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THEORY AND PRACTICE OF NEAR CRITICAL
PRESSURE DIRECT CONTACT HEAT EXCHANGE

Final Report

By
Edward F. Wahl
Frederic B. Boucher

Date Published - October 1977

Work Performed Under Contract No. E (04-3)-1076

MASTER

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ENERGY RESEARCH AND DEVELOPMENT ADMINISTRATION

Division of Geothermal Energy

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FINAL REPORT
THEORY AND PRACTICE OF NEAR CRITICAL PRESSURE
DIRECT CONTACT HEAT EXCHANGE

Prepared for:
Energy Research and Development Administration
Under Contract No. E(04-3)-1076

By:
OCCIDENTAL RESEARCH CORPORATION
1855 Carrion Road
La Verne, California 91750

October, 1977

E.F. Wahl
F.B. Boucher

Abstract

The direct contact heat exchange (DCHE) system designed, developed, and tested by ORC operated successfully without scale deposition. Conceptual design, optimization and cost analysis showed that an additional advantage is the reduction in cost due to the high capacity of the DCHE compared with conventional heat exchangers.

The technical and economic feasibility of DCHE operated near or above the critical pressure of isobutane using isobutane in direct contact with geothermal brine was assessed. The program consisted of conceptual design with a preliminary process evaluation and feasibility analysis, design and construction of the experimental apparatus, experimental tests, a final revised process evaluation and feasibility analysis based on the experimental results, and preparation of a design procedure.

Countercurrent direct contact heat exchange systems using packed or sieve tray columns, or mixer-settler cascades are the most promising processes for large commercial applications. A mixer-settler cascade may have a cost advantage for the supercritical isobutane system over other direct contact heat exchange devices.

No scaling problems were present in the direct contact heat exchanger or the test unit. Test runs show that no scale was deposited in either the packed column section or the sieve tray section for continuous runs of 76 and 160 hours using East Mesa well 6-2 brine. In fact, the sieve tray section after 160 hours operation was visibly cleaner after the run than before. Both the packed column section and the sieve tray section operated according to theoretical thermodynamic and hydraulic predictions. Operation of the supercritical isobutane direct contact heat exchange and associated test unit with geothermal brine was routine with automatic controls and without operator attention.

A design procedure including the requisite equations for sieve tray direct contact heat exchange columns was developed. The procedure gives the optimum ratio of isobutane to brine, optimum number of trays and tray spacing, and the sieve tray layout variables, namely downcomer area, active hole area, bubbling area, and column cross sectional area.

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INTRODUCTION

In December, 1974, ERDA (then AEC) issued a request for proposal to assess the technical and economic feasibility of direct contact heat exchange for geothermal applications, and to provide an impartial analysis identifying any engineering problems requiring particular attention in future R&D activities. The geothermal brine was to be in contact with a working fluid, which would vaporize and be superheated by transfer of heat from the brine. One particular application of direct contact heat exchange for geothermal energy recovery is for the production of electricity. In such a system the working fluid is expanded through an expander-generator to produce electricity and is then cooled, condensed, and returned to the direct contact heat exchanger as in a conventional power cycle.

ERDA contract E(04-3)-1076 was awarded to Occidental Research Corporation (ORC) on July 1, 1976, to study the theory and practice of near critical pressure direct contact heat exchange using isobutane as the working fluid. This near critical concept was based on prior development by ORC. The scope of work for this study is limited to the operation of the direct contact heat exchanger. The steps of using the working fluid in an expander-generator to produce energy, cooling and condensing the working fluid after being used in the expander-generator, and the recycling of the working fluid back to the direct contact heat exchanger is excluded from the scope of work under the contract.

SUMMARY

The technical and economic feasibility of direct contact heat exchange (DCHE) operated near or above the critical pressure of isobutane was assessed for isobutane in direct contact with geothermal brine. The program consisted of conceptual design

with a preliminary process evaluation and feasibility analysis, design and construction of the experimental apparatus, experimental tests, a final revised process evaluation and feasibility analysis based on the experimental results, and preparation of a DCHE design procedure.

