother, less turbulent entry of the brine into the rising isobutane drops. The modified distributor is in the form of a wheel with brine entering at the hub and flowing through the spokes to the perforated rim. A small improvement in heat transfer was noted following this modification.

3.2.6 Operating Experience - Boiler

Initial boiler tests were run with isobutane flows ranging from 2 to 4 GPM. Some clouding of the brine in the preheater occurred as contrasted with the preheating only runs in which the brine remains quite transparent as seen through the bullseyes. When isobutane flow was increased beyond about 3 GPM, the preheater dispersion became completely opaque and large quantities of isobutane liquid were collected in the separator. These problems were caused by entrainment of isobutane in the brine outlet at the bottom of the boiler. This two-phase mixture passed out of the boiler and was emulsified in the brine pump.

This boiler deficiency was corrected during the first inspection and modification period. Extensions were added to the tangential isobutane inlets inside the vessel to effectively increase the distance between these inlets and the brine outlet. A vortex breaker and a perforated steel plate were also installed in the brine outlet. The steel plate is preferentially wet by the brine and therefore tends to exclude random drops of unvaporized isobutane. These modifications solved the entrainment problem and only small amounts of isobutane collected in the separator thereafter.

It was noted during this inspection that the inside carbon steel surfaces of the boiler and preheater were coated with a black, powdery substance similar to the scum collecting at the brine-isobutane interface. This may have been largely corrosion product (ferric sulfide) caused by the action of mercaptan dissolved in the brine. For safety's sake, the isobutane used was initially stenching with ethyl mercaptan which, by its strong odor, gives warning of leaking gas. This compound was gradually extracted from the isobutane by contact with the brine and, after a week or two, the isobutane had entirely lost the distinctive mercaptan odor.

Although evaporation of the isobutane accounts for most of the heat transferred in the boiler, heating of sub-cooled liquid and superheating of vapor may also take place in this vessel. Flows of isobutane and brine are essentially counter current with the brine passing first through the isobutane vapor, then the saturated liquid and finally the sub-cooled liquid. The factor that determines how much heating of the liquid and superheating of the vapor occurs is the relationship of the isobutane saturation temperature to the temperatures of the incoming brine and isobutane. The temperature at which boiling occurs depends, of course, on the boiler pressure which is controlled by the vapor throttling valve in this case. If the boiling temperature is the same as the temperature of the isobutane leaving the preheater, boiling can begin immediately. Superheating of the vapor will occur if the temperature of the incoming brine is higher than the minimum required to completely evaporate the isobutane. In most of the tests all three types of heat transfer between the brine and isobutane take place.

It was planned to run the high temperature mode with a preheater isobutane inlet temperature of 170°F and a temperature rise of 50°F. Isobutane at 220°F would then be evaporated in the boiler without further preheating. In practice, difficulty in controlling condenser pressure and degree of subcooling of the isobutane condensate

along with isobutane pump suction problems restricted preheater isobutane inlet temperatures to 150°F or less. Since it was not possible to obtain the required temperature rise in the preheater with the available thermal driving force, considerable additional preheating was supplied by the boiler to reach the desired boiling temperature.

Boiler operation was usually started with the water level about 24 inches above the tangential isobutane inlets. If the saturation temperature in the boiler was only a few degrees above the inlet temperature, the isobutane was completely vaporized by the time it reached the brine free surface and the level in the boiler did not rise. If more preheat was required, a layer of isobutane appeared to accumulate in the boiler above the brine. Boiler level was maintained between about 24 and 33 inches above the isobutane inlets by controlling the heat input to the boiler and by adding make-up brine from the surge tank as required.

3.2.7 Operating Experience - Combination Preheater - Boiler

The last boiling heat transfer test, E-8, was somewhat different from the others. As a preliminary check on the feasibility of preheating and vaporizing in the same spray column, the preheater was operated as a combination preheater-boiler. The test was run by starting normal preheater operation and then lowering the preheater pressure until the isobutane began to vaporize in the upper part of the column. Liquid level was controlled with brine valve CV-100. Since the piping and valve for the preheater isobutane outlet was sized for liquid flow, this test was limited to about 2 gpm isobutane flow. Heat transfer performance in the preheating and boiling zones at this low flow was comparable to that of the separate preheater and boiler, respectively.

3.2.8 Miscellaneous Operating Problems

The two pumps in the system were probably the greatest single source of operating problems and unplanned shutdowns. Most persistent was the vapor binding of the isobutane pump caused by marginal suction conditions. Since it is impractical to elevate the isobutane hotwell above the pump to any significant extent, the required net positive suction head must be provided by subcooling below the saturation temperature. For cold flow tests, no subcooling is possible since there is no heat input and the entire unit operates at the same temperature. In this case it was found expedient to pressurize the condenser slightly with nitrogen gas in order to provide the necessary head. Heat transfer tests were run without this nitrogen blanket as it was considered unnecessary and detrimental to condenser performance as well. However, suction problems still occurred because of operating transients and difficulties in controlling the amount of subcooling and, for no apparent reason, at hotwell temperatures of over 150°F. Nitrogen was therefore added for some heat transfer runs. Fortunately, the condenser contained sufficient heat transfer surface that non-condensable blanketing was not a problem.

Problems of a strictly mechanical nature were experienced also, in-

cluding replacement of three suction and discharge hoses, a broken relief valve and excessive isobutane pump plunger stuffing box leakage. Evaporation of leaking isobutane often causes ice formation at the point of leakage if water is present also. The ice crystals then prevent the packing from sealing the leak until they can be melted. The same situation occurs at vent and drain valve seats when liquid isobutane is allowed to escape into the atmosphere.

Numerous minor piping modifications were made over the course of the test program. These included adding larger steam and isobutane vapor valves, the addition of a strainer and non-return valve on the isobutane pump discharge, revised vents and drains, and re-routed cooling water line.

Considerable trouble was caused throughout the test program by rupture discs bursting well below their rated pressure. Because of this lack of reliability, we were especially reluctant to operate the condenser close to its design pressure since most of the isobutane in the unit was contained in the hotwell. The supply of spare discs was also used up before the end of the test program, forcing lower pressure operation during the last week.

Because of the flammability and high vapor pressure of isobutane it was not feasible to make many inspections, modifications or repairs while the test unit was charged with isobutane. These tasks had to be deferred until the unit was gas freed. This operation involved pumping as much of the isobutane liquid as possible back into storage bottles and then venting the rest to the atmosphere. Any vessels not opened and gas freed were filled with water to eliminate vapor spaces while work was in progress.

3.2.9 Preparation of Simulated Geothermal Brine

Simulated geothermal brine was substituted for fresh water during the last month of testing. The main objective of using simulated brine was to adjust the physical properties of the brine to simulate the heat transfer and hydraulic performance of the design conditions. Other areas of interest were the reduction of isobutane solubility due to dissolved salts ("salting out") and any detrimental effects on the performance of the direct contact heat exchangers caused by suspended or deposited solids. Formation of large quantities of scale was not expected, however, due to the closed nature of the brine loop and the fact that significant concentration of the brine does not occur during operation.

A high salinity brine was prepared which approximated actual geothermal brine only in respect to a few major constituents. Sodium chloride and calcium chloride were included since they represent the bulk of total dissolved solids in high salinity brines. Silica was added because it is known to be a major contributor to fouling problems in geothermal equipment and in order to obtain possible corrosion inhibition benefits in the carbon steel vessels and piping. The following salts were added to approximately 500 gallons of fresh water:

<u>Salt</u>	<u>Lbs. Added</u>	<u>ppm in 500 gal.</u>
NaCl	700	51,200 as Na
CaCl ₂ • 2H ₂ O	500	25,300 as Ca
Na SiO ₃ • 5H ₂ O	2	106 as SiO ₂

Including 124,000 ppm chlorides, the first batch contained 20% solids by weight as mixed.

To prepare the brine, 500 gallons of fresh water was added to a 1400 gallon mixing tank. Next, 100 pound bags of chloride salts were emptied into the tank while an agitator was operating. When the sodium meta-silicate was added, the pH of the mixture increased from 7 to 10. Concentrated (32%) hydrochloric acid was used to bring the pH back to about 7. To completely dissolve the salts, it was necessary to run the agitator for about four hours. The sodium chloride, in the form of coarse rock salt, required the most time to dissolve. After mixing, the brine was transferred to a 500 gallon holding tank from which it was pumped to the test unit as needed.

Sodium meta-silicate dissolves readily in water to give a clear solution of high pH. Adjusting the pH of this solution with a strong acid like HCl consumes two moles of HCl per mole of sodium meta-silicate¹⁰. The meta-silicate is converted to silicic acid which forms a flocculent at room temperature that slowly settles out. It was necessary, therefore, to agitate the brine in the holding tank daily.

3.2.10 Brine Analysis

Heat transfer tests D-1 and E-1 through E-4 were run on the first batch of brine. Considerable dilution of the brine occurred in the test unit because the fresh water had not been completely drained out prior to adding the brine. The laboratory analysis of one sample taken from the test unit with batch one brine is given in Table 3-1.

After a week, the brine was drained from the unit and replaced by the second batch. This batch was the remaining 450 gallons of brine with the silica content increased to 500 ppm. Sodium meta-silicate was mixed separately with several gallons of water and neutralized with HCl before adding to the brine in the mixing tank. Tests E-5, E-6, E-7 and D-2 were run using this brine. About 430 ppm of HCO₃ and additional 80 ppm of SiO₂ were added to the brine for the third batch which was used for tests D-3, D-4, D-5 and E-8.

After the test unit was filled with each batch, the brine was circulated (but not heated) and a sample was taken for analysis. The results are shown in Table 3-1. Note that SiO₂ does not show up in the concentration originally added.

TABLE 3-1

Analysis of Brine During Operation (ppm)

<u>Item</u>	<u>Batch 1</u>	<u>Batch 2</u>	<u>Batch 3</u>
*Cl	73,300	142,700	141,400
*Na	30,200	58,700	58,200
Ca	14,300	27,300	24,800
Alk	24	51	24
SiO	96	148	120
TDS	118,800	231,200	229,200
pH	6.9	7.8	6.2

*Calculated from TDS

3.2.11 Effects of Brine

A primary concern regarding operation on brine rather than fresh water were those aspects influenced by interfacial tension, namely isobutane drop formation at the dispersion plate and coalescing of the drops. Since the addition of NaCl and CaCl₂ to water are known to increase surface tension, it was feared that larger isobutane drops might be formed at the dispersion plate and that these drops might tend to coalesce prematurely in the column. Fortunately, neither effect was evident during operation. The only adverse effect of the brine on preheater operation occurred a few days after the second batch of brine was added to the test unit. It was noticed that the dispersion was very non-uniform when the unit was first run at low isobutane flows. As flow increased, the dispersion became more uniform until, at about 6 gpm, it was essentially normal in appearance. Subsequent reduction of flow could be accomplished without the dispersion again becoming non-uniform. It therefore seemed that the problem might be caused by blockage of the dispersion plate orifices which was dislodged at higher orifice velocities. Inspection of the dispersion plate upon disassembly indicated that this was probably the case. It was covered with fine salt crystals except for a small area around each orifice. The salt crystals were soluble in fresh water and could be easily wiped away. FMC Corporation reported a similar problem with a spray tower heat exchanger operating on seawater.¹²

No unusual additional corrosion was noted due to operation on the simulated brine. The same powdery, black corrosion products were found that were present during operation on fresh water. No hard mineral scale was found in the preheater, boiler or brine heater.

3.3 TEST RESULTS - PREHEATER HEAT TRANSFER

3.3.1 Summary

Overall volumetric heat transfer coefficients of 3500 BTU/hr-ft³-°F were obtained in the liquid-liquid preheater. The volumetric heat transfer coefficient was influenced strongly by the isobutane (dispersed phase) flow and, to a lesser extent, by brine (continuous phase) flow. The number of transfer units for a given isobutane

flow ranged from 1.7 to 2.7. The highest values occurred at the highest brine to isobutane superficial velocity ratios. Heat transfer results for fresh water and simulated geothermal brine are comparable. Temperature and LMTD do not significantly influence preheater performance.

It is believed that axial mixing causes a significant deviation from ideal countercurrent flow, resulting in reduced thermal driving force for heat transfer. Overall preheater performance thus falls far below predictions based on local drop heat transfer coefficients and LMTD's. Two different axial mixing models indicate that mixing behavior has a more important influence on overall heat transfer performance than local coefficients. Experimental evidence regarding the effects of column length and diameter on overall performance and the temperature profiles of both phases is required to validate and refine a mixing model before spray columns can be reliably scaled up in size.

3.3.2 Data Reduction

Heat transfer surface in a direct contact heat exchanger is continuously formed and eliminated, making it difficult to define a heat transfer area that can be used to characterize overall heat exchanger performance. Reference to a conventional overall surface conductance, U , has therefore been avoided in favor of an overall volumetric heat transfer coefficient, U_v , defined as:

$$U_v = \frac{Q_f}{\Delta T_{LM}(v)} = \frac{UA}{v}$$

Where Q_f is the heat transferred to the isobutane per hour, ΔT_{LM} is the apparent log mean temperature difference between the two fluid streams and v is the operating volume of the vessel. The heat transferred to the isobutane was determined from isobutane flow rates and inlet and outlet temperatures. Brine is the continuous phase in the preheater and is in contact with the walls of the vessel. Heat is transferred from the brine to the ambient as well as to the isobutane, so Q_b is always somewhat larger than Q_f . The temperature difference used to calculate U_v is a logarithmic mean temperature difference (LMTD) based on inlet and outlet temperatures of the fluid streams and assumes countercurrent flow. This basis for calculating LMTD is used throughout this report.

Another parameter, the number of heat transfer units for counter flow, or NTU has also been used to express preheater performance. This parameter is defined as:

$$NTU = \frac{Q_f}{\Delta T_{LM}C_f} = \frac{UA}{C_f}$$

This is valid for C_f less than or equal to C_b , where C_f and C_b are the hourly heat capacities of the isobutane and brine streams, respectively. Heat transfer effectiveness can be determined directly from knowledge of NTU and the ratio of hourly heat capacities for a given flow arrangement (e.g. countercurrent). NTU may be taken as a dimensionless measure of the portion of the total working fluid temperature rise required for a binary cycle that can be economically provided in a particular heat exchanger. When scaling a direct contact heat exchanger up in size,

NTUs are probably more meaningful than volumetric heat transfer coefficients since axial mixing effects or drop coalescence may impose a limit on the NTU that can be obtained in a single contact. It should be noted that the number of transfer units is the same as the number of equilibrium stages if the ratio C_b/C_f is unity (i.e. the temperature change of each fluid is the same). Otherwise, the two terms are not equivalent. NTU has been used here because it is a convenient heat transfer correlation for this application.

Heat transfer tests were run with isobutane flows ranging from 500 to 3000 pounds per hour, preheater brine inlet temperatures to 221°F and brine flows from 1300 to 3200 pounds per hour. Preheater data were recorded during combined preheating and boiling tests, as well as preheating only tests. A large number of steady state operating points, about 130, were reduced. Test data for all runs are tabulated in Appendix A.

The method used to reduce preheater operating data is given in Appendix A. Physical properties of pure isobutane from reference (33) have been used in calculations. Brine properties were evaluated for aqueous sodium chloride solutions of equivalent total dissolved solids weight concentration from reference (33).

3.3.3 Heat Losses

Because of the fire hazard posed by the use of isobutane, the test unit was constructed out-of-doors. Testing was performed during the winter and spring, often under conditions of low ambient temperature, moderate winds and even rain. It is not surprising, therefore, that the heat lost by the brine exceeded the heat transferred to the isobutane by a considerable amount. This discrepancy was usually between 10 and 30% of the heat lost by the brine, the higher percentage losses occurring for tests with low Q_f . Heat transfer results have been based on Q_f determined from isobutane temperature measurements and flow rates. Brine temperatures have been used only to calculate mean temperature differences for heat transfer which should not be seriously in error. The variability of the heat loss does, however, interfere with steady state operation and probably causes some of the scatter in the data.

To see if the discrepancy in Q 's could be accounted for by losses to the atmosphere, a heat loss test was performed by circulating hot brine through the rig without isobutane flow at different steady temperatures. Heat losses for the boiler and preheater were determined as a function of temperature. An over-all U of about 2 BTU/hr ft²°F for the boiler and 3.5 BTU/hr.-ft²°F. for the preheater were obtained. The vessels were insulated with half inch thick plastic foam but some areas, such as the flanges, were left exposed. By correcting the Q_b 's for heat loss based on the above U and the difference between ambient and average brine temperature, the difference between Q_b and Q_f was reduced to $\pm 10\%$ for most runs.

3.3.4 Data Analysis

Volumetric heat transfer coefficients, U_v , are principally influenced by the interfacial area available for heat transfer. Over the narrow range of drop sizes encountered in these tests, interfacial area may be considered proportional to the dispersed phase hold-up. For figure 3-4 the hold-ups were not measured for each point but were taken as a

function of dispersed phase (isobutane) flow only from a curve drawn through experimental data. Hold-up is practically independent of continuous phase (brine) flow and varies only with isobutane flow, especially for the dispersed mode of preheater operation in which most of the heat transfer data were taken.

Point C-10-9 has been omitted from Figure 3-4 because the heat balance was grossly in error (300%) and the U_v was unusually low. The five values of U_v over 3500 were obtained after a new brine distributor was installed in the preheater and appears to reflect a slight improvement in heat transfer performance due to reduced flow disturbance in the upper part of the column.

Preheater heat transfer performance is not appreciably affected by average brine temperature over the range of approximately 100°F to 190°F covered by these tests. Data for simulated brine and fresh water runs also fall together despite considerable variation in physical properties due to temperature and concentration. Prandtl numbers of the continuous phase, for instance, vary by over 300%.

Although the U_v values have been plotted against hold-up (or as a function of the isobutane flow only), actually they show a slight positive dependence on the brine flow. This influence causes some of the scatter in figure 3-4, especially in the large groups of data around holdups of 20 and 30%. Figure 3-5 is a dimensionless representation of the trend. Preheater NTUs are plotted against the ratio of superficial phase velocities, V_b/V_f , for these hold-ups. Superficial velocity is defined as the volumetric flow per unit cross sectional area of the column. The trend of increasing NTU with brine flow at constant butane flow may be due to an influence of brine flow on holdup undetected in the experimental holdup determinations. Or the trend might be explained by the relationship of the V_b/V_f ratio to backmixing effects which is discussed later in this section. The points near 3 NTU for 30% hold-up were obtained after the modified brine distributor was fitted.

The brine-to-isobutane mass flow ratio for the conceptual design described in Section 4 is 1.192. The NTU for eight preheater tests using simulated brine with isobutane flows of about 2500 lbs/hr and mass flow ratios approximating the conceptual design is 2.5. This value was, therefore, used as a basis for the conceptual design although it is likely that higher isobutane superficial velocities would yield larger NTUs at the same flow ratio.