To minimize the capital cost of direct contact heat exchangers in liquid-liquid heat transfer, the DCHE is designed for maximum throughput. Of column type DCHE, sieve plate and packed columns can be designed to achieve the highest throughput of brine and isobutane. Consequently, for this study two 8 inch I.D. DCHE columns were operated intermittently for performance data and continuously for scale deposition data using USBR's East Mesa Well 6-2. A 10 foot high column containing 40 inches of Goodloe stainless steel packing was operated 76 continuous hours and intermittently. At the conclusion of the 76-hour run the column was inspected and was found free of scale deposits. The other, an 18 foot high column containing 12 carbon steel sieve plates with 1 foot spacing, was operated continuously for seven days. No scale deposition was present and the column was actually cleaner after the run than before. Both the continuous and intermittent operation of either column and the test apparatus was trouble-free without scale deposition. Steady operation was maintained by the automatic control system for both subcritical and supercritical operation.

Typically the geothermal brine entered the top of the column at 330°F and heated isobutane entering the bottom of the column from 80°F to 290°F at the top of the column. Column pressure was varied from 500 to 650 psia. The packed column was operated at rates up to 2000 lb/hr (5730 lb/hr ft²) and the sieve plate column at rates up to 4800 lb/hr (14,000 lb/hr ft²). The efficiency of the sieve plate column was typically 75% and varied from 50 to 100%. Flooding, the column brine flow rate limit, occurred at 5000 lb/hr (14,500 lb/hr ft²).

A computer model based on stagewise energy and material balances of the DCHE, throughput estimation, stage efficiency, conventional cost prediction methods, and work generation equations was used to predict the power output and cost of electricity for the DCHE system. The calculations of the thermal efficiency and power output show that for more than eight theoretical stages the increase in efficiency and net output become quite small. The cost analysis, based on the equations for column flooding velocity, and size, shows that the minimum cost of electricity occurred in the 5 to 8 stage range, depending on the heat exchanger system. This would correspond to seven to eleven 75% efficient plates.

Assuming a working fluid such as isobutane and a column type such as a sieve plate column have been selected, a design procedure was developed consisting of selecting the continuous phase, designing the sieve plate configuration, and determining the number of plates. The continuous phase liquid is selected so as to minimize scale deposition and maximize throughput consistent with satisfactory tray efficiency. For the layout design, either column area or brine throughput is set and the design equations are solved stepwise for the downcomer area, active hole area, bubbling area, column cross sectional area, brine mass flow rate, and isobutane volumetric rate. The hole diameter, number of holes, downcomer area, and tray spacing are specified so as to maximize throughput. The optimum number of plates is determined by balancing the cost of the column, which increases with increasing number of plates, against the increased work output so as to minimize the cost of electricity.

CONCLUSIONS

1. All pilot plant data has been obtained that is needed to design a sieve plate direct contact heat exchanger (DCHE) and associated auxiliary equipment for any commercial size power plant.

2. The capital cost of the DCHE is a small part of the total cost of a geothermal powered electricity generating facility, typically about 10-15% of the cost of conventional shell and tube heat exchangers for a comparable "binary" system.

3. Scaling does not occur in the direct contact heat exchanger or anywhere in the test unit using East Mesa well 6-2 brine which has a TDS of 4380 ppm, contains CO_2 , and exhibits CaCO_3 and silica-silicate deposition.

4. Noncondensibles from the East Mesa reservoir are handled by the system and no venting facilities are necessary because of the nature of the DCHE system.

5. Of the various countercurrent direct contact heat exchange systems, packed or sieve tray columns or mixer-settler cascades result in minimum cost for large commercial applications.

6. Based on the conceptual design study, a mixer-settler cascade may have a cost advantage for the supercritical isobutane system over other direct contact heat exchange devices.

7. Operation of the direct contact heat exchanger and associated equipment at supercritical conditions poses no unusually difficult design or operational problems involving liquid-liquid thermodynamics and hydraulics.

8. The design procedures for sieve tray direct contact heat exchange columns predict actual performance of any commercial size direct contact heat exchange sieve plate column.