One approach to understanding the overall heat transfer performance of liquid/liquid spray towers is to examine the local film coefficients on the inside and outside of individual drops as they pass through the continuous phase. Heat transfer occurring during droplet formation has been neglected in the following discussion since drops are formed very quickly at the dispersion plate orifice. Inside the drops, a film coefficient corresponding to heat transfer by conduction only with no convection at all may be taken as a lower limit. An approximate Nusselt number for a rigid sphere given by Treybal⁵² is:

$$Nu_d = \frac{h_d}{k} = \frac{2\pi^2}{3} = 6.6$$

FIGURE 3-4

PREHEATER VOLUMETRIC HEAT TRANSFER COEFFICIENT (U_V)

VS

HOLDUP OF DISPERSED PHASE (H)

FIGURE 3-5

NUMBER OF HEAT TRANSFER UNITS (NTU)

vs

SUPERFICIAL VELOCITY RATIO (V_b/V_f)

HOLDUP
○ = 30%
● = 20%

It is known, however, that circulation occurs in drops moving through viscous fluids. Sideman and Shabtai⁴⁵ surveyed various equations for heat transfer inside drops based on different models of drop circulation. Kronig and Brinks⁶⁵ solution for a spherical drop with internal circulation gives a Nu_d value of about 17. Furthermore, at Reynolds numbers of practical interest (>200), drop oscillation begins and combines with internal circulation to increase mixing in the drop. The Handlos and Baron model⁶⁶ and the experiments of Johnson and Hamielec⁶⁷ show that heat transfer coefficients up to 70 times those of the solid drop model may be expected at high Reynolds numbers.

For the outside film, the droplet may again be approximated by a rigid sphere. The correlation recommended by McAdams³¹ for this case is

$$Nu_c = (0.97 + 0.68 Re^{0.5}) Pr^{0.3}$$

which gives a value for Nu_c of about 40 for the conditions of experimental preheater operation. Models that take into account the effect of internal drop circulation on the outside film coefficient predict Nu_c values two or three times higher. For rigid spheres most of the heat transfer at Reynolds numbers of interest occurs from the frontal half.⁶⁰ In contrast, several investigators have reported that heat transfer at the rear of a liquid drop predominates.^{61,62,63} The following correlation⁶⁴ for droplets larger than an eighth inch gives a Nu_c value of 68 BTU/hr-ft²-°F.

$$Nu_c = 0.42 (\Delta \rho g d^3 \rho_c / \mu_c^2)^{1/3} Pr_c^{1/3}$$

Note that a buoyancy term is included so that a separate velocity calculation is not required.

The average drop diameter in the preheater was about 0.15 inch at high isobutane flow rates. By evaluating inside and outside film coefficients using the above Nusselt numbers, a local heat transfer coefficient can be calculated. For inside and outside Nusselt numbers of 17 and 68, respectively, the following coefficients are obtained.

$$\text{Inside film} : h_d = 68 \text{ BTU/hr.-ft}^2 \text{ } ^\circ\text{F}$$

$$\text{Outside film} : h_c = 2100 \text{ BTU/hr-ft}^2 \text{ } ^\circ\text{F}$$

$$\text{Both films} : U = 60 \text{ BTU/hr-ft}^2 \text{ } ^\circ\text{F}$$

Sideman and Taitel⁴⁴ experimentally determined heat transfer coefficients for drops of liquid butane rising in stagnant distilled water and obtained values for U of about 60 BTU/hr-ft²-°F. (This is a unit surface conductance, not to be confused with the volumetric coefficient, U_v). For comparison, the volumetric heat transfer coefficient calculated from Test D-5-12 assuming countercurrent flow is 3438 BTU/hr-ft²-°F. With a holdup of 48%, and mean drop diameter of 0.15 inch, the interfacial area per cubic foot of operating preheater volume is 230 ft²/ft³, with a unit surface conductance, U , of only 15/BTU/hr-ft²-°F. This is only 25% of the value measured by Sideman and calculated above for individual drops. In another study, Sukhatme and Hurowitz⁵⁰, using oil

dispersed in water in a 3" diameter spray column, obtained U values of about 20 BTU/hr-ft²°F. at high hold-ups. Again, a U of about 60 BTU/hr-ft²°F might have been expected for the conditions of their experiments. This sort of discrepancy is usual for liquid-liquid spray column heat exchangers when the overall coefficients are calculated assuming true counter current flow and casts doubts on the validity of using a logarithmic mean temperature difference to calculate overall heat exchanger performance from local drop coefficients as is done for a conventional surface heat exchanger.

Ideally, in a column designed for countercurrent contact of two phases, each phase should be in "plug" flow to obtain the maximum theoretical driving force for heat transfer. Any deviation from plug flow caused by axial mixing tends to diminish the effective driving force. Then, the overall heat transfer coefficient for a heat exchanger, calculated by assuming plug flow, will be lower than the local coefficient. The extent of the reduction will depend on the deviation from plug flow which can range to the other extreme of perfect mixing. In order to adequately characterize heat exchanger performance for design purposes, axial mixing must be quantified and related to heat exchanger efficiency.

Sleicher⁴² developed an eddy diffusion model to relate mixing in liquid extraction columns to mass transfer performance. The model relates four dimensionless parameters: an eddy Peclet number for each phase; the number of transfer units; and the extraction factor (similar to the hourly heat capacity ratio for heat transfer). For each phase, the eddy Peclet number is defined as:

$$Pe = LV/\alpha$$

Where V is the superficial velocity of the phase, L is the column height and α is the eddy diffusivity of the phase. The Peclet number expresses the extent of axial mixing. The less the Peclet number, the greater the mixing. Difficulty arises, however, in trying to determine its magnitude. It is known generally that spray column operation is characterized by considerable axial mixing in the continuous phase but very little mixing in the dispersed phase.^{29,52}

Orcutt, Mixon and Whitaker³⁶ applied Sleicher's eddy diffusion model to heat transfer in a spray column and performed experiments to determine Peclet numbers through a radioactive tracer technique. They defined an average Peclet number as:

$$\bar{Pe} = PR/(R + PF)$$

Where P and R are the dispersed phase and continuous phase Peclet numbers, respectively, and $F = C_b/C_f$. For the case $F = 1$, they recommended a value for $\bar{Pe}$ of 2.1 for columns up to six inches in diameter. No definite conclusion was reached regarding the effect of column length and diameter on $\bar{Pe}$. Figure 3-6 is adapted from Orcutt and illustrates the manner in which this approach can be used to explain spray column heat exchanger performance. This figure is restricted to the case where

FIGURE 3-6

APPARENT HEAT TRANSFER UNITS

VS

ACTUAL HEAT TRANSFER UNITS

(ORCUTT ET AL., 1966)³⁵

$F=1$. This condition is approximately satisfied by preheater test D-5-12. In figure 3-6 apparent heat transfer units are the values of NTU calculated from overall heat exchanger performance based on the assumption of plug flow. Actual transfer units are the NTU calculated from the local heat transfer coefficients. For Test D-5-12 the apparent NTU is 2.2. For $Pe=2.1$, the actual NTU from the curve is 10. Therefore, the actual local coefficient predicted from the previously calculated apparent coefficient for D-5-12 of $15 \text{ BTU/hr ft}^2\text{-}^\circ\text{F}$ is $68 \text{ BTU/hr-ft}^2\text{-}^\circ\text{F}$ which is about the same measured by Sideman and predicted by the general correlations.

The other curves of figure 3-6 for infinite Pe and $Pe = 0$ are for plug flow and perfect mixing, respectively. This model suggests that for small Pe and actual NTU over about 20, the performance of the heat exchanger will be insensitive to the local heat transfer coefficients and surface area. Instead, it will depend on the axial mixing characteristics of the heat exchanger.

Mixing in the continuous phase of spray towers is believed to be largely caused by wakes of rising dispersed phase drops.⁸ Hendrix, Dave and Johnson¹⁵ found that significant portions of the continuous phase were conveyed in the wakes of droplets, and that some of the original material in the wake reached the top of the column, the amount decreasing with column height. Observations of temperature profiles in spray column heat exchangers led Letan and Kehat⁵⁸ to propose the following model for heat transfer based on the hydrodynamic behavior of drops. Formation of wakes behind rising drops removes a large portion of the continuous phase from the lower part of the column and causes a rapid increase in drop temperature as heat is transferred from the new wake material. As each drop rises, part of its wake is continually shed and replaced in the middle of the column. At the top, the drops coalesce and all of the wake material mixes with the incoming continuous phase. This causes a sudden decrease in continuous phase temperature which reduces the effective driving force for heat transfer. By estimating the volume of wakes and the rate of shedding from temperature profiles in a spray column, Letan and Kehat were able to successfully predict temperature profiles for several other studies with dispersed phases ranging from benzene to mercury.

Their model was applied to the example preheater test, D-5-12, used previously. From the isobutane inlet and brine outlet temperatures at the bottom of the preheater, the isobutane outlet temperature was predicted within an error of 12%. The only other variables for a given fluid system are flow rates, average hold-up and column length. Fully mixed drops and wakes are assumed in the model so knowledge of local heat transfer coefficients is not required. Table 3-2 summarizes the isobutane outlet temperature predictions for several other preheater tests.

TABLE 3-2
CALCULATION OF PREHEATER ISOBUTANE OUTLET
TEMPERATURES BY THE LETAN-KEHAT MODEL⁵⁸

Run No.	C-8-9	E-1-12	B-7-17	D-2-12	D-5-12
Holdup, %	20	20	30	30	48
Z, cm	168	168	183	183	183
R	1.57	1.62	1.80	1.70	2.50
r	0.339	0.357	0.335	0.367	0.351
M	0.83	0.83	0.83	0.83	0.83
m, cm ⁻¹	0.030	0.030	0.030	0.030	0.023
T _{f3} °F	138	136	107	100	123
T _{b3} °F	158	152	134	136	134
T _{f4} °F	182	163	160	172	149
T _{f4} (calc.) °F	173	165	158	170	146
% error	20	7	4	3	12

The ability of the Letan-Kehat model to predict spray column heat transfer performance indicates that the resistance to heat transfer at the drops is unimportant in comparison with the fluid dynamics of the system. Some other qualitative observations can be made regarding this model. The influence of the superficial velocity ratio, V_b/V_f , on the NTU value in figure 3-5 may be explained in terms of the sudden temperature drop which the continuous phase undergoes at the top of the preheater. The amount of the continuous phase transported to the top of the column in the wakes of the drops depends on the dispersed phase flow rate. As the ratio V_b/V_f increases, the magnitude of the temperature drop should decrease and the value of NTU should improve. Another important consequence on of the wake model is the effect of column length on mixing behavior. As column length decreases, the sudden temperature drop at the top becomes a greater percentage of the overall continuous phase temperature drop for the column and it approaches the behavior of a perfect mixer. As length increases, the reverse is true, and the temperature distribution in the column approaches a plug flow profile. This implies that the NTU should increase as column height increases and that axial mixing does not impose a "ceiling" on the NTU that can be obtained in a single spray tower.

If drop wakes are the primary mechanism by which mixing of the continuous phase occurs, the addition of longitudinal baffles to a large diameter spray tower should have no beneficial effect on the heat transfer performance. FMC Corporation found this to be the case in their work with 16 inch diameter columns.¹²

Since the sudden temperature drop which the continuous phase undergoes at the top of the column is said to be caused by dispersed phase drops coalescing and expelling wake material, a significant change in the mixing behavior of the spray column might be expected if, instead of coalescing, the drops were vaporized. This suggests that the heat transfer performance in the preheating section of a combination preheater-boiler might be significantly different than that of a similar preheater.

3.3.5 Isobutane Holdup

Holdup is the fraction of the active preheater volume occupied by the dispersed liquid phase, isobutane in this case. For a given isobutane drop size, the interfacial area for heat transfer is directly proportional in size to the holdup, so determinations of holdup are useful in characterizing heat exchanger performance. Holdup also assumes significance in the mixing behavior of spray towers.

Three different techniques were used to measure holdup in the preheater. The first method is based on the fact that the test unit brine loop is closed and therefore contains a fixed amount of brine. As the holdup in the preheater increases, brine is displaced from the preheater into the boiler and surge tank. By measuring the brine levels in these vessels, the displaced volume and holdup can be determined.

The first method proved cumbersome in practice so another method, used previously by others, was tried. It involves suddenly isolating the preheater by quickly closing the brine and isobutane outlets. Isobutane settles out in a layer above the brine and the volume between the settled interface and the operating interface is measured. Since this method requires disruption of stable operation, it was generally used when shutting the unit down after a heat transfer test.

A third method allows continuous monitoring of holdup by assuming that the pressure differential across the preheater measures the average "density" in the column. The neglect of wall friction in this case is justified by the low superficial brine velocity, about 200 ft/hr. Figure 3-7 illustrates the relation of manometer to preheater. By balancing pressures from A to B it can be shown that:

$$P_B = P_A + L_1 \rho_f + \rho_b (L_3 + L_4 + L_5 + L_6) - \rho_{1.75} (L_3 + L_4)$$

and

$$P_B = P_A + \rho_f L_7 + \rho_b L_8 + (L_2 - L_7 - L_8) (\rho_b (1-H) + \rho_f H)$$

where the subscript 1.75 denotes manometer fluid with specific gravity 1.75. Combining these expressions and inserting numerical values for the preheater (length in inches and density in pounds per cubic inch) gives the following expression for holdup, H:

$$H = \frac{4.371 - .0215(L_3 + L_4) \rho_f (27 - L_1 - \Delta L) - \rho_b (88 + L_1 + \Delta L)}{(73 + L_1 + \Delta L) (\rho_f - \rho_b)}$$

ρ_f and ρ_b are evaluated at preheater operating temperatures. L_7 is not a constant since the position of the brine-isobutane interface varies and must be recorded along with the manometer readings. Also, the apparent position of the interface as seen in the gage glass is not the true position in the column because the gage glass itself is a type of

FIGURE 3-7

PREHEATER MANOMETER

manometer and the level in the glass is influenced by holdup. Therefore, $L_7 = 27 - L_i + \Delta L$ where L_i is the indicated height of the interface (measured from the bottom of the gage glass) and ΔL is a correction which depends on holdup. This correction also applies to the "quick shut-off" method of measuring holdup.

The manometer proved somewhat erratic in practice, presumably due to failure to expell all gases from the pot in the upper leg. Adjustments required to maintain flows or preheater interface position interfered with measurements and sometimes blew the fluid out of the manometer. An improved manometer system is planned for future operations.

Experimental holdup values are tabulated in Table 3-3 and plotted in Figure 3-8. For comparison, holdups predicted by analogy with the Zenz correlation for fluidized beds^{5,52} are shown for two values of continuous phaseflow. The Zenz correlation, as applied to liquid-liquid systems, relates holdup and the ratio of slip velocity to terminal drop velocity. Slip velocity is the relative velocity between the continuous phase the the dispersed phase drops, and is given by

$$V_s = \frac{V_d}{H} + \frac{V_c}{1-H}$$

where V_d and V_c are the superficial velocities of the dispersed and continuous phases. An expression recommended in reference (52) was used to calculate a terminal velocity of 0.69 ft/sec. for a 0.15 inch diameter isobutane drop rising through the brine. Physical properties for typical test conditions were used in the calculations. This correlation predicts some dependence of holdup on continuous phase flow although none was observed experimentally, perhaps because of limitations in the accuracy of the methods used. The curves in Figure 3-10 are for the straight section of the tower whereas the experimentally determined holdups are for the volume between the dispersion plate and the upper interface.

For all preheater heat transfer tests through D-2, holdups were below 40%. This mode of spray column operation is generally referred to as "dispersed" because distances between drops are large and the drops do not interfere with each other appreciably.²⁶ A "close packed" mode was obtained in tests D-3, D-4 and D-5 by lowering the interface to the top of the straight section of the column. This reduces the area of the isobutane-brine interface and slows the rate at which drops coalesce, causing them to "pile up" below the interface. Close packing resulted in holdups greater than 40%.

At certain critical dispersed and continuous phase flows, determined by tower cross sectional area and physical properties of the liquids, the close packed drops extend down to the dispersed phase distributor and "rejection" occurs. This hydraulic limit was never reached during testing. As Figure 2-6 (section 2) indicates, pump capacities were not quite high enough to reach the flows for rejection predicted by the Minard and Johnson correlation.

3.3.6 Isobutane Drop Size

The size of the isobutane drops produced by the preheater dispersion plate was estimated from enlarged photographs similar to Figure 3-9.

FIGURE 3 - 8
ISOBUTANE HOLDUP
VS
ISOBUTANE FLOW

TABLE 3-3
PREHEATER ISOBUTANE HOLDUP

<u>RUN</u>	<u>GPM ISOBUTANE</u>	<u>GPM BRINE</u>	<u>HOLDUP %</u>
A-2	1.82	2.25	3.7
A-2	1.82	2.25	2.5
A-2	3.63	2.25	6.4
A-2	5.45	2.25	13.5
A-2	7.27	2.25	21.9
A-2	9.09	2.25	31.9
A-5	8.18	4.50	38.0
A-5	5.45	4.50	18.8
A-5	3.63	4.50	10.6
A-5	9.09	4.50	40.3
A-5	9.09	0.00	38.8
A-5	7.27	0.00	29.1
A-5	5.45	0.00	12.3
A-5	3.64	0.00	9.6
B-6	9.90	5.35	32.4
B-7	8.93	5.02	26.4
B-8	9.30	2.26	26.6
B-8	7.40	2.26	21.2
B-8	5.55	2.26	15.6
B-8	10.17	4.52	32.9
C-10	9.69	5.38	30.7
D-2	10.17	5.76	35.0
D-2	9.35	5.39	33.4
D-2	9.17	4.90	28.8
D-4	11.18	4.37	45.3
D-5	8.31	4.07	24.2
D-5	9.43	4.07	29.2
D-5	11.04	3.97	38.7
D-5	10.79	3.98	49.4
E-6	7.42	4.90	18.3

FIGURE 3-9
ISOBUTANE DISPERSION AT ORIFICE VELOCITY OF 1.1 FT/SEC.
(UP →)

From measurements of about 25 drops for each nozzle velocity, a Sauter mean diameter (volume-to-surface ratio) was calculated from the following relationship:

$$d_p = \frac{\sum n_i d_i^3}{\sum n_i d_i^2}$$

Based on the drop diameters and standard deviations tabulated below, it appears that orifice velocity of about 2.2 feet/second yields the best combination of drop size and uniformity. Observations also indicate that this velocity is high enough to keep the orifice edges free of salt crystals which tend to interfere with drop formation.

TABLE 3-4

ISOBUTANE DROP SIZE

<u>Orifice Velocity</u> <u>ft/sec.</u>	<u>Diameter, d_p</u> <u>in</u>	<u>Standard Deviation</u> <u>%</u>
0.5	0.11	28
1.1	0.13	31
1.6	0.11	37
2.2	0.11	23
2.7	0.15	33
3.2	0.15	30

The photographs from which drop sizes were estimated were taken during ambient temperature operation of the preheater with simulated geothermal brine. It is believed, based on observation, that drop sizes for operation on fresh water were not appreciably different from those of Table 3-4.

The isobutane dispersion plate assembly was removed from the preheater and flow tests were conducted to determine its pressure drop characteristics. Although the pressure drop required to form the dispersion is small, it is important if it must be supplied by the difference in density between brine and isobutane. Such is the case if preheating columns are to be vertically staged without interstage pumps. Knowledge of the pressure drop allows the depth of isobutane liquid beneath the dispersion plate to be computed.

The test was conducted by flowing fresh water through the dispersion plate assembly and collecting it in a known volume while timing with a stop watch. Head loss was measured with a water filled manometer and averaged 14 velocity heads as shown in figure 3-10.

FIGURE 3-10

ISOBUTANE DISPERSION PLATE ASSEMBLY PRESSURE DROP

VS

NOZZLE VELOCITY

3.4 TEST RESULTS - BOILER

3.4.1 Summary

The UA factors for boiling in the experimental boiler ranged between 3200 and 6800 BTU/hr-°F, averaging 4700 BTU/hr-°F for all boiling tests. As expected, boiling performance is far superior to liquid-liquid heat transfer in the preheater in terms of UA factors and volumetric heat transfer coefficients. Boiling performance was not appreciably influenced by flow rates of brine or isobutane or by the thermal driving force. The average approach temperature difference calculated for the boiling zone was 8.5°F. For reduced saturation pressures above 0.3, boiling UA factors decline with increasing pressures. The boiler exhibited an unexpected ability to preheat subcooled isobutane liquid with maximum UA factors, for preheating of over 5000 BTU/hr-°F. As much as 30°F superheat was added to the vapor in the boiler; UA factors for superheating were, however, relatively low. The amount of water collected in the isobutane condensate from the boiler was generally less than the expected amounts. Discrepancies in heat balances and saturation temperatures indicate that water collection was incomplete.

3.4.2 Data Reduction

All heat transfer in the preheater takes place between liquid brine and liquid isobutane drops. In contrast, heat transfer to isobutane in the boiler involves the following four mechanisms: (1) liquid brine to liquid isobutane; (2) liquid brine to boiling isobutane; (3) liquid brine to isobutane vapor; and (4) water vapor to isobutane vapor.

As in the preheater, the heat transfer surface is difficult to define. But whereas preheater volume is at least fixed, the volumes of the various heat transfer zones within the boiler depend on operating conditions and can not be estimated on visual observation. To avoid rather arbitrary assumptions of volumes, heat transfer performance has been expressed in terms of a product of an overall unit area conductance, U , and area, A , for each of the three isobutane heat transfer phase regimes: liquid; mixed phase (boiling); and vapor. For each of the three sections, a UA factor is determined from the relation

$$UA = \frac{Q_f}{\Delta T_{LM}}$$

where Q_f is the heat transfer to the isobutane and ΔT_{LM} is a logarithmic mean temperature difference. Use of this temperature difference involves the assumptions of counterflow operation and constant local coefficients over the range of local temperature difference.

Q_f for each section was determined from the isobutane heat balance rather than the brine. The sensible heat lost by the brine, Q_b , in the boiler is, in general, equal to five components:

$$Q_b = Q_{f,PH} + Q_{f,BL} + Q_{f,SH} + Q_L + Q_{wv}$$

Where $Q_{f,PH}$, $Q_{f,BL}$ and $Q_{f,SH}$ are the heat transfer to the isobutane for preheating the liquid, vaporization and superheating the vapor respectively, Q_L is the rate of heat loss to the ambient and Q_{wv} is the heat rate of the water vapor generated in the boiler.

A temperature versus heat transfer duty profile shown in Appendix A illustrates the manner in which the intermediate brine temperatures were determined from brine inlet and outlet temperatures. Temperatures T'_b and T''_b were determined simply by proportioning the brine temperature to drop in the boiler according to the fraction of the total heat transfer to the isobutane represented by each section (i.e. preheating, boiling and superheating). Implicit in this method is the assumption that brine heat loss to the atmosphere and water vapor is evenly distributed throughout the three sections.

The isobutane saturation pressure in the boiler was determined from the total boiler pressure corrected for water vapor pressure. During boiler operation water was drained from a sump in the line connecting the condenser with the hotwell at frequent intervals. From the amount of water measured, the mole-fraction and partial pressure of the water vapor in the gas leaving the boiler was computed.

With knowledge of the saturation temperature, the amount of preheat, latent heat and superheat added to the isobutane was calculated from the flow rate and inlet and outlet conditions. The method of data reduction is illustrated by example in Appendix A.

3.4.3 Boiler Data Analysis

Isobutane liquid entering the boiler usually requires additional heating before it can begin to boil. The extent of preheating that must be done in the boiler depends on the difference between isobutane inlet and saturation temperatures which can range from 0 to more than 20°F. Because of the expected high degree of mixing in the boiler it should be capable of providing only about one heat transfer unit of preheat. For many runs, however, the observed NTU exceeds 2. UA values for the preheating section of the boiler are given for each test in Table A-3. The maximum UA values are comparable to, and even exceed in several cases the best UA values for the preheater. (Preheater UA factors are the volumetric coefficients, U multiplied by the operating volume of 1.2 cu. ft). Corresponding volumetric heat transfer coefficients are much higher for preheating in the boiler since the operating volume is smaller. This unexpected ability of the boiler to efficiently perform preheating duty indicates that it may be advantageous to preheating heat transfer performance if boiling takes place in the same spray tower.

The heat transfer performance of the liquid-liquid preheater was shown in the previous section to be principally dependent on the flow rates of isobutane and, to a lesser extent, brine. In light of this and the

experimental results of others,^{6,56} flow rates might be expected to influence the UA factor in the boiling section of the boiler. UA values for the boiling section in Table A-1 show little if any dependence on either isobutane or brine flow rates, however. The boiler was operated to completely evaporate all isobutane at ratios of isobutane-to-brine mass flow ranging from 0.24 to 1.17. Since complete evaporation required large brine temperature changes at these flow ratios, the ΔT at the "hot" end was much larger than at the "cold" end of the boiling section. For a fixed cold end approach temperature and brine flow, the heat transferred in the boiling section and the LMTD are both functions of isobutane flow. Thus, the increase in LMTD tends to offset increases in Q_f in the ratio:

$$UA = \frac{Q_f}{\Delta T_{LM}}$$

It is also possible that the influence of the LMTD on U noted by Huang, et al.¹⁸ and Sideman and Taitel⁴⁴ tends to oppose the increase in heat transfer area, A, which should accompany increased isobutane flow. However, both of those studies indicated that, beyond 18°F, thermal driving force has little effect on U for butane drops evaporating in water. LMTD's in the present tests were generally greater than 18°F.

Figure 3-7 indicates that the boiling UA factor is influenced by the saturation pressure (and temperature). The UA factors for isobutane flows between 7.1 and 8.1 GPM ($\approx$ 2000 lbs/hr) have been plotted against the reduced boiling pressure. Extrapolation to critical pressure indicates that a UA of only 2600 BTU/hr-°F would be obtained in this boiler. A correlation for pool boiling of organic liquids at a heated metal surface indicates a decline in peak heat flux at reduced boiling pressures above 0.30. This has been explained in terms of a buoyancy factor ($\rho_f - \rho_g$), the difference between liquid and vapor densities, which becomes zero at the critical pressure. Similarly, the trend shown in Figure 3-11 may be due to the declining change in volume accompanying change in phase as the pressure increases.

By assigning an operating height to the boiling section, volumetric boiling coefficients can be calculated for comparison with the results of other studies. If the height is taken, somewhat arbitrarily, to be 24 inches, the volume in the boiling section is 0.40 cubic feet. For this assumed volume, volumetric boiling coefficients for this study range from 7900 to 16,900 BTU/hr-ft³-°F. In work with an n-pentane-water system in a spray column, Sideman and Gat⁶⁰ obtained volumetric boiling coefficients ranging from 6200 to 12,500 BTU/hr-ft³-°F. Blair, Boehm and Jacobs⁶ boiled refrigerant 113, which has a specific gravity of 1.55, by direct contact with water in a type of sparged boiler; volumetric coefficients ranged from 260 to 4261 BTU/hr-ft³-°F with approach temperatures of 25°F and up. Ganiaris, et al.¹³ used butene-1 and n-butane as refrigerants to freeze brackish and sea water in a large crystallizer that was both sparged and mechanically agitated. Volumetric coefficients up to 2500 BTU/hr-ft³-°F for butene-1 and 3500 BTU/hr-ft³-°F for n-butane were reported. In this process, however, heat was transferred through two phase changes: refrigerant evaporation and water freezing. As might be expected, the coefficients from these tests compare best with those of Sideman and Gat whose tests with n-pentane were physically most similar.

FIGURE 3- 11
BOILING UA
 VS
REDUCED BOILING PRESSURE

The local temperature difference for boiling heat transfer in the present tests were as high as 147°F, which is higher than the above studies and beyond the range of studies of evaporation of single butane drops. Calculated cold end approach temperatures in the boiling section for all of the present test runs averaged 8.5°F. In general, approach temperatures increase with increasing isobutane to brine mass flow ratios. These fairly low approach temperatures appear to be typical of evaporation in systems of impure fluids although high liquid superheats have been found necessary to initiate boiling in high purity liquids. Thus, the impurities (i.e. n-butane and propane) in the isobutane used for these tests may have affected boiler performance by providing nucleation of vapor bubbles.

In most of the boiler tests, the isobutane vapor leaving the boiler was superheated to some extent. This superheat is added as the vapor passes through the hot brine spray at the top of the boiler. Contact time is quite short and the temperature driving force high so that the UA factor for superheating is low. For a given isobutane to brine mass flow ratio and isobutane saturation temperature, a certain minimum boiler brine inlet temperature is required to completely vaporize the isobutane. If the brine inlet temperature exceeds this minimum, superheat is added to the vapor.

3.4.4 Water Vapor Carryover

When hot brine is sprayed into the boiler some water vapor is produced at a temperature higher than that of the surrounding isobutane vapor. As the two vapors mix and approach a common temperature, the water vapor cools and any water vapor present in excess of the amount corresponding to the vapor pressure at the equilibrium temperature must condense. In so doing, additional heat is transferred to the isobutane vapor. At equilibrium the mole ratio of water vapor to isobutane vapor leaving the boiler should be equal to the ratio of the water vapor pressure at the mixed vapor exit temperature to the isobutane vapor pressure of the saturation temperature. At a boiling temperature of 220°F and 10°F of superheat, for instance, the water vapor should be 2.0% by weight of the vapor leaving the boiler.

As previously mentioned, water was collected from the condenser and measured. The corresponding water vapor pressure was subtracted from the total boiler pressure to arrive at the isobutane saturation temperature. The latent heat of vaporization of the water collected was also subtracted from the heat given up by the brine in the boiler for heat balance calculations. Figure 3-12 is a plot of the water collected expressed as a percentage of the amount of water vapor calculated based on boiler temperatures. The expected amount of water was collected only at very low isobutane flows, suggesting that all of the water was not trapped and measured at higher flows.

Two discrepancies in the temperature data also point to the possibility of incomplete water collection. First, the calculated saturation temperature based on measured amounts of water are a few degrees higher than the vapor exit temperatures actually measured for several tests. The other discrepancy involves the isobutane-brine heat balances in the boiler. The heat loss test, mentioned previously, explained the differences in brine heat loss and isobutane heat gain satisfactorily for the

FIGURE 3-12

preheater. For the boiler, however, part of the brine heat remains unaccounted for after deducting the heat transferred to the isobutane in the three sections (preheating, boiling and superheating), the heat loss to the atmosphere and the latent heat of vaporization of the water as measured. If the expected amount of water vapor is used for this last term instead of the measured amount, much better agreement is obtained in the heat balance.

The specific conductivity of a sample of water collected from the condenser during boiler operation on 21% brine was measured as a check on liquid brine carry-over in the vapor leaving the boiler. The measured value of 460 micromhos indicates that the solids carry-over amounts to about 230 ppm of water vapor or 5 ppm of the total vapor flow. Solids carryover may not be a problem with direct contact boiling since no flashing of the brine occurs.

3.5 ISOBUTANE LOSS

3.5.1 General

Brine in the test unit is continuously recirculated and becomes saturated with respect to isobutane. For a slightly soluble gas like isobutane, Henry's law should apply and the equilibrium concentration is given by:

$$S = \frac{P_f}{K}$$

where P is the partial pressure of the isobutane gas and K is the Henry's "constant" which is a function of temperature and brine salinity. It has been assumed in this section that the Henry's constants for a low pressure gas above liquid brine in reference(33) apply to the situation in which only the two liquid phases, brine and isobutane are present.

In addition to the isobutane in solution, small droplets of liquid isobutane may be entrained in the brine leaving the preheater. Since the brine is normally saturated with dissolved isobutane, any concentration in excess of equilibrium represents entrained isobutane liquid.

3.5.2 Methods of Determination

A Beckman Model 915 total organic carbon analyzer was available in the test facility laboratory for analysis of isobutane concentration, but its use was limited to low pressure brine samples. To determine the isobutane content of higher pressure brine samples (up to 340 psig) while the test unit was operating, the small stripping apparatus shown schematically in Figure 3-13 was devised. This apparatus was adapted for use at the test unit, requiring no electrical ancillaries such as vacuum pump.

The measurement is based on the pressure rise in a known volume caused by the release of isobutane gas from a measured quantity of brine. The volume of the stripping chamber (1) is small, $3.96 \times 10^{-3} \text{ ft}^3$ (112 cc), in the interests of obtaining a large rise in pressure from a brine sample of no more than one or two gallons. It was assumed that no gases other than isobutane were dissolved in the brine that would come out of the solution at one atmosphere pressure, i.e. the total pressure in the

stripping chamber was the sum of the partial pressures of isobutane and vapor. Further, it was assumed that the brine was stripped of isobutane to some constant fractional approach to the equilibrium concentration at the stripping chamber pressure and temperature.

Each concentration measurement is made in the following manner. While the direct contact heat exchanger test unit is in operation, the sample valves (A) are opened to admit brine to the stripping chamber (1) through the spray nozzle (2). The gas evolved is vented through (C). This is done to purge the chamber of air and clear the sample lines. After about 10 minutes, valves (A) and (C) are closed, leaving a slight positive pressure in the chamber. To take a sample, (C) is opened briefly to equalize the chamber at atmospheric pressure and one of the sample valves is opened. Sample flow is about 0.1 GPM. After a reduction in pressure from full test unit pressure to chamber pressure through (A) and the spray nozzle, the brine is drained through a loop seal (4). A slight amount of liquid is maintained in the bottom of the chamber by adjusting (B) as chamber pressure slowly rises. When a pressure of about 20 inches of mercury is reached the sample valve is closed and any brine in the bottom of the chamber drained off through (B) which is then closed. Final temperature and pressure of the chamber are noted.

To calculate the concentration of the sample, two more assumptions are made. First, it is assumed that Henry's law applies (i.e. the equilibrium concentration of isobutane in water at a given temperature is proportional to its partial pressure in the gas phase). Second, it is assumed that the ideal gas law can be used to calculate the chamber pressure. The resulting differential equation for the rate of chamber pressure increase with brine flow has the solution:

$$P_f = kS(1 - \exp(-RTBE/KV)) + P_{f0} \exp(-RTBE/kV)$$

where P_f = partial pressure of isobutane, psf

P_{f0} = initial partial pressure of isobutane, psf

S = concentration of isobutane in brine sample,
lb-moles/lbm brine

k = Henry's constant, psf/lb-mole/lbm brine

R = universal gas constant, 1545 ft-lbf/lb-mole-°R

B = mass of brine sample, lbm

T = temperature, °R

V = chamber volume, cu.ft.

E = fractional approach to equilibrium defined as $(S - S_a)/(S - S_e)$
where subscripts a and e denote actual and equilibrium
concentrations at chamber conditions.

FIGURE 3-13
ISOBUTANE CONCENTRATION
MEASURING DEVICE

Solubility data from reference (33) has been converted to values of Henry's constant and is given in Figure B-1. It is estimated that the accuracy of this experimental method is $\pm 10\%$ if the value of E is accurately known. Huang, et.al. obtained values for E of 0.80 to 0.99 stripping butane from distilled water in a similar chamber. It was originally intended to determine E experimentally by checking the isobutane content of the drains from the chamber with the carbon analyzer but this method was discarded for lack of an adequate sampling technique. Instead, E was determined by a few trial runs with the stripping apparatus using a saturated sample of known concentration. A value for E of 0.95 yielded results within the estimated experimental accuracy and was used for subsequent runs.

3.5.3 Results

Data for fresh water are shown in Figure 3-14. These points were obtained during operation of the preheater only for various isobutane and brine flows. As expected, the concentration was generally at about the saturation level corresponding to conditions at the preheater brine outlet. Carry-under of isobutane droplets, it appears, occurs only randomly. No influence of variations in isobutane or brine flows in the normal operating range was noted. The samples were taken at isobutane flows from 2 to 12 gpm and brine flows from 0 to 5 gpm.

When samples were taken while the test unit was operating in the combined preheating-boiling mode, no measurable increase in chamber pressure could be obtained, even for large water samples (i.e. several gallons). In preheater only operation, the water is not heated enough to cause the isobutane to come out of solution. During boiler operation, however, much of the isobutane is stripped from the water as it enters the boiler. Apparently, subsequent mass transfer taking place in the preheater is insufficient to produce a concentration detectable by this stripping method.

It was noticed that when a fresh batch of water was added to the test unit, equilibrium concentrations of isobutane were not obtained until the preheater had been operated for an hour or two. These observations suggest that the concentration in the exit brine might be much lower than equilibrium for once-through brine flow.

Samples of simulated geothermal brine produced no measurable pressure rise in the stripping chamber even for preheater only tests. As the dashed lines in Figure 3-14 for sodium chloride solutions indicate, equilibrium concentrations are much less than for fresh water due to a "salting" out effect. The isobutane concentration in the simulated geothermal brine was apparently too low to be measured by this technique.

To draw quantitative conclusions regarding the magnitude of working fluid loss to exit brine, tests using once-through brine flow and sampling techniques capable of detecting smaller isobutane concentrations will be necessary. In particular, such tests would allow the relationship between the heat and mass transfer performance of direct contact heat exchangers to be explored. This relationship may

FIGURE 3-14
ISOBUTANE CONCENTRATION

VERSUS
PREHEATER WATER OUTLET TEMPERATURE
(FRESH WATER DATA)

have significant impact on the optimization of binary fluid cycles employing direct contact heat transfer.

Another test was run using the stripping apparatus to determine the effectiveness of the liquid-liquid separator. An emulsion was deliberately produced by allowing both isobutane and brine to be simultaneously drawn through the same pump. If allowed to stand, such an emulsion requires several hours to settle out. The emulsion was circulated while samples were taken before and after the separator. One pair of samples showed an inlet concentration of 3100 ppm and 1500 ppm at the outlet. A second set, taken later, showed 1300 ppm in and 1000 out. Thus, it appears that the coalescing pad of stainless steel mesh and fiberglass was not especially effective in coping with large concentrations of very finely divided isobutane. A coalescer made of an oleophilic material might improve separator performance.

3.6 NOTATION FOR SECTION 3

A	=	interfacial heat transfer area (ft ²)
B	=	brine mass, (lbm)
C _b	=	brine hourly heat capacity, (Btu/hr°F)
C _f	=	isobutane hourly heat capacity (Btu/hr°F)
cp _b	=	brine specific heat, (Btu/lbm°F)
cp _f	=	isobutane specific heat, (Btu/lbm°F)
d	=	diameter, (ft.)
d _p	=	drop diameter, (in)
E	=	fractional approach to equilibrium
F	=	C _b /C _f
g	=	acceleration due to gravity, (ft./hr ²)
H	=	dispersed phase holdup
h	=	liquid head, (ft.)
h _c	=	continuous phase film coefficient, (Btu/hr. ft. ² °F)
h _d	=	dispersed phase film coefficient, (Btu/hr. ft. ² °F)
K	=	Henry's constant, (psf/lb-mole/lbm brine)
k	=	thermal conductivity, (Btu/hr. ft.°F)
L	=	length, (ft.)
LMTD	=	log mean temperature difference, (°F)
M _b	=	brine mass flow, (lbm/hr)
M _f	=	isobutane mass flow, (lbm/hr)
M	=	ratio of wake to drop volumes
m	=	volume of wake elements shed per volume of drop and unit length of column, (cm ⁻¹)
NTU	=	number of heat transfer units
P	=	dispersed phase Peclet number
P _f	=	partial pressure of isobutane, (psf)
P _{sat}	=	saturation pressure of isobutane, (psia)
Pe	=	average peclet number for both phases.
Pr	=	Prandtl number
Q	=	rate of heat flow, (Btu/hr.)
R	=	continuous phase Peclet number
R	=	universal gas constant, (1545 ft lbf/lb-mole°R)
Re	=	Reynolds number
S	=	concentration of isobutane in brine, (lb-moles/lbm)
T	=	Temperature, (°F)
U	=	overall unit surface heat conductance, Btu/hr. ft. ² °F)
U _v	=	volumetric heat transfer coefficient, (Btu/hr. ft. ³ °F)
V	=	velocity, (ft/sec)
v	=	volume, (ft. ³)
X _i	=	partial pressure of H ₂ O as % of total pressure
z	=	column length, (cm)
α	=	eddy thermal diffusivity, (ft ² /sec)
μ	=	absolute viscosity, (lbm/ft sec)
ρ	=	density, (lbm/ft ³)

SUBSCRIPTS

1	=	boiler hot end
3	=	preheater cold end
4	=	preheater hot end
5	=	boiler cold end
a	=	actual (isobutane concentration)
b	=	brine
c	=	continuous phase
d	=	dispersed phase
e	=	equilibrium (isobutane concentration)
f	=	isobutane, or liquid
g	=	gas
i	=	indicated, or "i 'th" term
BL	=	boiling
PH	=	preheating
SH	=	superheating
LM	=	logarithmic mean
sat	=	saturation

SECTION 4

CONCEPTUAL DESIGN & COST ESTIMATE

4.1 CONCEPTUAL DESIGN

4.1.1 Design Basis

Preliminary designs of a direct contact heat exchanger system for a 50 MWe geothermal power plant were prepared based on the information generated by our analytical studies and the test results from the experimental direct contact heat exchangers. The assumptions and ground rules used for the preparation of the conceptual design are as follows:

- a. The conceptual heat exchange system should be constrained to a subcritical isobutane cycle operating with a moderate temperature, high salinity brine. The system design would not, at this time, be aimed at any specific fluid characteristics such as the content of particular scale formers or non-condensable gases.
- b. The design should be executed without considering the cost related to a surface type heat exchanger since surface units may not work at any cost with high salinity brines. However, the design would be aimed at minimizing production cost per Kilowatt hour.
- c. The conceptual design should be based on performance data obtained from the test unit. The performance improvements anticipated by proposed modifications to the design and operation of the test unit will not be incorporated into the conceptual design until they are confirmed by actual testing. This ground rule will tend to make the design somewhat conservative and perhaps counterbalance the uncertainties inherent to increasing the size of the heat exchanger by a factor of 16. (from a six inch diameter test unit to an eight foot conceptual design).
- d. Specific technical input to the power plant model used for the conceptual design is summarized in Table 4-1.

The design philosophy of the conceptual study is to develop a design with the lowest possible technical risk by utilizing state-of-the-art components where possible and by minimizing extrapolations in performance data and other design parameters for the direct contact heat exchanger. It is recognized that the concept of direct contact heat exchange is novel in itself so that every effort will be made to maintain a conservative approach. Because of this, it was further recognized that the conceptual design may not be the "lowest cost" answer to the direct contact heat exchange approach. It should be kept in mind, however, that a number of technical improvements promise to reduce the costs developed from this conceptual study. These improvements are discussed later in the next section in connection with the recommended future development and testing of direct contact heat exchangers.

TABLE 4-1

INPUT TO POWER PLANT/HEAT EXCHANGER MODEL

Plant Size:	50,000 KWe net
Geothermal Fluid	325°F, 210,000 TDS, $C_p(\text{avg}) = .817$
Working Fluid:	Commercial Grade Isobutane
Type of Cycle:	Subcritical, Without Superheat
Type of Preheater:	Direct Contact - Spray Tower
Type of Boiler:	Direct Contact - Spray Tower
Preheater Performance:	2.5 NTU, 3500 BTU/HR-°F-Ft ³
Boiler Performance:	10,000 BTU/HR-°F-Ft ³ (avg.)
Cooling Water:	80°F
Condensing Temperature:	110°F
Condensing Type:	Direct Contact
Turbine Efficiency:	.85
Pump Efficiency;	.80

The development of the conceptual design was initiated by establishing a thermodynamic model of the isobutane binary cycle which incorporates direct contact heat exchangers and brine booster pumps. The thermodynamic model differs slightly from other published models as follows: (a) Brine Booster pumps are included as a parasitic loss; (b) The boiler pressure is slightly higher than the saturation pressure of the isobutane by an amount equal to the water vapor pressures; (c) The brine is 210,000 ppm total dissolved solids and has an average specific heat of .817; (d) the thermodynamic model was constrained to work with one, two, or three stages of a direct contact preheater each having a size of 2.5 NTU's. In determining the cycle state points and system parameters the effect of the water vapor in the vapor leaving the boiler was not included. This is discussed in Appendix D.

For each preheater the log mean temperature difference required was determined by dividing the isobutane ΔT for that preheater by 2.5 for each boiling temperature. The total isobutane ΔT is simply the boiling temperature minus the hot well temperature, $T_{sat}-105^{\circ}\text{F}$. From this, the brine temperature profile was adjusted until the total isobutane preheat was attained in exactly 6 NTU. An equivalent of one NTU was assumed for preheating in the boiler, thus a total of 6 NTU was available for the entire preheating with two stages of preheaters at 2.5 NTU each.

The design process was repeated for isobutane saturation temperatures of 210°F to 260°F in 10°F increments, and a net power output per pound of brine was calculated at each temperature. It soon became apparent that with direct contact preheaters each having 2.5 NTU, two stages was most practical. The results of this analysis showed that the effectiveness (BTU Network/lb brine) was nearly constant for all temperatures, varying from 14.44 to 14.76. Table 4-2 and Figure 4-1 summarize the results.

Figure 4-2 illustrates the final temperature-enthalpy diagram for the conceptual design. This does not show the effects of the water vapor (see Appendix D). The temperature-enthalpy line for the brine is virtually linear at a total dissolved content of 210,000 ppm. Also the "pinch point" ΔT does not coincide with the saturation temperature at saturation temperatures above 230°F due to the convex curvature of the isobutane saturated liquid line. The graphical trial and error procedure for establishing the brine temperature profile and flow rate required first choosing a pinch point ΔT . This established a straight line for the brine profile from the brine inlet temperature through the pinch point to the brine outlet. The brine flow rate per pound of isobutane is then simply the ratio of the total isobutane enthalpy change to the total brine enthalpy change. It should be noted that the graphical method was for determination of inlet and outlet temperatures of each section which were used in calculating the MTDs. It was not graphical integration for determining actual MTDs.

The two preheaters were then fitted to the temperature profiles thus established to see if the given isobutane ΔT could be attained with two preheaters of 2.5 NTU size. If not, a slightly larger "pinch point" ΔT was chosen and the process repeated. It turns out that for each 10°F increase in the total isobutane ΔT (10°F rise in boiling temperature) an increase of roughly 1°F was required in pinch point ΔT to allow the preheating to be accomplished in two preheaters.

TABLE 4-2
CONCEPTUAL DESIGN STUDY
THERMODYNAMIC CYCLE SUMMARY

PARAMETER	BOILING TEMPERATURE (°F)					
	210	220	230	240*	250	260
Hf ₁	165.7	167.0	168.2	169.5	170.7	172.0
Hf ₂	143.0	143.0	143.0	143.0	143.0	143.0
Hf ₃	1.5	1.5	1.5	1.5	1.5	1.5
Hf ₅	72.0	80.5	89.5	99.5	109.0	120.0
Hb ₁	244.0	244.0	244.0	244.0	244.0	244.0
Hb ₃	93.0	96.0	100.0	103	108.0	112.0
Mf/Mb	0.923	0.894	0.865	0.839	0.802	0.744
Wt	19.3	20.4	21.42	22.52	23.55	24.65
P ₁	296.0	331.0	368.0	407	450.0	497.0
Wpf	1.48	1.72	1.97	2.24	2.54	2.87
Wpb	0.66	0.82	0.99	1.18	1.42	1.68
Mcw/Mf	5.18	5.20	5.20	5.21	5.22	5.22
Wpcw=.291 Mcw/Mf	1.51	1.51	1.51	1.51	1.52	1.52
Wp total	3.65	4.05	4.47	4.93	5.48	6.07
E =	<u>14.44</u>	<u>14.62</u>	<u>14.66</u>	<u>14.76</u>	<u>14.49</u>	<u>14.38</u>
ΔTp =	10.0	11.0	12.0	13.0	14.0	15.0
T _{B3} =	140.0	144.0	148.0	152.0	159.0	164.0

* Conceptual design basis

FIGURE 4-1

SYSTEM PARAMETERS
VS
BOILING TEMPERATURE

FIGURE 4-2
CONCEPTUAL DESIGN

ISOBUTANE CYCLE STATE POINTS

TEMPERATURE - °F

ISOBUTANE ENTHALPY BTU/LB

The use of one and three preheaters was investigated also, but neither offered any advantage. The single unit would require a very large LMTD and a corresponding high brine rejection temperature with low cycle effectiveness. The use of three preheaters lowered the LMTD only slightly and the slight gain in cycle effectiveness did not warrant the expense of an additional preheater.

Table 4-3 lists the thermal design parameters for the selected cycle corresponding to Figure 4-2. A boiling temperature of 240°F was selected because it has the highest effectiveness (14.76 BTU/LB Brine). The mechanical and hydraulic design is based on this cycle as well as the cost estimates. In retrospect, it appears that selecting a boiling temperature of 220°F may lead to a lower net cost per KWH. The boiler pressure at 220°F is 331 psia whereas at 240°F the boiler pressure is 407 psia. The pressure vessels for 331 psia could be as much as 10% cheaper than at 407 psia, and the brine pumps and pumping power would also be less. The accuracy of the cost and performance factors at this stage, however, do not warrant such a refinement or optimization routine.

The total height of each preheater was fixed by the height of the effective control zone. This zone is illustrated by the darker dispersion pattern in drawing 051-2206 (Figure 4-3). The height of the contact zone was set at eight feet or two feet higher than the equivalent height in the experimental unit. The extra two feet is a safety factor based on mere intuition. The remaining vertical distances are taken up by the conical ends, coalescing zones, and distributor plates as shown on the drawing. The height of the boiler was determined by arbitrarily designating the boiling volumetric coefficient at 10,000 BTU/HR - °F-ft³ and the preheating (occurring in the boiler) coefficient at 3500 BTU/HR - °F-ft³. The preheating and boiling therefore require 7.8 feet of height. An additional 12 feet was added for vapor space and demisters.

4.1.3 Hydraulic Design

The diameter of the contacting zone in the preheaters is sized by the superficial velocities of the brine and isobutane required to produce the desired holdup at the required flow ratio. The flow ratio, M_f/M_b , is set by the heat balance of the cycle. The superficial velocities were estimated by Minard and Johnson's relationship (see Figure 2-6) to be 550 cubic feet per hour per square foot of tower cross sectional area for isobutane. The same drop-let diameter as used, and observed in the experimental unit, .15 inch, was used for evaluating the Minard and Johnson relationship. The superficial velocities are somewhat higher for the conceptual design because of the higher average fluid temperatures. The superficial velocity for isobutane of 550 cu ft/hr-ft² was evaluated at an average fluid density of 29.5 LB/ft³, so that the corresponding volumetric flow rate for isobutane is 329,984 cubic feet per hour. The total tower area is then calculated by dividing the volumetric flow rate by the superficial velocity. Therefore, the area is $329,984/550 = 600$ sq.ft. and for ten vessels, each must have 60 sq.ft. cross-sectional area and have a diameter of 8.75 feet.

TABLE 4-3

CONCEPTUAL GEOTHERMAL BRINE/ISOBUTANE HEAT EXCHANGERS

<u>ISOBUTANE</u>	<u>PREHEATER #1</u>	<u>PREHEATER #2</u>	<u>BOILER</u>
Inlet Temperature, °F	105.0	185.5	227.2
Inlet Pressure, psia	428.0	421.0	412.0
Inlet Enthalpy, BTU/lb.	1.5	53.0	87.0
Outlet Temperature, °F	185.5	227.2	241.0
Outlet Pressure, psia	421.0	412.0	407.0
Outlet Enthalpy, BTU/lb.	53.0	87.0	169.5
 <u>GEOTHERMAL FLUID</u>			
Inlet Temperature, °F	206.5	240.0	325.0
Inlet Pressure, psia	421.0	412.0	410.0
Inlet Enthalpy, BTU/lb.	146.3	174.8	244.0
Outlet Temperature, °F	152.0	206.5	240.0
Outlet Pressure, psia	429.0	421.0	412.0
Outlet Enthalpy, BTU/lb.	103.1	146.3	174.8
 <u>SUMMARY - H.X.</u>			
Total Heat Transfer, 10 ⁶ BTU/HR	499.63	329.85	800.38
Percent of Total Heat, %	31%	20%	49%
Volumetric H.T. Coefficient,	3500	3500	
Boiling (avg), BTU/HR-°F-Ft ³			10,000
Preheating, BTU/HR-°F-Ft ³			3500
LMTD, °F	32.3	16.7	38.3
Isobutane ΔT, °F	80.5	41.7	13.8
Geothermal ΔT, °F	54.5	33.5	85.0
Effective Volume Required, Ft ³	4420	5650	2,600(Preheat) 2,090(Boiling)
Volume Supplied - Ft ³	9944	9944	9944
NTU	2.5	2.5	1.0 (Preheat)

TYPICAL CROSS SECTION

SCALE: 1/8" = 1'-0"

D&S ENGINEERS, Inc.	
PROJECT NO. _____	
DATE _____	
DRAWN BY _____	
CHECKED BY _____	
APPROVED BY _____	
CONCEPTUAL DESIGN	
BRINE / ISOBUTANE	
HEAT EXCHANGERS	
FOR 50 MWE SYSTEM	
FIGURE 4-3	

Referring to the drawing of the conceptual heat exchanger, figure 4-3 the reader will notice that 8' - 10" is shown for the diameter of the boiler as well as the effective contact zone in the preheater.

The pressure vessel itself is not 8' - 10", but 12' - 5". The 8'-10" diameter contact zone is provided by an internal thin wall baffle. The pressure vessel diameter of 12' - 5" was set by the inlet and outlet conical ends of the preheater. This configuration is the conventional Elgin Spray Tower design which was also used for the experimental preheater, and which we know works satisfactorily. However, for the large preheaters of the conceptual design, it would be desirable to eliminate or reduce these large diameter conical ends since they cause the pressure vessel diameter to be nearly 40% greater than the effective direct contact zone (12.4 ft. vs. 8.8 ft). The conical ends provide a quiescent zone where the brine is disengaged from the bottom of the preheater stage without carrying under droplets of isobutane, and also to allow the brine to flow around the intermediate isobutane distributor plates. The conical ends also provide stability to the position of the flooded zone of the dispersion. The 12.4 foot diameter is quite generous and could undoubtedly be reduced, but we feel that these conical ends may ultimately be entirely eliminated. Spray towers have been built and operated without the conical ends, but these have had problems with controlling the stability of the interface between the dispersed and continuous phase. The tower operated by FMC was straight-sided, (without conical ends), and by means of an elaborate interface control system they were able to operate the spray tower successfully. Since the elimination of the conical ends would reduce the cost of the towers considerably, we recommend investigating this possibility in the future.

4.1.4 Design Summary

Table 4-4 summarizes the main features of the power plant used as a model for the conceptual design. The net output of 14.76 BTU/LB geothermal brine (effectiveness) is based on a brine having 21% total dissolved solids and an average specific heat of .817. To estimate the effectiveness for a brine with higher specific heat, the effectiveness must be factored upward by the ratio of the higher specific heat to .817. For example, the same plant operating with fresh water at the same temperature would have an effectiveness of approximately $1.0/.817 \times 14.76 = 18.07$ BTU/LB "Brine". For both cases, the brine inlet temperature is constant at 325°F. A higher brine inlet temperature will yield a higher effectiveness.

4.1.5 Direct Contact Condenser

Table 4-5 summarizes the condenser requirements for the conceptual design. A direct contact condenser was assumed for this service to avoid the installation of surface heat exchangers in an otherwise "direct contact heat exchange" system. While this is not part of this development program, we feel that the heat rejection service may be a greater problem economically than the brine to working fluid heat exchangers. Since the efficiency of geothermal systems is quite low, the amount of heat rejected is approximately 85% of the heat

TABLE 4-4
50 MW GEOTHERMAL POWER PLANT
CONCEPTUAL DESIGN SUMMARY

ISOBUTANE FLOW:	.839 LBS/(LB Geothermal Fluid) 9,701,535 LBS/HR 36,540 GPM @ 105° F
GEOTHERMAL FLUID FLOW:	1.192 LBS/LB Isobutane 11,563,200 LBS/HR 21,960 GPM @ 325°F
COOLING WATER FLOW:	5.21 LBS/LB Isobutane 50,533,000 LBS/HR 101,300 GPM @ 80°F
GROSS POWER OUTPUT:	22.52 BTU/LB Isobutane 64,010 KW
FEED PUMP:	2.24 BTU/LB Isobutane 6,370 KW
BRINE PUMP:	1.18 BTU/LB Isobutane 3,350 KW
COOLING WATER PUMP:	1.51 BTU/LB Isobutane 4,290 KW
TOTAL PARASITIC POWER:	4.93 BTU/LB Isobutane 13,990 KW
NET POWER OUTPUT:*	17.59 BTU/LB Isobutane 14.76 BTU/LB Geothermal Fluid 50,000 KWe 4,325 KWH/10 ⁶ LB Brine 231 LBS Brine/KWH
TOTAL HEAT EXTRACTED:	1630 x 10 ⁶ BTU/HR
TOTAL HEAT REJECTED:	1411 x 10 ⁶ BTU/HR

*Note: To estimate the net power per pound of geothermal fluid for a geothermal fluid having an average specific heat other than .817, multiply by the ratio of $C_{pnew}/.817$.

TABLE 4-5
DIRECT CONTACT CONDENSER SUMMARY

CONDENSER

Isobutane Inlet Temperature	130°F (20°F S.H.)
Condensing Temperature	110°F
Condenser Pressure	83 psia
Isobutane Outlet Temperature	105°F (5° Subcooling)
Cooling Water Inlet Temperature	80°F
Cooling Water Outlet Temperature	108°F
Isobutane Inlet Enthalpy	146.98 BTU/lb
Isobutane Outlet Enthalpy	1.50 BTU/lb
Cooling Water Inlet Enthalpy	48.02 BTU/lb
Cooling Water Outlet Enthalpy	75.95 BTU/lb
Cooling Water Flow	5.21 LB/LB isobutane 50.533 x 10 ⁶ LB/HR 101,300 gpm @ 80°F
Total Heat Transferred	1425 x 10 ⁶ BTU/HR

The condenser pressure is sufficient to transport the cooling water to the cooling tower without pumps. The pumping power for cooling water shown in the plant summary is for returning the cooling water from atmospheric pressure to condenser pressure plus piping losses.

input to the power plant. However, the heat rejection temperature difference is smaller than that of the heat input service. As a result, the amount of heat transfer surface required is very large. A direct contact condenser for this service could possibly create large capital cost savings. The condenser would operate by spraying cooling water into a chamber and the isobutane would condense directly on the droplets of water. The water and isobutane, due to the large density difference, would separate rapidly in the hot well. The cooling water would be withdrawn from the bottom of the hot well and sent to a cooling tower. The problem with this scheme is that if a conventional wet cooling tower was used the isobutane loss to the atmosphere might be excessive. The isobutane is ten times more soluble in fresh cooling water than brine, and the cooling water flow is five times that of the geothermal brine. Therefore, the isobutane losses may be as much as 50 times higher than expected in the direct contact brine to isobutane heat exchangers. The only solution to this problem is the possible use of a low cost plastic heat exchanger instead of a wet cooling tower. Thin walled plastic heat exchangers could reject the heat from the cooling water to the atmosphere without losing isobutane. Cooling water make-up needed by a wet tower would also be eliminated. The question may arise as to why the isobutane would not be condensed directly in the plastic air-cooled heat exchangers. The objection is that the condensing pressure of 83 psia and temperature of 105°F would require the plastic walls to be made thicker and may obviate the economic advantage of the concept. Also, isobutane would require a more expensive plastic since it softens many of the common low cost plastics. The use of atmospheric pressure cooling water would allow the use of thin low cost plastics.

The use of this system for heat rejection results in cooling water pump work about 8 times greater than that for a conventional surface heat exchanger using available cool water on a once-through basis. This larger pump work does not effect the previous analysis and determination of the thermodynamic optimum. However, it does show that the type of heat rejection system must be considered in overall economic optimization studies.

4.2 SYSTEM COSTS

4.2.1 Introduction

The purpose of this section is to provide the reader with the cost factors inherent to the direct contact brine to isobutane heat exchange system. The absolute and relative costs compared to a surface heat exchanger are summarized for a conceptual 50,000 KW system and the impact of technological improvements on the costs are discussed.

4.2.2 Method and Basis

The 50MW conceptual design used for cost estimating is based on the performance parameters obtained from the small experimental test unit operated during this contract. The thermal performance assumed is therefore conservative, since no credit has been taken for anticipated improvements to the test performance.

Also, the conceptual design has not been optimized by an integrated performance/cost model for minimum cost.

For the operating cost estimate, an annual average fixed charge of 13% has been used. This is typical for a public utility with federal support for financing. The sensitivity of the unit production costs caused by varying the fixed charge rate is also considered.

The capital cost estimate includes only costs associated with the brine/isobutane heat exchange system. The battery limits for the sake of our analysis therefore extends from the discharge of the isobutane feed pump to the turbine inlet. Three major items included in the capital costs are:

- (a) Direct Contact Heat Exchanger Vessels
- (b) Brine Booster Pumps. These are considered as part of the direct contact heat exchange system because they may not be required for a surface heat exchange system.
- (c) Isobutane Storage. Likewise this item results from the use of direct contact heat exchangers by virtue of increased isobutane losses.

Costs for installation of the above equipment include materials and labor costs for foundations, piping, structural supports, electrical equipment, instrumentation and controls. Insurance, freight, taxes and construction indirect costs are also included.

All costs are based on 1976 dollars.

4.2.3 Heat Exchange Towers

Table 4-6 summarizes the costs for the direct contact heat exchanger vessels. The vessel cost was estimated by taking the weight of steel from the vessel drawing, (No. 051-2206, figure 4-3) and applying a unit cost of \$.90/lb steel. The cost of fabricating the three isobutane distributor plates was added to this.

4.2.4 Brine Pumps

The cost of the brine pumps was based on using ten pumps of equal capacity. The total head assumes a suction pressure of 120 psia or about 20 psi above the saturation pressure of the brine at 325°F. The discharge pressure is equal to the boiler pressure plus the hydrostatic head of 60 feet and 10 psi for piping losses. The boiler pressure, as mentioned earlier, is taken as the isobutane saturation pressure plus the water vapor pressure at the isobutane boiling temperature. The electric motors specified are explosion proof which increases the cost of the motors by about 20%. The pumps specified have cast iron cases and stainless steel impellers and fittings, which is an absolute minimum from a materials standpoint. The use of better materials may be necessary depending on the composition of the brine and this would increase the total installed cost for the brine pump and installation by about 10%. See Table 4-7 for the cost breakdown.

Note that the actual power requirement for the brine pumps selected is 4020 KW compared to the theoretical power of 3350 KW given in Table 4-4.

TABLE 4-6

CAPITAL COST ESTIMATE

HEAT EXCHANGER TOWERS

Tower Size:	5 MW (net) each
Number of Towers:	10
Total Weight of Steel:	1,342,340 lbs
Unit Cost:	\$.90/lb Steel*
Vessel Cost:	\$1,208,100
Extra Cost of Internals:	<u>\$ 60,000</u>
Total Equipment cost, FOB	\$1,268,100
Installation Materials	\$1,302,338
Direct Field Labor	\$1,232,593
Freight, Insurance, Taxes	\$ 101,448
Indirect Costs	<u>\$1,254,151</u>
Total Installation Costs	\$3,890,530
TOTAL INSTALLED COST (for 10 units, 50,000KW total)	<u><u>\$5,158,630</u></u>

* Includes heat treatment and x-ray inspection for ASA 515-70
Carbon Steel Construction.

TABLE 4-7.

COST ESTIMATE

BRINE PUMPS

PLANT SIZE:	50,000 KW (net)
NUMBER OF PUMPS:	10
CAPACITY OF EACH PUMP:	2,196 GPM
SUCTION PRESSURE:	120 psia
DISCHARGE PRESSURE:	440 psia
TOTAL HEAD:	320 psi (709 ft.)
PUMP TYPE	Centrifugal, double suction
PUMP MATERIAL	Cast Iron w/Stainless Steel Trim
ELECTRIC MOTOR HORSEPOWER:	512 H.P. each @ 75% Pump Efficiency
TOTAL POWER REQUIREMENT:	402 KW @ 95% Motor Efficiency
MOTOR TYPE:	Explosion Proof

	<u>PUMP</u>	<u>DRIVER</u>
EQUIPMENT COST	\$25,000	\$14,280
INSTALLATION, MATERIAL & LABOR	26,750	4,320
INSTALLATION, INDIRECTS	<u>18,250</u>	<u>5,880</u>
SUB TOTAL:	\$70,000	\$24,480

TOTAL INSTALLED COST FOR 10 UNITS:	<u>\$944,800</u>
------------------------------------	------------------

TABLE 4-8

CAPITAL COST ESTIMATE

ISOBUTANE STORAGE

CAPACITY: 100,000 GALLONS
NUMBER: TWO TANKS, 50,000 GALLONS EACH
TYPE: HORIZONTAL PRESSURE VESSELS
DIMENSIONS: 12 FT. DIA. X 60 FT. LENGTH
TOTAL INSTALLED COST: \$165,000

BASIS: 40 PPM ISOBUTANE IN REJECTED BRINE
3,646,570 LBS/YR ANNUAL CONSUMPTION
40 DAY ADDITIONAL STORAGE CAPACITY
OFFSITE ISOBUTANE STORAGE FACILITIES ALREADY PRESENT
COST REPRESENTS 40 DAYS ADDITIONAL STORAGE REQUIRED
FOR INCREASED USAGE BY DIRECT CONTACT HEAT EXCHANGERS

4.2.5 Isobutane Storage

Table 4-8 summarizes the cost for the isobutane storage tanks. These costs are based on the assumption that an isobutane "binary" cycle plant would normally have off-site isobutane storage facilities with associated piping and civil works. As a result, the costs for this item reflects only the addition of 100,000 gallons of storage capacity. The use of direct contact heat exchangers without stripping columns and associated auxiliary recovery equipment is expected to use 800,000 to 1,000,000 gallons per year over and above that required of a binary cycle equipped with surface heat exchangers.

4.2.6 Operating Costs - Isobutane Loss

The cost of the isobutane dissolved in the rejected brine is considered an additional annual operating expense over a system using surface heat exchangers. The amounts of isobutane lost and the annual costs corresponding to anticipated rates of loss are tabulated in table 4-9. As discussed earlier it is difficult to accurately predict the exact isobutane losses for a very large once-through system such as the conceptual design. From equilibrium solubility data a loss of 40 ppm of isobutane in the brine effluent is the basis for this cost estimate. This rate assumes that the brine is in equilibrium with the isobutane at the exit temperature of 150°F. At a brine flow rate of 11,563,200 lb/hr, the isobutane loss of 40 ppm would be 462 lbs/hr or about 1.75 gallons per minute or 843,500 gallons per year. At this rate, the annual cost would be \$273,493 based on an isobutane cost of 7.5¢ per pound (delivered).

Thus, it is possible to recover most of the isobutane that may be dissolved in the brine by stripping in a reduced pressure tower. A system to recover 30 ppm of the isobutane was conceived, and a preliminary cost estimate indicated the total capital cost of over \$500,000. This would save one half of the \$273,490 annual cost for isobutane and \$10,750 annualized cost for one half of the isobutane storage for a net annual saving of \$82,000. This is approximately 7.5% of the projected annual cost of \$1,088,390. Considering the accuracy of this cost estimate, the accuracy of our estimates of dissolved isobutane and recovery efficiency, it is not practical to include credit for isobutane recovery at this time. Precise knowledge of the brine composition, especially the non-condensibles, will be required in order to make this decision for each particular application.

4.3 COST ANALYSIS

4.3.1 Production Costs

Table 4-10 summarizes the production costs attributed to the direct contact brine/isobutane heat exchangers for a 50,000 KW (net) plant. These figures are based on a 13% annual fixed charge rate which is average for a public utility with Federal Government assistance for financing. The effect of different rates of annual fixed charges on the production costs, are shown in figure 4-4. These rates are shown at different isobutane loss rates as well.

The relative importance of each cost center in the direct contact heat exchanger production cost component are shown on page 98 as a percentage of the total busbar component cost. In order to determine

TABLE 4-9

ANNUAL OPERATING COSTS

FOR ISOBUTANE LOSS

(Without isobutane recovery system)

<u>ISOBUTANE CONCENTRATION IN REJECTED BRINE (PPM)</u>	<u>ISOBUTANE CONSUMPTION LBS/YR.</u>	<u>ANNUAL COST, \$</u>	<u>COST MILS/KWH</u>
20	1,823,285	\$ 136,746	0.35
30	2,734,928	205,120	0.52
40	3,646,570	273,493	0.69
50	4,558,213	341,866	0.86
60	5,469,856	410,239	1.04
70	6,381,499	478,612	1.21

Basis: Plant Size: 50,000 KW (net)

Brine Flow: 11,563,200 lbs/hr

Load Factor: 90%

Isobutane Cost: 7.5¢/lb delivered

TABLE 4-10

COST SUMMARY FOR 50,000 KWDIRECT CONTACT HEAT EXCHANGE SYSTEM

I	CAPITAL COSTS	
	a. Heat Exchangers:	\$5,158,630
	b. Brine Pumps:	944,800
	c. Isobutane Storage:	<u>165,000</u>
	d. TOTAL INSTALLED COST:	<u>\$6,268,430</u>
	e. Unit installed Cost	\$ 125/KW
II	ANNUAL FIXED CHARGES (average @ 13%)	<u>\$ 814,896</u>
III	ANNUAL OPERATING COSTS (incremental for Isobutane losses)	<u>\$ 273,493</u>
IV	TOTAL ANNUAL COSTS	<u>\$1,088,390</u>
V	BUSBAR COMPONENT COST (90% Load Factor)	\$.00276/KWH

FIGURE 4-4
BUSBAR RATE COMPONENT VS.
AVERAGE FIXED CHARGES

the relative cost of brine pumping power (4020 KW), the brine pumping power was added to the net plant output of 50 MW, which brings the plant output to 54,020 KW, and reduces the total busbar component to \$.00255/KW-hr. The pumping power is then considered as an annual cost and based on the electric production cost of \$.00255/KWH and figured as a component of that production cost.

Heat Exchanger Towers	57.1%
Isobutane Losses	23.2%
Brine Pumps	10.4%
Pumping Power (brine pump)	7.5%
Isobutane Storage	1.8%
	<u>100.0%</u>

This is far from accurate since it ignores the incremental costs of other production components such as the wells, turbine, condenser, etc. Thus, the actual contribution of the brine pumping power requirements would be considerably greater than that given above. Total power production costs should be estimated to determine overall economic optimum conditions and the relative contributions of each component more exactly.

4.3.2 Cost Reductions

The cost estimate for the 50 MW direct contact system is considered very conservative. With continuing development of the system it is expected that these costs can be substantially reduced. Referring back to the relative costs of each component given under paragraph 4.3.1, it is apparent that the most costly part of the direct contact heat exchange system is the heat exchanger towers. This is expected, obviously, since this is the principal component. Optimistically, it is estimated that the cost of this item can be reduced by 50% by technical improvements. Since this is responsible for 57.1% of the total direct contact system, a cost reduction of 28% in the direct contact system could be achieved.

The cost of isobutane losses, responsible for 23.2% of this production cost could optimistically be reduced by 20%. This would reflect in a 5% savings for the total production cost. Thus, the total estimated saving from technical improvements could possibly be as high as 33% of the production cost estimated earlier.

4.4 COST SUMMARY

The capital costs of direct contact heat exchangers (brine to isobutane) plus brine pumps, isobutane storage and isobutane losses contribute \$.00276/KWH to total busbar production costs based on the concept and parameters set forth herein. This cannot, at this time, be realistically compared with a binary cycle using surface exchangers (closed cycle) since the optimum operating conditions for the two cycles would be different. The optimum economic conditions for the direct contact system were not determined.

SECTION 5

CONCLUSIONS AND RECOMMENDATIONS

5.1 GENERAL CONCLUSIONS

5.1.1 Commercial System Development

The analytical and experimental program just completed has probed several aspects of direct contact heat exchange between geothermal brine and isobutane. The overall conclusion is that there are no major technical obstacles to the development of a commercial system. Furthermore, the prospects for the early development of a commercial system appear favorable, particularly for highly saline brines. However, a cautious approach should be taken in the development of the system until the economic comparisons with conventional or other developmental heat exchange systems can be made accurately. Accuracy in comparisons will require that additional trials with all types of heat exchangers be conducted with the more perverse types of geothermal brines in order to document their performance and operating costs under these conditions.

5.1.2 Capital Costs .

The total installed cost for a 50 MW direct contact heat exchange system with brine pumps was estimated to be \$6,268,000 or \$125/KW capacity. This estimate assumes however, that the performance of the direct contact unit is the same as the small experimental unit tested and no credit was taken for anticipated improvements.

5.1.3 Annual Costs

The annual costs due to the direct contact system was estimated at \$1,088,390 for a 50 MWe net power plant. This includes a 13% fixed charge rate and \$273,493 for isobutane losses.

5.1.4 Production Costs

The busbar unit production cost component for the direct contact system was estimated at \$.00276/KWH. This does not consider all the costs contributed due to the pumping power required by the brine pumps(s). For this, it would be necessary to determine the total production costs and incremental costs with and without the power needed for the brine pumps. However, it should also be realized that an overall system cost analysis is needed to determine the economic optimum point. This was not accomplished in this phase of the research work.

5.1.5 Design Data

The effects of varying the tower diameter and height on the backmixing and resulting tower performance is unknown. Additional heat transfer data of a more localized nature will be required in order to completely characterize the heat transfer phenomenon taking place. Gross overall performance data such as the volumetric heat transfer coefficients obtained in the experiments were not sufficient for more than characterizing the existing test unit. Additional instrumentation is required to make an accurate thermal profile of both isobutane and brine in the spray tower preheater. With this knowledge, the effects of backmixing can be established as well as the effects of tower length and nozzle types. This will facilitate the design of larger units to some extent

but will not define the effect of varying diameter on backmixing. The heat transfer data referred to above must be obtained in an experimental unit resembling the complete system and be done in a tower of significant diameter. Tests for backmixing and other variables should be determined in tests with larger and larger diameter columns as the development program proceeds.

5.1.6 Technical Improvements

Several means for improving the performance of the preheater have been conceived but not yet implemented in the experimental unit. A conservative estimate sets the anticipated performance gains as follows:

- a. The number of heat transfer units (NTU) may be increased from 2.7 to 3.25 (+20%) by increasing the hold-up of isobutane in the preheater. This should result in a 20% reduction in height and a corresponding capital cost savings.
- b. The throughput of fluids (superficial velocities) through the unit can be increased 20%. This will allow the diameter to be reduced 10% and provide a capital cost saving.

5.1.7 Cost Reductions

The capital and operating costs of the 50 MW plant presented in this report can be reduced by economic optimization of the conceptual design. Potential savings may be realized in the following areas:

- a. The isobutane boiling pressure (and temperature), if reduced, will allow the use of thinner pressure vessel walls. Lower operating pressures also reduce the brine booster pump capital and operating costs. The reduction in operating pressures and attending cost savings must be weighed against reduced plant efficiency.
- b. The diameter of the conceptual preheater may be reduced. The exterior shell diameter was set by the brine flow requirements at the conical inlet and outlet ends of each preheater stage. There is a possibility that this configuration can be modified to reduce or eliminate the end cones and, as a result, the entire shell diameter can be reduced.
- c. In conjunction with "a" and "b", above, the size and number of vessels for a 50 MW plant may be optimized to reduce the total capital cost for vessels.

5.2 SPECIFIC CONCLUSIONS FROM EXPERIMENTAL PROGRAM

5.2.1 Operation

Continuous and stable operation of the experimental isobutane preheater and boiler were easily initiated and maintained. No automatic controls were utilized during the test.

5.2.2 Preheater Heat Transfer

The six inch diameter, six foot high preheater tested during this pro-

gram transferred heat from a 21% brine solution to isobutane at a rate of 3500 BTU/hr-°F-ft³. The heat exchanger effectiveness was 80% which for the design hourly heat capacity ratio corresponds to 2.7 heat transfer units (NTU). The NTU obtained in the preheater is governed by the axial mixing behavior of the heat exchanger and not by local drop heat transfer coefficients. The preheater was not operated at high enough flows and hold-up to achieve a "close-packed" mode of flooding and therefore did not achieve what we believe is full potential with regard to both thermal performance as well as hydraulic capacity. The flows were limited by pump capacity.

5.2.3 Boiler Heat Transfer

The six inch diameter experimental boiler transferred heat at rates of up to 17,000 BTU/hr-°F,ft³ based on an effective volume of .4 cu.ft. The boiler also exhibited a capacity to preheat the isobutane quite efficiently. The boiler can be operated to provide at least one heat transfer unit of liquid-liquid preheating. In other words the boiler acts as a perfect mixed heat exchanger for liquid-liquid heat exchange, and the possibility of a combined preheater-boiler is promising.

5.2.4 Droplet Size

The isobutane droplet size and distribution appeared uniform as observed through site glasses at two locations. The isobutane droplets did not exhibit any premature coalescence either with each other or with the steel shell. Droplet size averaged .15 inches in diameter. The plastic coalescing screen at the top of the unit apparently functioned properly as no brine was observed to enter the zone of coalesced isobutane at the top of the unit. The interface between the brine (with dispersed isobutane) and the coalesced isobutane was operated at levels below the coalescing screen on several occasions. This has raised the questions of whether the coalescing screen and conical ends are necessary.

5.2.5 Working Fluid

Isobutane as a working fluid is easy to work with and observations disclosed no undesirable characteristics such as sludge formation, foaming or chemical reactions in the heat exchangers. The isobutane quickly dissipates to the atmosphere when small leaks occur and no explosive concentrations were detected more than a few inches away from small leaks. The entire charge of isobutane was blown out of a vent stack when a rupture disc failed and no explosive concentrations were detected at ground level even though the isobutane is heavier than air.

5.2.6 Water Vapor

Measured amounts of water vapor in the isobutane vapor from the boiler ranged from 0.2 to 1.9%. This amount of water vapor is lower than predicted by assuming the water vapor temperature to be equal to the isobutane temperature at the boiler outlet. The amounts of water vapor measured were quite erratic with time and the accuracy of the sampling method is in question. A more comprehensive investigation of the amount of water vapor in the isobutane vapor leaving the boiler is definitely required. Also, formulas for cycle state points and overall performance should be modified to include the appropriate factors for the

effect of the water vapor produced.

5.2.7 Dissolved Isobutane

The method of measuring the dissolved isobutane in the brine effluent from the experimental heat exchangers was unsatisfactory. Attempts to use spectographic carbon analyzers require the sample to be at atmospheric pressure and room temperature. Since the samples from the heat exchangers were at a pressure of 300 psia and the temperature from 130 to 180°F, it was impossible to accurately measure dissolved isobutane by this method. A physical measurement was made of the dissolved isobutane by flowing a small sample stream through a stripping chamber. This method was successful with fresh water, but not sufficiently sensitive to the low concentrations (20-40ppm) encountered with higher salinity brines.

5.3 RECOMMENDED DEVELOPMENT PROGRAM

5.3.1 Introduction

Direct Contact Heat Exchangers in conjunction with an isobutane binary cycle appear to have outstanding potential for use with liquid dominated geothermal resources. Most of the recommendations and their underlying objectives are a direct result of the recently completed program. Typical of a development program of this nature, the original questions and objectives have been replaced by new ones generated by the test operations. The original question - does it work? has been answered affirmatively, but numerous possibilities for improving the performance and economics of the system have presented themselves to the investigators. It is, of course, impossible to test all of these ideas, so the main task of formulating the recommended development program was the selection of the most promising, or demanding, of these new ideas and incorporate them into a workable development plan. A development program is recommended below that will utilize the existing test apparatus in conjunction with additional analytical studies for the next phase of development.

5.3.2 Operation With Geothermal Brine

The recommended program is aimed at satisfying three general objectives. The first objective should be to demonstrate the ability of the system to operate reliably with geothermal brine in a field application. This objective can be satisfied by operating the test unit at the new East Mesa Geothermal Facility on a continuous basis for a total on stream time of at least four months. The existing test unit should be modified to operate round-the-clock, and a duplicate direct contact pre-heater should be installed in order to duplicate the entire temperature should prove the system's resistance to scaling and corrosion which is prevalent with equipment exposed to geothermal brines. These field tests should also be designed to document the exact amount of isobutane loss from the system and the amount of water vapor carryover.

5.3.3 Improve Performance and Reduce Costs

The second objective should be the further development of the process to improve the performance and reduce the costs of the system. To satisfy this objective several experiments should be initiated before the unit is moved from its present location in North Carolina to the East Mesa Facility. The effect of increasing the throughput and varying the length of the direct contact preheater should be explored. Subsequently the boiler and preheater should be mated to form a combination preheater-boiler, since this is incorporated in the conceptual design and appears to be desirable for large scale units. These experiments can be done with existing hardware. Additional experiments should be initiated at East Mesa concurrent with the long term testing to further improve performance and cost characteristics.

5.3.4 Engineering Design and Cost Analysis

The third objective should be the creation of "second generation" conceptual designs and cost estimates for commercially sized plants. The conceptual designs should explore such problems as scale-up from pilot to commercial size plants, optimum module sizes, cycle thermodynamics, heat rejection, and an economic analysis. The results of the conceptual studies should then be utilized to prepare a detailed pilot plant design and cost estimate.

The pilot plant design should be complete with drawings and specifications suitable for obtaining accurate bids for construction. A detailed cost estimate should be prepared based on actual quotations.

5.3.5 Specific Development Tasks

A more specific set of recommendations in accordance with the above three objectives is presented on the following pages. The recommended development tasks are listed according to their general objective areas and not in any particular order of execution. It is certain that these will be many areas of overlapping functions.

5.4 RECOMMENDED FIELD DEMONSTRATION

5.4.1 Automatic Operation

The existing test unit should be modified so that the unit can be operated continuously 24 hours a day with only "supervisory" attendance. This modification would require installing standard pneumatic controls so that the unit can be tested over a reasonably long period of time under actual field conditions. Many of the effects of geothermal fluids on process equipment can be judged only by lengthy exposure. The heat exchangers will ultimately be expected to give 20 years of service in a commercial application and it is not unreasonable to test such units at this stage of development for several months. Pneumatic pressure and level controls will allow the unit to be tended by shift personnel at the East Mesa Facility and thereby avoid the cost of two additional shift personnel. A schematic of the recommended controls is shown in Figure 5-1.

5.4.2 Additional Preheater

The existing test unit incorporates one preheater capable of increasing the temperature of the isobutane by 35°F to 80°F. This isobutane temperature rise is not enough to simulate the entire binary cycle. The experiments were conducted by simulating a "truncated" cycle in such a manner that one half to one third of the temperature range of a complete binary cycle was simulated at any given time. For the field tests, however, it is important that the entire temperature range in the preheater be simulated. The reason being that corrosion, precipitation and scale deposition are a function of brine temperature. Therefore, we recommend installing a duplicate preheater in series with the existing preheater and thus enable the unit to simulate a complete binary cycle and operate over the full brine temperature range. At design conditions the brine will be cooled, from 325°F to approximately 138°F. Figure 5-2 shows the modified flow sheet for field testing. The duplicate preheater is shown in phantom lines.

5.4.3 Measuring Water Vapor Carry-over

The method of measuring water vapor carryover should be improved so that accurate long term data can be obtained during the field testing. The existing method for collecting condensed water vapor at the condenser outlet should be modified. Either install a large settling chamber between the existing condenser and hotwell or enlarge the sump capacity of the hotwell. In addition, differential water vapor pressure instrumentation should be installed directly at the boiler outlet to monitor the water vapor content continuously.

5.4.4 Isobutane Losses

It is extremely important to obtain long-term data on isobutane losses during the recommended field tests. To this end we recommend installing a stripping column, or degassifier, that is sized for 5 to 6 gpm discharge of brine from the test unit. The unit should operate between 1/2 to 1 atmosphere of pressure. The stripping chamber would be quite small for this amount of brine flow and should be inexpensive. The isobutane collected from the chamber would amount to only one or two ounces an hour so that small laboratory equipment could be used to evacuate the isobutane from the chamber and condense it. The brine effluent from the degassifier will be at atmospheric pressure and can be analysed with a spectographic carbon analyzer in order to complete the isobutane loss measurement.

5.5 HEAT EXCHANGER PERFORMANCE IMPROVEMENTS

5.5.1 Operate The Existing Preheater With Increased Hold-up and Achieve A "close packed" Mode of Hold-up.

The hold-up of isobutane, expressed as a percentage of the volume of the tower, has been measured at several operating conditions and

FLOW SHEET

EAST MESA BRINE TAKEN AS 20,000 P.P.M. TDS FOR PURPOSE OF HEAT BALANCE

STREAM	1	2	3	4	5	6	7	8	9	10
PARAMETER	I.B. HIGH PRESSURE OF 1ST CONDENSATE PREHEATER	I.B. OUT OF 1ST PREHEATER	I.B. OUT OF 2ND PREHEATER	I.B. VAPOR OUT OF BOILER	GEOTHERMAL BRINE FROM WELL	BRINE TO BOILER	BRINE TO 2ND PRE-HEATER	BRINE TO 1ST PRE-HEATER	BRINE TO STRIPPER	BRINE TO WASTE
FLOW, LB./HR.	2650	2650	2650	2650	2292	2292	2292	2292	2292	2292
TEMPERATURE, °F.	110	165	220	220	325	325	228	184	138	138
PRESSURE, PSIA	320	325	320	315	90	320	325	320	310	40
ENTHALPY, BTU/LB.	5	42.5	80	165	289	289	192	149	104	104
DENSITY	32.9	30.2	26.4	3.7	57.5	57.5	602	61.3	62.2	62.2

DEVELOPMENT OF DIRECT CONTACT HEAT EXCHANGERS FOR GEOTHERMAL BRINES
PHASE II
EAST MESA OPERATIONS

FIGURE 5-2

varies from zero (at no flow) to 45%. Most of the heat transfer data was obtained at 25 to 35% hold-up. It was not until the last week of testing that a method of manipulating the numerous flow control valves to attain significantly higher hold-up (45%) was discovered. This mode of hold-up is what is referred to in the liquid extraction literature as "close-packed" hold-up, whereby the isobutane droplets become packed together much like a jar of marbles, each droplet in contact with others. The limited number of data points at this mode of operation indicates an improvement in volumetric heat transfer coefficients, but testing with this condition at all temperatures will require increasing the throughput of both isobutane and brine per unit of tower cross sectional area. (Superficial velocity).

The positive displacement isobutane and brine pumps were operating at their limits. However, it may be possible to increase their output by simply changing the sheaves on the pumps, thereby increasing their speed and output. Another method to increase superficial velocities is to reduce the cross sectional area of the column by installing a dummy core in the column along its vertical axis. This core could also be used to contain thermocouples and thereby obtain an accurate temperature profile. (see recommendation 5.5.6.)

5.5.2 Determine The Effect Of Varying The Preheater Column Height.

It has been shown analytically that increases in column height will increase the effectiveness of the preheater. The height of the existing column is six feet. This includes five feet of straight section and and one foot of the tapered sections at the ends. Increasing the height (and volume) will not increase the volumetric heat transfer coefficient, in fact, it will undoubtedly decrease. However, the heat exchanger effectiveness may be increased by additional height. By virtue of the increased interfacial surface created in the taller tower closer approaches can be attained.

The increase in heat exchanger effectiveness, however, is not linear with length, but increases asymptotically toward the theoretical limit. In other words, the increase in effectiveness is diminishing with increased length so that an optimum with respect to performance and cost must be determined. The experimental unit tested has an effective height of six feet. We recommend testing the same unit with mid-sections of different length so that the performance can be determined for four, eight and ten foot heights. Greater heights should be tested if the data indicates significant increases in performance with taller columns occurs. This experiment should not begin until the flooded or "close-packed" mode of hold-up is attained as recommended above.

5.5.3 Combine Preheater and Boiler

An experimental change to the existing hardware is recommended whereby the preheater and boiler are combined into one integral unit. The original test unit was designed to have a separate boiler and preheater in order to determine the characteristics of these units separately before a combination was attempted. The units were designed so that they could ultimately be bolted together with mating flanges to form a combination boiler-preheater. The benefits to a commercial plant derived

from this combination include the elimination of one pump, one set of level controls, and interconnecting piping. It is envisioned that isobutane droplets in the preheater would rise directly from the preheater into the boiler. This would eliminate the costly top head assembly of the preheater where the isobutane is presently coalesced and separated from the brine, and the brine is introduced after being pumped from the boiler. It would also eliminate the bottom section of the existing boiler. See figure 5-3.

5.5.4 Test Different Nozzles and Distributors

Increased performance can be obtained by improvements in the isobutane and brine nozzles and distributors. The heat transfer performance of the preheater is expected to be increased if smaller isobutane droplets could be created. The present distribution plate at the bottom of the preheater contains 390 holes, 1.5 mm diameter and the drop size observed in the sight glasses appeared to be about 3 mm in diameter. Since the interfacial surface area is directly proportional to the reciprocal of droplet diameter at constant isobutane hold-up, a 20% reduction in droplet diameter should theoretically create 20% increase in heat transfer. The existing distributor plate was not fouled or plugged despite the small hole diameter. However, we feel that it is unwise to reduce the diameter further in an attempt to produce smaller droplets. We recommend testing at least one atomizing nozzle. One type of nozzle is actuated by the momentum of the isobutane itself and doesn't require external power to activate it. At least one type of porous ceramic nozzle should be tested as well.

The preheater brine distributor at the top of the unit has been modified once to create a smoother, less turbulent entry of the brine into the bed of isobutane droplets. There was a small improvement in heat transfer, but there was still some turbulence being produced. Therefore, we recommend a new version of the distributor that will provide even smoother entry of the brine. It has been suggested that the brine/isobutane interface at the top of the preheater be lowered to the mid point of the preheater and thus create an "inverted" dispersion in the upper half of the preheater, i.e., brine droplets dispersed in isobutane. While this promises to increase the thermal performance we do not recommend doing this unless a non-fouling brine nozzle can be found that will produce the necessary small droplets at low pressure drops.

It is recommended that the spray type of brine inlet in the boiler be eliminated and introduce the brine below the free surface. The boiler brine inlet nozzle currently sprays the brine into the top of the boiler vapor space. Although it is a swirl type nozzle that contains no small orifices, it may be prone to plugging when used with real geothermal brines. The original objective of this nozzle was to accomplish superheating of the isobutane. It is now evident that more than 20°F superheat is unnecessary for the isobutane cycle. In fact this is undesirable unless a regenerator, or economizer is used in the power cycle. Apparently the isobutane picks up sufficient superheat from the hot water vapor present above the brine free surface level. We recommend eliminating the spray type of brine inlet in the boiler and introducing the brine below the free surface level of the brine.

DEVELOPMENT OF DIRECT CONTACT HEAT EXCHANGERS FOR GEOTHERMAL BRINES

FIGURE 5-3
FLOW SHEET FOR COMBINATION
BOILER-PREHEATER

5.5.5 Working Fluid

Mixtures of working fluids should be investigated and tested in the experimental unit. The existing equipment is well suited for testing mixtures of binary fluids and no major modifications should be necessary to do so.

5.5.6 Temperature Measurement

Temperature instrumentation sufficient to make a detailed temperature profile of both brine and isobutane should be installed in the preheater. The effects of axial backmixing of the brine has been extremely difficult to ascertain. We recommend installing additional thermocouples at a minimum of one foot vertical spacing on the preheater. At each location two thermocouples will be required; one with an exposed steel sheath which will be wetted by brine and the other thermocouple enclosed in a device with plastic packing so that it will capture and be wetted by isobutane droplets. This information will enable the actual local temperature differences between brine and isobutane to be determined and the local heat transfer coefficients and backmixing effects determined.

5.6 RECOMMENDED ENGINEERING DESIGN AND COST ANALYSIS

5.6.1 Refine Conceptual Designs

Future test data, especially from field testing must be applied to the refinement of conceptual designs of commercial systems. In addition, the heat exchanger concepts must not be considered alone but incorporated into complete reservoir/power plant/reinjection systems. Special attention should be given to the restraints and interactions imposed by the heat rejection system whether it is a wet cooling tower, dry tower, or direct contact condenser/dry tower hybrid.

5.6.2 Design Data with Different Brines

The conceptual designs and optimizations should be enlarged to accommodate several specific types of brines. To create a more accurate economic picture of the direct contact system, at least one specific brine temperature and salinity must be selected. The brine conditions at the inlet to the heat exchangers will require knowledge of the treatment the brine receives before it reaches the heat exchanger, such as whether it is pumped with or without flashing, the amount of concentration, composition, and most important, the value of the brine and its heat content.

5.6.3 Optimize Designs

A design optimization should be made considering the water vapor produced, the boiler pressure, plant efficiency, vessel wall thickness, number of vessels and their diameter, and brine pump capital and operating cost. Lower boiler pressures will lead to thinner vessel wall thickness and/or larger and fewer vessels, as well as lower brine pump capital and operating costs. New heat exchanger test data must be incorporated into a cost model which will lead to the lowest overall plant production costs.

5.6.4 Heat Rejection Studies

While not a part of the development effort reported herein, the problem of heat rejection should be dealt with at once. The size of the condensers, cooling towers and the cooling water requirements appear to be formidable. Direct contact condensation of isobutane on water followed by separation and cooling of the water in a conventional cooling tower should be included in the heat rejection study. In light of the large water requirements for cooling towers, dry heat exchangers may be necessary. The overwhelming cost of dry towers may be reduced if the plastic heat exchange surface being developed for the ocean thermal energy program can be incorporated.

5.6.5 Design a Pilot Plant

The results of the conceptual design studies should be utilized to prepare a detailed pilot plant design and cost estimate. This design should be at least 20 times larger (100 gpm brine) than the present test unit and include the complete binary power cycle with direct contact heat exchangers. It should also include a turbine generator with a net electric output of 100 KW min. The system with 100 gpm of 350°F brine will produce sufficient vapor to make about 300 KWe.

The pilot plant design should be complete with drawings and specifications suitable for obtaining accurate bids for construction. The cost estimate should be based on quotations for equipment required.

APPENDIX A
TEST RESULTS AND DATA
ANALYSIS EXAMPLE

APPENDIX A

TABLE A - 1
PREHEATER TEST RESULTS

TEST	M _f	M _b	T _{f3}	T _{f4} -T _{f3}	ΔT _{LM}	Q _f	U _v	NTU
C-2-1	793	2,105	69.0	59.0	17.5	28,099	1,339	3.37
C-2-7	768	2,211	69.0	63.0	26.0	29,036	928	2.42
C-2-8	768	2,207	68.5	67.5	27.6	31,628	954	2.45
C-2-9	751	2,092	97.5	46.5	21.1	21,663	854	2.20
C-2-10	752	2,092	97.0	48.0	20.9	22,395	892	2.30
C-2-11	751	1,925	98.5	47.0	18.4	21,896	993	2.55
C-2-12	751	1,812	99.5	43.0	18.6	20,032	898	2.31
C-2-13	729	2,369	94.0	54.5	22.1	24,629	926	2.47
C-2-14	942	2,090	111.5	37.5	15.7	22,249	1,182	2.39
C-2-15	895	2,088	106.5	41.5	15.6	23,038	1,234	2.66
C-3-3	521	2,263	111.0	48.5	24.3	15,927	545	2.00
C-3-5	496	2,206	111.0	52.5	23.9	16,419	572	2.20
C-3-7	494	2,085	118.0	60.0	29.4	19,263	545	2.04
C-4-1	588	2,215	73.0	58.0	36.3	20,465	469	1.60
C-4-6	743	2,211	70.0	83.0	39.2	37,595	799	2.12
C-4-8	692	2,209	68.0	89.0	39.3	36,742	779	2.21
C-4-12	1,534	2,191	71.0	76.0	32.4	69,954	1,799	2.35
C-4-16	1,583	2,187	74.0	75.0	31.7	72,419	1,905	2.37
C-4-14	1,585	2,245	71.0	77.0	33.8	74,458	1,833	2.28
C-4-23	1,513	2,242	89.0	75.0	25.3	70,366	2,317	2.96
C-5-6	1,566	2,172	64.0	86.0	37.5	80,826	1,797	2.30
C-5-7	1,541	2,230	66.0	86.0	37.5	79,501	1,768	2.30
C-5-10	1,534	2,175	73.0	86.0	31.0	80,478	2,160	2.77
C-5-15	2,010	2,219	94.0	65.0	24.0	80,991	2,814	2.71
C-6-5	1,955	2,050	127.5	24.5	10.2	27,533	2,454	2.40
C-6-12	1,927	2,046	142.0	18.0	7.2	21,294	2,590	2.50
C-7-4	999	2,132	70.0	74.0	30.7	44,338	1,202	2.41
C-7-9	1,529	2,213	100.0	56.0	23.3	53,083	1,899	2.40
C-7-12	2,007	2,094	112.0	44.0	17.9	55,647	2,585	2.45
C-7-17	2,171	2,037	145.0	18.0	9.6	25,405	2,195	1.88
C-8-5	1,996	2,201	120.0	44.0	16.5	56,212	2,839	2.67
C-8-6	1,961	2,188	127.0	48.0	17.8	61,200	2,860	2.69
C-8-9	1,939	2,186	138.0	44.0	16.8	56,318	2,790	2.62
C-8-11	1,925	1,870	146.0	34.0	15.2	43,853	2,398	2.23
C-8-14	1,415	1,329	151.0	31.0	17.0	29,394	1,442	1.83
C-8-17	983	1,276	150.0	36.0	22.5	23,720	879	1.60
C-9-4	1,976	2,200	118.0	50.0	17.5	63,221	3,006	2.86
C-9-7	1,957	2,084	142.0	35.0	14.7	45,200	2,561	2.38
C-9-10	1,917	2,148	149.0	37.0	18.4	47,519	2,154	2.01
C-9-13	1,809	2,034	145.0	41.0	21.5	49,706	1,930	1.91
C-9-15	1,775	2,032	149.0	40.0	19.7	47,570	2,010	2.03
C-9-18	1,793	2,139	153.0	43.0	20.3	52,433	2,152	2.12
C-10-9	2,410	2,297	163.0	7.0	10.3	11,303	914	0.68
C-11-4	2,499	2,144	138.0	12.0	6.0	19,195	2,672	2.00
C-11-5	2,561	2,367	128.0	29.0	13.4	47,540	2,945	2.16
C-11-7	2,566	2,192	131.0	31.0	12.4	50,920	3,411	2.50
E-1-3	663	2,213	75.0	59.0	24.0	23,487	815	2.46
E-1-6	1,005	2,371	93.0	55.0	26.5	34,265	1,074	2.08
E-1-9	1,460	2,474	132.0	28.0	14.2	26,162	1,534	1.97
E-1-12	1,954	2,498	136.0	27.0	15.7	34,301	1,815	1.72
E-1-14	2,469	2,720	140.0	23.0	12.2	36,908	2,519	1.89
E-1-16	2,565	2,717	140.0	25.0	12.9	25,014	2,533	1.94

TABLE A - 1

PREHEATER RESULTS (Con't)

TEST	M _f	M _b	T _{f3}	T _{f4} -T _{f3}	ΔT _{LM}	Q _f	U _v	NTU
E-2-4	2,069	2,443	134.0	20.0	11.9	26,485	1,843	1.68
E-2-7	1,996	2,433	134.0	33.0	21.0	42,816	1,697	1.57
E-2-10	2,020	2,543	135.0	45.0	22.0	60,007	2,269	2.04
E-3-3	2,058	2,489	140.0	35.0	17.9	47,547	2,210	1.96
E-3-6	1,958	2,287	141.0	43.0	21.4	56,419	2,197	2.01
E-3-10	2,304	2,658	150.0	30.0	14.5	46,308	2,662	2.07
E-4-3	1,982	2,377	127.0	32.0	18.0	40,594	1,881	1.78
E-4-6	2,002	2,658	145.0	37.0	17.4	49,642	2,370	2.12
E-4-9	1,999	2,485	147.0	34.0	17.2	45,544	2,201	1.97
E-4-12	2,168	2,261	147.0	33.0	16.2	47,932	2,460	2.04
E-5-5	1,988	2,550	152.0	33.0	17.5	43,947	2,093	1.89
E-6-4	1,124	2,447	78.0	74.0	27.8	50,761	1,522	2.66
E-6-8	1,489	1,976	110.0	34.0	18.4	31,394	1,423	1.85
E-6-11	1,722	2,262	120.0	34.0	17.9	36,897	1,715	1.90
E-6-13	1,934	2,545	127.0	33.0	16.4	40,847	2,069	2.01
E-7-1	2,042	2,839	121.0	42.0	18.8	54,901	2,427	2.23
E-7-2	2,052	2,830	128.0	44.0	19.7	58,693	2,483	2.23
B-1-1	1,019	1,777	76.0	74.0	30.8	46,009	1,243	2.40
B-1-2	1,021	1,779	75.0	72.0	28.7	44,831	1,300	2.51
B-1-3	1,524	1,611	82.2	62.3	26.1	57,905	1,845	2.38
B-1-4	1,544	1,663	85.4	57.6	25.7	54,242	1,761	2.24
B-1-5	1,578	1,721	100.0	43.0	18.3	42,067	1,912	2.35
B-2-1	1,967	2,227	91.0	39.0	16.9	47,568	2,344	2.30
B-2-2	1,920	2,398	109.0	26.0	9.8	30,948	2,626	2.65
B-2-3	1,980	2,395	115.0	19.5	9.8	23,934	2,022	1.98
B-2-4	1,822	2,220	103.0	45.0	14.9	50,834	2,839	3.02
B-2-5	1,787	2,000	110.0	34.0	16.8	37,673	1,866	2.02
B-3-1	1,764	2,391	142.5	16.5	6.7	18,919	2,343	2.46
B-3-2	1,761	2,389	144.5	13.5	5.4	15,455	2,368	2.50
B-3-3	1,758	2,389	145.5	13.5	8.7	15,430	1,480	1.55
B-3-4	1,686	2,389	147.4	13.6	6.5	14,905	1,903	2.09
B-3-5	1,755	2,500	148.8	14.7	7.0	16,766	1,984	2.10
B-4-1	978	2,001	97.0	52.0	27.3	31,539	962	1.90
B-4-2	1,428	1,806	100.0	54.0	26.9	47,798	1,481	2.01
B-4-3	1,728	2,167	100.0	53.0	24.4	56,789	1,942	2.17
B-4-4	1,756	2,165	96.0	57.5	24.4	62,597	2,138	2.36
B-4-5	1,763	2,715	110.0	57.0	23.9	63,316	2,209	2.38
B-4-6	1,781	2,662	112.0	53.5	22.7	60,022	2,205	2.36
B-4-7	1,756	2,660	113.0	60.5	20.4	67,996	2,769	2.97
B-4-8	1,800	2,766	119.0	59.0	23.3	69,030	2,469	2.53
B-4-9	1,778	2,208	117.0	62.0	29.4	71,657	2,034	2.11
B-5-2	2,305	1,559	96.0	28.0	15.1	39,377	2,167	1.85
B-5-3	2,238	1,671	107.0	25.5	12.5	35,389	2,360	2.04
B-5-5	2,354	1,889	113.0	30.0	15.5	43,793	2,355	1.94
B-5-8	2,186	2,054	124.0	33.5	14.9	46,860	2,619	2.25
B-5-10	2,105	2,108	127.5	35.5	16.0	47,834	2,486	2.22
B-5-12	2,116	2,159	122.0	47.0	19.0	63,657	2,781	2.47
B-6-4	2,500	2,450	101.0	37.0	18.8	57,343	2,536	1.96
B-6-8	2,473	2,493	115.0	45.0	23.1	70,120	2,525	1.94
B-6-13	2,491	2,215	105.0	58.0	28.0	91,018	2,701	2.07
B-6-17	2,494	2,444	117.0	53.0	29.5	84,606	2,390	1.80
B-6-21	2,530	2,205	99.0	72.0	35.5	114,747	2,695	2.03

TABLE A - 1
PREHEATER TEST RESULTS (Con't)

TEST	M _f	M _b	T _{f3}	T _{f4} -T _{f3}	ΔT _{LM}	Q _f	U _v	NTU
B-6-24	2,680	2,253	100.0	77.0	37.7	132,069	2,921	2.04
B-7-6	2,416	2,624	106.5	25.0	9.6	36,847	3,206	2.61
B-7-12	2,354	2,616	125.0	26.0	14.5	38,556	2,217	1.79
B-7-14	2,420	2,452	118.0	40.0	21.5	60,988	2,364	1.86
B-7-17	2,388	2,438	107.0	53.0	24.9	79,724	2,663	2.13
D-1-4	2,498	2,886	114.0	14.0	6.5	21,683	2,785	2.16
D-1-8	2,537	2,933	119.0	23.0	10.5	36,765	2,920	2.19
D-1-10	2,496	2,997	112.0	36.0	15.1	56,607	3,127	2.39
D-1-13	2,513	3,054	94.0	53.0	20.6	82,589	3,341	2.57
D-1-18	2,498	2,943	115.0	51.0	21.2	81,538	3,208	2.41
D-2-5	2,399	2,823	114.0	18.0	7.5	26,775	2,979	2.41
D-2-7	2,478	3,050	114.0	36.0	14.3	56,209	3,263	2.51
D-2-9	2,446	3,157	97.0	58.0	23.7	87,949	3,090	2.45
D-2-12	2,555	3,143	100.0	72.0	29.3	115,883	3,295	2.46
D-2-15	2,809	3,192	84.0	96.0	44.4	169,899	3,188	2.16
D-3-4	2,930	2,236	97.5	21.5	9.8	38,118	3,245	2.19
D-3-7	2,914	2,585	106.0	27.0	12.5	48,393	3,213	2.16
D-4-2	2,547	3,004	109.5	45.5	5.9	85,423	4,053	3.19
D-4-4	2,546	2,910	96.5	39.5	11.0	70,110	2,713	2.10
D-4-7	2,503	3,048	112.0	38.0	15.4	59,925	3,250	2.47
D-4-10	2,589	3,042	98.0	23.0	20.2	33,718	3,675	2.75
D-4-13	2,634	2,888	90.5	19.0	22.6	29,039	3,849	2.81
D-4-17	2,910	2,490	97.0	63.5	19.7	104,529	2,958	3.22
D-4-20	2,980	2,453	95.0	55.5	19.4	89,080	3,661	2.86
D-5-12	2,956	2,817	123.0	26.0	11.7	48,414	3,438	2.22
D-5-15	2,974	2,868	111.0	41.0	17.0	76,819	3,770	2.41

TABLE A - 2

BOILER TEST CONDITIONS

TEST	T _{b1}	T _{b5}	T _{f5}	T _{f1}	P _t	M _b	M _f	M _w
C-2-1	182	131.5	128	134.5	118	2,105	793	7.19
C-2-7	194	140.5	132	163.5	122	2,210	768	11.55
C-2-8	202	145.0	136	196.0	122	2,207	768	11.55
C-2-9	212	152.5	144	178.0	154	2,091	751	11.51
C-2-10	215	152.5	145	178.0	145	2,091	752	8.25
C-2-11	218	151.0	145.5	180.0	145	1,924	751	8.47
C-2-12	224	149.0	142.5	171.0	145	1,811	751	8.89
C-2-13	209	155.5	148.5	186.0	142	2,368	728	12.49
C-2-14	218	155.0	149.0	159.5	161	2,089	941	8.99
C-2-15	221	153.0	148.0	182.0	155	2,088	895	17.36
C-3-3	205	170.0	159.5	178.0	198	2,262	521	5.91
C-3-5	211	172.5	163.5	179.5	200	2,205	496	4.85
C-3-7	234	191.0	178.0	194.5	248	2,084	493	5.43
C-4-6	223	170.5	153.0	194.0	180	2,198	742	9.33
C-4-8	225	171.5	155.0	192.0	183	2,196	692	6.67
C-4-12	273	165.0	147.0	168.0	183	2,170	1,534	8.19
C-4-14	268	166.5	148.0	171.0	182	2,228	1,585	8.41
C-4-16	266	167.5	149.0	171.0	187	2,174	1,582	8.32
C-4-23	276	176.0	164.0	182.0	195	2,224	1,513	8.62
C-5-6	270	170.0	150.0	185.0	189	2,172	1,566	8.80
C-5-7	264	172.0	152.0	190.0	199	2,230	1,540	8.94
C-5-10	263	172.0	159.0	180.0	203	2,175	1,534	14.17
C-5-15	285	174.0	159.0	182.0	213	2,218	2,009	6.08
C-6-5	302	160.0	152.0	167.0	183	2,049	1,955	5.73
C-6-12	310	170.0	162.0	182.0	210	2,046	1,927	4.84
C-7-4	243	157.0	144.0	196.0	143	2,132	998	8.92
C-7-9	298	170.0	156.0	227.0	176	2,213	1,528	13.02
C-7-12	315	169.0	156.0	202.0	181	2,094	2,007	12.73
C-7-17	335	174.5	163.0	195.5	195	2,037	2,171	5.14
C-8-5	323	176.0	164.0	204.0	202	2,201	1,996	12.41
C-8-6	335	188.5	175.0	215.5	225	2,188	1,961	10.87
C-8-9	336	196.0	182.0	218.0	245	2,186	1,939	8.26
C-8-11	335	194.5	180.0	220.0	240	1,869	1,925	5.23
C-8-14	342	200.0	182.0	224.5	295	1,329	1,415	7.24
C-8-17	343	208.0	186.0	237.5	313	1,276	983	8.18
C-9-4	320	181.0	168.0	208.0	211	2,199	1,975	7.13
C-9-7	335	190.5	177.0	213.0	240	2,083	1,956	6.09
C-9-10	326	202.0	186.0	225.5	273	2,148	1,916	5.86
C-9-13	335	206.0	186.0	225.0	309	2,033	1,809	5.02
C-9-15	337	207.5	189.0	233.5	315	2,032	1,775	5.35
C-9-18	336	213.0	196.0	237.0	312	2,139	1,793	5.26
C-10-9	334	184.5	170.0	207.5	238	2,297	2,410	8.32
C-11-4	325	160.5	150.0	179.0	199	2,144	2,499	4.52
C-11-5	312	172.0	157.0	190.5	231	2,367	2,561	21.43
C-11-7	330	176.0	162.0	194.0	231	2,192	2,566	21.43
E-1-3	190	142.5	134.0	145.0	128	2,213	663	33.74
E-1-6	232	162.5	148.0	177.5	163	2,371	1,004	10.33
E-1-9	268	171.0	160.0	189.0	185	2,473	1,459	7.84
E-1-12	320	178.5	163.0	197.5	206	2,498	1,954	5.19
E-1-14	325	176.5	163.0	181.5	215	2,720	2,468	4.03
E-1-16	330	178.5	165.0	190.0	233	2,717	2,565	4.03

TABLE A - 2
BOILER TEST CONDITIONS (Con't)

TEST	T _{b1}	T _{b5}	T _{f5}	T _{f1}	P _t	M _b	M _f	M _w
E-2-4	314	167.0	154	174.0	182	2,442	2,069	8.30
E-2-7	325	183.0	167	192.5	219	2,433	1,996	8.25
E-2-10	332	198.5	180	207.5	279	2,542	2,020	8.00
E-3-3	333	191.0	175	201.0	263	2,488	2,058	9.19
E-3-6	337	203.0	184	217.0	315	2,486	1,958	9.38
E-3-10	335	195.0	180	212.5	210	2,657	2,303	7.61
E-4-3	328	176.0	159	185.5	211	2,376	1,982	10.76
E-4-6	332	198.0	182	215.0	269	2,658	2,002	7.87
E-4-9	337	197.5	181	214.0	310	2,484	1,999	7.82
E-4-12	335	195.5	180	215.0	315	2,260	2,167	7.22
E-5-5	328	202.0	185	213.0	217	2,549	1,987	8.16
E-6-4	269	162.5	152	219.0	161	2,447	1,124	20.85
E-6-8	336	160.0	144	209.0	160	1,976	1,489	17.33
E-6-11	328	170.0	154	205.0	180	2,262	1,722	11.67
E-6-13	330	175.0	160	202.5	192	2,544	1,934	13.03
E-7-1	316	177.5	163	190.0	211	2,839	2,042	12.02
E-7-2	328	187.2	172	203.5	224	2,829	2,052	11.77

TABLE A - 3

BOILER TEST RESULTS

TEST	T _{sat}	Q _{BL}	ΔT _{BL,LM}	(UA) _{BL}	Q _{PH}	ΔT _{PH,LM}	(UA) _{PH}	Q _{SH}	ΔT _{SH,LM}	(UA) _{SH}	%ΔQ
C-2-1	132	99,219	7.8	12,687	2,214	0.94	2,355	809	48.2	16	-3.1
C-2-7	134	95,250	23.0	4,132	967	7.71	125	11,556	41.2	280	-0.7
C-2-8	134	95,250	27.3	3,487	**	**	**	24,288	22.4	1,080	-3.4
C-2-9	152	88,664	17.1	5,163	3,847	5.07	758	10,354	43.2	239	9.4
C-2-10	148	89,549	22.5	3,965	1,444	6.38	226	11,908	47.4	250	16.3
C-2-11	148	89,415	20.7	4,309	1,202	4.58	262	12,683	48.9	259	14.6
C-2-12	148	89,415	21.0	4,253	2,644	4.50	587	9,116	60.7	150	20.7
C-2-13	144	87,465	27.2	3,213	**	**	**	16,071	37.2	431	11.7
C-2-14	157	109,246	13.2	8,238	4,897	2.43	2,015	1,271	59.3	21	6.3
C-2-15	152	105,653	17.9	5,899	2,327	3.47	670	14,236	49.2	289	2.6
C-3-3	175	58,816	10.2	5,526	5,332	5.16	1,033	883	28.2	31	14.4
C-3-5	176	54,110	8.9	6,066	4,157	3.38	1,230	981	32.9	29	26.4
C-3-7	195	49,391	8.9	5,518	5,793	3.89	1,489	*	*	*	35.3
C-4-6	167	83,165	21.9	3,787	6,757	11.46	589	11,026	38.6	285	5.4
C-4-8	170	76,848	20.0	3,825	6,750	9.95	678	8,453	40.9	206	17.5
C-4-12	170	170,283	37.0	4,594	22,581	12.16	1,857	*	*	*	16.2
C-4-14	170	175,959	35.6	4,938	22,319	12.44	1,794	879	97.2	9	9.2
C-4-16	173	174,122	33.2	5,239	24,693	11.72	2,107	*	*	*	5.0
C-4-23	175	164,944	34.6	4,767	11,152	9.35	1,192	5,931	95.8	61	15.7
C-5-6	173	170,737	34.9	4,886	23,417	13.39	1,748	10,526	88.3	119	2.6
C-5-7	177	166,397	29.5	5,631	25,036	11.90	2,103	11,316	77.8	145	2.0
C-5-10	177	165,681	26.2	6,316	18,224	7.55	2,413	2,623	83.8	31	0.6
C-5-15	184	211,019	32.5	6,480	33,160	9.26	3,581	*	*	*	0.2
C-6-5	171	217,055	35.6	6,086	24,149	5.42	4,455	*	*	*	17.8
C-6-12	183	204,264	33.9	6,020	26,708	5.28	5,058	*	*	*	19.3
C-7-4	150	117,834	32.4	3,631	3,834	11.02	348	29,207	61.5	393	24.6
C-7-9	165	172,764	40.1	4,306	8,944	11.82	756	52,135	86.6	601	14.4
C-7-12	169	222,829	44.4	5,017	16,963	10.85	1,563	26,766	119.6	307	6.7
C-7-17	175	236,679	49.6	4,771	17,197	10.40	1,653	25,149	142.4	176	13.2
C-8-5	177	215,588	44.5	4,837	17,386	10.25	1,696	30,721	123.8	248	13.7
C-8-6	187	194,193	48.0	4,045	16,006	12.29	1,302	32,424	123.9	261	22.4
C-8-9	196	191,994	47.2	4,065	19,005	12.57	1,512	25,385	121.4	209	21.4
C-8-11	193	196,361	47.3	4,151	17,268	12.90	1,338	30,666	119.6	256	6.0
C-8-14	212	125,953	48.7	2,586	29,719	15.55	1,911	10,879	119.1	91	9.7
C-8-17	217	85,560	48.3	1,769	21,340	18.25	1,169	12,499	108.6	115	28.0
C-9-4	183	209,419	42.8	4,888	19,855	10.71	1,853	28,400	116.7	243	14.9

TABLE A - 3

6.3 BOILER TEST RESULTS (Con't)

TEST	T _{sat}	Q _{BL}	ΔT _{BL,LM}	(UA) _{BL}	Q _{PH}	ΔT _{PH,LM}	(UA) _{PH}	Q _{SH}	ΔT _{SH,LM}	(UA) _{SH}	%ΔQ
C-9-7	194	197,627	46.3	4,262	22,952	11.7	1,960	22,120	124.8	177	20.8
C-9-10	206	180,179	41.6	4,329	26,835	12.9	2,065	22,613	104.0	217	13.4
C-9-13	217	157,424	46.8	3,363	39,256	16.6	2,363	8,975	111.1	80	21.7
C-9-15	220	150,875	43.8	3,443	39,067	15.1	2,582	14,586	105.5	140	22.1
C-9-18	220	152,420	42.5	3,582	30,986	14.2	2,172	18,900	101.7	186	23.1
C-10-9	192	243,415	47.5	5,115	36,054	12.3	2,926	22,039	128.7	171	10.9
C-11-4	177	269,933	44.5	6,065	43,864	8.2	5,349	2,849	146.2	19	9.9
C-11-5	190	263,828	39.1	6,747	55,788	10.0	5,534	749	121.5	6	-2.2
C-11-7	190	264,352	47.0	5,624	47,429	11.3	4,197	6,005	136.5	44	0.6
E-1-3	140	80,944	17.8	4,542	2,507	5.9	425	1,725	46.9	36	
E-1-6	157	116,653	29.0	4,021	5,878	11.2	522	11,123	61.5	180	-7.5
E-1-9	170	163,513	30.9	5,279	9,635	8.2	1,172	15,395	84.4	182	-2.7
E-1-12	180	209,129	48.0	4,356	22,261	13.1	1,699	19,495	125.7	155	11.5
E-1-14	185	265,749	49.9	5,145	35,846	11.6	3,071	*	*	*	11.7
E-1-16	192	259,120	48.7	5,315	46,410	11.5	4,032	*	*	*	8.9
E-2-4	170	231,740	48.4	4,780	21,518	11.0	1,952	4,593	140.6	32	9.6
E-2-7	185	197,612	51.8	3,814	24,072	14.3	1,683	8,682	133.5	65	15.9
E-2-10	208	185,879	50.2	3,702	39,034	15.9	2,455	*	*	*	19.1
E-3-3	202	199,655	49.6	4,025	37,790	13.7	2,758	*	*	*	15.5
E-3-6	220	166,457	49.3	3,376	49,349	16.4	3,009	*	*	*	19.2
E-3-10	182	246,513	53.6	4,595	3,179	14.7	215	40,403	127.9	315	2.1
E-4-3	182	212,088	53.7	3,949	30,088	14.6	2,060	3,989	143.0	27	13.4
E-4-6	205	188,234	46.3	4,065	32,240	13.6	2,370	12,115	118.5	102	17.8
E-4-9	218	171,938	47.6	3,606	51,041	14.2	3,592	*	*	*	20.8
E-4-12	220	184,271	44.8	4,113	59,834	12.9	4,638	*	*	*	5.2
E-5-5	185	206,718	54.4	3,797	-0-	-0-	-0-	32,001	120.4	265	6.1
E-6-4	155	130,445	35.0	3,725	2,192	9.6	227	38,503	68.0	565	11.2
E-6-8	157	172,756	54.6	3,158	12,390	14.2	870	41,431	136.6	303	20.2
E-6-11	167	192,924	52.0	3,710	14,555	14.1	1,029	36,328	130.0	279	13.0
E-6-13	175	210,808	50.6	4,166	19,146	13.1	1,458	29,783	132.3	225	16.0
E-7-1	182	218,540	46.0	4,746	26,000	11.9	2,181	9,395	127.4	73	17.8
E-7-2	187	213,429	49.2	4,330	20,932	13.4	1,558	19,639	127.2	154	19.0

* T_{sat} higher than T_{f1}.

** Slightly negative preheat. Possible flashing downstream of CV-110.

SAMPLE DATA REDUCTION SHEET

TEST C-9-18

PREHEATER

fluid properties

BRINE

$$\rho_b @ T_{b1} (336^\circ\text{F}) = 55.9 \text{ lbm/ft}^3 \quad \text{GPM} = (4.0) \frac{8.617(1.035^*)}{\sqrt{55.9}} = 4.77 \text{ GPM}$$

$$\rho_f @ T_{f3} (153^\circ\text{F}) = 30.8 \text{ lbm/ft}^3 \quad \text{CFS} = \text{GPM}/448.8 = 0.0106 \text{ CFS}$$

$$\rho_w @ T_{f3} (153) = 61.2 \text{ lbm/ft}^3 \quad V_b = \text{CFS}/.2006 = 0.0530 \text{ Ft/sec}$$

$$c_{pb} @ \text{Avg. } T_{b3,4} (195^\circ\text{F}) = 1.01 \text{ BTU/lbm}^\circ\text{F} \quad M_b = (\text{GPM})(8.02)(\quad) = 2139 \text{ lbm/hr}$$

$$c_{pb} @ \text{Avg. } T_{b1,5} (274.5^\circ\text{F}) = 1.02 \text{ BTU/lbm}^\circ\text{F} \quad T_{b4} - T_{b3} = 36^\circ\text{F}$$

$$c_{pf} @ \text{Avg. } T_{f3,4} (174.5^\circ\text{F}) = 0.68 \text{ BTU/lbm}^\circ\text{F} \quad Q_b = (2139)(1.01)(36) = 77,779 \text{ BTU/hr}$$

$$c_{pf} @ T_{f4} (196^\circ\text{F}) = 0.72 \text{ BTU/lbm}^\circ\text{F} \quad \Delta T_{LM} = (24-17)/\ln(24/17) = 20.30$$

ISOBUTANE

*Rotameter float correction

$$\text{GPM} = (7.5) \frac{5.345(1.005^*)}{\sqrt{30.8}} = 7.26 \text{ GPM} \quad UA = Q_f/\Delta T_{LM} = 2583 \text{ BTU/hr }^\circ\text{F}$$

$$\text{CFS} = \text{GPM}/448.8 = 0.0162 \text{ CFS} \quad U_v = UA/1.2 \text{ cu ft} = 2152 \text{ BTU/hr ft }^\circ\text{F}$$

$$V_f = \text{CFS}/0.2006 = 0.0806 \text{ ft/sec.} \quad C_f = (c_{pf})(M_f) = (0.68)(1793) = 1219 \text{ Btu/hr}^\circ\text{F}$$

$$M_f = (\text{GPM})(8.02)(30.8) = 1,793 \text{ lbm/hr} \quad \text{NTU} = UA/C_f = 2.12$$

$$T_{f4} - T_{f3} = (196 - 153) = 43^\circ\text{F} \quad \Delta Q = Q_b - Q_f = 25,346 \text{ Btu/hr}$$

$$Q_f = (1793)(0.68)(43) = 52,433 \text{ btu/hr} \quad \% \Delta Q = (\Delta Q/Q_b)100 = 33\%$$

SAMPLE DATA REDUCTION SHEET

TEST

BOILER

$$\begin{aligned} P_T &= \underline{297} \text{ psig} = \underline{312} \text{ psia} \\ M_w &= \underline{\quad (ml/15 \text{ min}) (.00014) \quad} (\rho_w) = \\ &\quad (\underline{615}) (.00014) (\underline{61.2}) = \underline{5.27 \text{ lbm/hr}} \\ \% \text{ by wt} &= (M_w/M_f)100 = \underline{0.29\%} \\ X_i &= (\text{From curve}) = \underline{0.005} \\ P_w &= (P_t)(X_i) = (\underline{312}) (\underline{.005}) = \underline{1.56 \text{ psia}} \\ P_{sat} &= (\underline{312} - \underline{1.56}) = \underline{310} \\ T_{sat} @ P_{sat} &= \underline{220^\circ\text{F}} \\ h_{fg,w} @ P_w &= \underline{1028 \text{ BTU/lbm}} \\ Q_w &= (M_w)(h_{fg,w}) = \underline{5417 \text{ BTU/hr}} \\ \\ h_{g,f} @ T_{sat} &= \underline{165 \text{ BTU/lbm}} \\ h_{f,f} @ T_{sat} &= \underline{80 \text{ BTU/lbm}} \\ h_{fg,f} &= (\underline{165} - \underline{180}) = \underline{85 \text{ BTU/lbm}} \\ Q_{BL} &= (h_{fg,f})(M_f) = \underline{152,421 \text{ BTU/hr}} \\ T_{sat} - T_{f_5} &= \underline{24^\circ\text{F}} \\ Q_{PH} &= (T_{sat} - T_{fs})(c_{pf})(M_f) = \underline{30,986 \text{ BTU/hr}} \\ T_{f1} - T_{sat} &= \underline{17^\circ\text{F}} \\ Q_{SH} &= (T_{f1} - T_{sat})(c_{pf})(M_f) = \underline{18,900 \text{ BTU/hr}} \\ \Sigma Q_f &= Q_{BL} + Q_{PH} + Q_{SH} = \underline{202,307 \text{ BTU/hr}} \\ T_b' &= (Q_{PH}/\Sigma Q_f)(T_{b_1} - T_{b_5}) + T_{b_5} = \underline{231.8^\circ\text{F}} \\ T_b'' &= T_{b_1} - (Q_{SH}/\Sigma Q_f)(T_{b_1} - T_{b_5}) = \underline{324.5^\circ\text{F}} \\ T_{LM,BL} &= (104-12) / \ln(104 / 12) = \underline{42.6^\circ\text{F}} \end{aligned}$$

SAMPLE DATA REDUCTION SHEET

TEST

BOILER

$$\Delta T_{LM,PH} = (17-12) / \ln(17 / 12) = \underline{14.3}$$

$$\Delta T_{LM,SH} = (105-99) / \ln(105 / 99) = \underline{101.7}$$

$$U_{BL} = Q_{BL} / T_{LM,BL} = \underline{3.582 \text{ BTU/hr } ^\circ\text{F}}$$

$$U_{APH} = Q_{PH} / T_{LM,PH} = \underline{2.173 \text{ BTU/hr } ^\circ\text{F}}$$

$$U_{ASH} = Q_{SH} / T_{LM,SH} = \underline{186 \text{ BTU/hr } ^\circ\text{F}}$$

$$Q_b = (T_{b1} - T_{b5})(c_{pb})(M_b) - Q_w = (123)(1.02)(2139) - (5417) = \underline{262,958}$$

$$\Delta Q = Q_b - \Sigma Q_f = \underline{60,652 \text{ BTU/hr}}$$

$$\Delta\% Q = (\Delta Q / Q_b) 100 = \underline{23.1\%}$$

APPENDIX B

FIGURE 8-1

SOLUBILITY OF ISOBUTANE
IN AQUEOUS SODIUM CHLORIDE SOLUTIONS
(REFERENCE 33)

WEIGHT PERCENT NaCl

25

20

15

10

5

0

TEMPERATURE °F

K&E SEMI-LOGARITHMIC 46 4972
2 CYCLES X 70 DIVISIONS
KETCHUM & EBER CO.

HENRY'S CONSTANT $[PSF/LB-MOLE/LBM BRINE] \times 10^{-10}$

APPENDIX C
P&I DIAGRAM AND
VESSEL DRAWINGS

APPROVALS		TITLE	
DESIGNED	DATE	DIRECT CONTACT HEAT EXCHANGER TEST P&I DIAGRAM	
CHECKED	BY		
REVIEW			
ELECT			
CIVIL			
PROJECT ENGINEER		DRAWING NO. 051-6101	
P.E. REG. NO.		SCALE NONE SHEET 1 OF 2

WELD CONNECTION DETAILS

NO SCALE

HALF OR FULL COUPLING CONNECTION

FLANGE CONNECTION

BUTT WELD CONNECTION

PIPE THRU VESSEL CONNECTION

CONE CONNECTION

NOTES:

1. NOTES APPLY TO ALL DRAWINGS.
2. ALL WELDS AND WELDWORK TO BE IN ACCORDANCE WITH LATEST ASME BOILER & PRESSURE VESSEL CODE SECTION VIII, DIV. 1 FOR DESIGN PRESSURES AND TEMPERATURES SHOWN ON CODE STAMPS HEREWITH. VESSELS TO BE HYDROSTATICALLY TESTED. CERTIFIED TEST SHEETS TO BE FURNISHED.
3. GAGE GLASS CONNECTION ALIGNMENT IN VERTICAL PLANE (ALONG LONGITUDINAL AXIS) MUST BE TRUE.
4. ALL THREADED HALF COUPLINGS MUST BE FURNISHED WITH FORKS.
5. ALL THREADED AND SOCKET WELD FITTINGS ARE TO BE OF POWERED STEEL CONFORMING TO ASTM A-302 GRADE 1, ASA D36.11.
6. ALL PIPE EITHER TO BE SEAMLESS CARBON STEEL TO ASTM A305 GR. B, OR WELDED CARBON STEEL TO API 5L GR. B, OR EQUIV.
7. ALL STRUCTURAL STEEL TO ASTM A36 OR EQUAL.
8. DETAILS SHOWN ON THIS SHEET APPLY TO ALL SHEETS.
9. WHEN WELDED PIPE IS USED ALL CONNECTIONS ARE TO BE LOCATED OFF THE LONGITUDINAL WELD. THIS NOTE ALSO APPLIES TO WELDED CODE REDUCTIONS.
10. ALL GASKET SURFACES TO BE 125 FINISH FLAT & TRUE. ALL FLANGED SURFACES TO HAVE RAISED FACES.
11. COMPRESSED ASBESTOS GASKETS TO BE FURNISHED AT ALL PLANNED JOINTS 1/16" THICKNESS PER ASME D36.21 TABLE 1.
12. BOLTING SHALL BE ALLOW STEEL CLASS ASTM A305 GR. B7, EACH WITH TWO HEX NUTS ASTM A304 CL.28.

~~8-19-75 GENERAL REVISION NOTES~~
~~7-19-75 GENERAL REVISION NOTES~~

DBS ENGINEERS, Inc.

APPROVALS		DATA	
DESIGNED BY	CK 7/75	PRESSURE VESSELS	
DESIGNED BY	DM 7/75	FOR	
DESIGNED BY	OLYMPIAN 7/75	DIRECT CONTACT	
DESIGNED BY		TEST UNIT	
DESIGNED BY		PROJECT NO.	
DESIGNED BY		051-2201	
DESIGNED BY		REV. D	
P. E. OGDEN, INC.		AKA. MARYLAND

PRE HEATER E-100

DESIGN PRESSURE ——— 950 P.S.I.A. Δ

DESIGN TEMPERATURE—220°F.

SECTION 'A-A'

SECTION 'B-B' & 'C-C'

8-14-75 GENERAL REVISIONS NOTED ▲
7-29-75 GENERAL REVISIONS NOTED ▲

DBS ENGINEERS, Inc.

APPROVALS		DATE	
DESIGNED BY	DATE	PRESSURE VESSELS FOR DIRECT CONTACT TEST UNIT	
CHECKED BY	DATE		
DESIGNED BY	DATE		
CHECKED BY	DATE		
DATE	DATE		
PROJECT NUMBER		DATE	
001-1000-001		051-2202	
001-1000-001		001-1000-001

7-29-75 GENERAL REVISIONS NOTED
3-14-75 DELETE CYCLONE SEPARATOR

DSS ENGINEERS, INC.			
APPROVALS	DATE	BY	FOR
DESIGN	02/11/75	JTB	PRESSURE VESSELS
CHECK	02/11/75	JTB	FOR
CONSTRUCTION	02/11/75	JTB	DIRECT CONTACT
TEST			TEST UNIT
PROJECT NUMBER	051-2203		
DATE	02/11/75

SECTION 'A-A'

DESIGN PRESSURE — 220 PSIA
DESIGN TEMPERATURE — 185 °F.

HOTWELL V-100

SECTION 'A-A'

BOILER E-101

DESIGN PRESSURE — 314 PSIA
DESIGN TEMPERATURE — 325 °F.

8-14-78 GENERAL REVISIONS NOTED
7-27-78 GENERAL REVISIONS NOTED

DSS ENGINEERS, Inc.

APPROVALS		DATE	FOR
DESIGNED BY	PK	1-18	PRESSURE VESSELS FOR DIRECT CONTACT TEST UNIT
CHECKED BY	QJM	1/18	
DATE	1/18/78		
PROJECT NUMBER			
DRAWN BY			
P.E. DES. NO.		051-2204	B

APPENDIX D

EFFECTS OF WATER VAPOR

D.1 INTRODUCTION

In vaporizing the isobutane in the direct contact boiler, the total pressure of the vapor would theoretically be the sum of the vapor pressures of brine and isobutane at the selected temperature. Then, if this acts as an ideal gas and has reached equilibrium, it will contain a weight fraction of water vapor proportional to its partial pressure and molecular weight.

With these conditions the weight of the water vapor in the mixture would be 1.2% to 1.7% of the working fluid. However, because of the considerably high latent heat of vaporation, the formation of this water vapor takes 5% to 15% of the total heat supplied by the brine. To counter balance this, the water vapor in the mixture from the boiler will produce work as it expands through the turbine.

The analytical expressions used for the cycle studies reported herein (see paragraph 2.3 and 4.1.2) do not include factors to account for water vapor produced from the geothermal brine. Also, the cycle state diagrams, figures 2-1 and 4-2 do not reflect the water vapor effect. Therefore, it is worthwhile to set forth here some of the possible effects that this may have on system design and optimization procedures.

D.2 EXPERIMENTAL WORK

The theoretical amount of water vapor can be calculated assuming an ideal gas and equilibrium conditions. However, since these assumptions may not be valid, the test work included measurement of the water vapor condensed in the isobutane. This is explained in paragraph 3.4.4, page 67 of this report. The measured amounts of water vapor were taken into account in analyzing the test data and in determining the boiler performance.

It can be seen from figure 3-12 that the measured amount of water vapor was considerably less than the calculated theoretical amount of vapor at high flow rates. Discrepancies in the test data indicates that all the water vapor may not have been collected. Thus, a correlation for amount of water vapor produced is a % of the theoretical could not be established. Further testing will be done with more accurate collection and measurement means to establish this.

D.3 EFFECT ON CYCLE DESIGN

If the total theoretical amount of water vapor is produced in the boiler this would change the cycle state points since the slope of the brine T vs. H line would change. This change for the conceptual design is shown in figure 4-2A. From this it can be seen that a smaller pinch point would be used and the distribution of heat transferred in each preheater would change but the total size of the preheaters remain the same. In this example, the exit brine temperature increases slightly from 152°F to about 158°F. At the same time, additional turbine work is obtained from the water vapor. Thus, there is very little decrease in the overall cycle effectiveness.

FIGURE 4-2A
CONCEPTUAL DESIGN

ISOBUTANE CYCLE STATE POINTS

At lower boiling temperatures and for higher brine temperatures, the effect of water vapor produced on the brine T vs H curve is less. In some cases there would be an actual increase in the overall effectiveness due to water vapor. Thus, consideration of the water vapor may have more significance in determining the optimum operating points.

D.4

FUTURE WORK

Continued development of direct contact heat exchangers under Phase II of the work includes expanding the analytical expressions to include water vapor and then using these in a computer model for cycle analysis and optimization.

Additional experiments will be done to accurately measure the water vapor in the isobutane vapor. The results can be utilized to modify the theoretical expressions developed.

The final expressions and resulting analysis will then be the basis for overall system economic optimization.

APPENDIX E

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