9. A capacity factor (operating factor) of 90-95% can be expected for commercial DCHE units because of the absence of scaling problems and trouble free operation of the DCHE unit, and because the plant is similar to petrochemical or petroleum refinery operation where these high capacity factors are commonly achieved.

RECOMMENDATIONS

1. Steps should be initiated to fund a program for the design, construction, test and evaluation of a 50 megawatt prototype electricity generating plant using the direct contact heat exchanger designed as described in this report. The program should include:

- i. Selection of a suitable geothermal reservoir. For the demonstration plant, the reservoir should be representative with perhaps 1-5 wt % TDS and with a temperature both high enough to be a good candidate for generating electricity and low enough to have a reasonable incidence of occurrence, probably in the 400-500°F range.
- ii. Operation of the DCHE test unit at the reservoir selected. It is not possible to predict the reactions of the complex brine chemistry with certainty. Therefore, before designing the prototype plant, the test unit should be operated using the reservoir fluid at the site selected to ensure that scale deposition, if any, and non-condensibles are not a problem.
- iii. Design of the total system using the procedure described in Recommendation 3, including selection of the working fluid, DCHE operating pressure and working fluid recovery stripper pressure. The demonstration plant should be designed to minimize the cost of electricity.
- iv. Design evaluation.
- v. Construction and operations.

- vi. An overall economic evaluation of the secondary working fluid system. After operation of the commercial size prototype plant, a thorough economic analysis of the system is needed to put the value of electricity generation from geothermal brine into perspective with regard to the total energy picture and in relation to other geothermal powered generating systems.

2. The existing ORC test unit should be tested using geothermal fluids from other reservoirs to verify the universal application of the isobutane direct contact heat exchange system. The sequence of reservoir tests should be:

- i. a 400-500°F, less than 3% TDS brine so that the ability of the system to handle a high silica depositing brine can be observed.
- ii. a 400-500°F, about 5% TDS reservoir to observe the ability of the system to handle medium high TDS brines.
- iii. a 400-500°F, about 5% TDS, high carbon dioxide and/or other noncondensibles reservoir to observe the ability of the DCHE to handle a complex system consisting of the elements studied above together with carbon dioxide already studied using the East Mesa brine.
- iv. a 400-600°F, about 10% TDS reservoir such as the Niland reservoir which represents one of the most difficult to handle geothermal reservoirs.

On completion of the tests, an elementary feasibility analysis with mention of general economics should be prepared based on the existing economic data and the test results.

3. Total system design procedures should be developed to optimize the cost of electricity. The fundamental work equations developed by Wahl (1977) and the equations developed along with the system analysis computer program, together or separately provide a basis for the system design procedure. Also, they will become elements of the system design procedure. Some objectives of the system design are:

- i. Selection of the secondary working fluid to minimize the lost work in the DCHE.
- ii. Optimization of the DCHE operating pressure.
- iii. Optimization of the working fluid recovery system stripping pressure.

4. The mixer-settler system should be studied since this may be the most economic and reliable system for direct contact heat exchange of geothermal brines. A limited test should be undertaken using a simple modification of the present test unit.

DISCUSSION

1. Objectives

The principal objective of this work was to assess the technical and economic feasibility of near supercritical direct contact heat exchange for geothermal applications. This objective includes design procedures for a commercial size direct contact heat exchanger, economic and technical feasibility analyses of direct contact heat exchange for geothermal applications and an assessment of technical areas which may need further investigation.

The above information was to be obtained by selecting the best conceptual direct contact heat exchanger, then designing, constructing, and operating a pilot scale direct contact heat exchanger to obtain the experimental data necessary for the technical assessment, and finally preparing the assessment of the technical and economic feasibility. In addition, design procedures were to be developed, engineering problems were to be identified, and a program of follow on R&D was to be prepared. The above objectives are summarized in Table 1-1, together with the task name described in Section 2, Plan.

Assessment of Feasibility

The feasibility of using a direct contact heat exchanger to recover energy from geothermal brines was to be determined with respect to a power cycle in which the heated working fluid is expanded through an expander-generator for the production of electricity and the working fluid cooled, condensed, and returned to the direct contact heat exchanger.

Selection of Conceptual Design. The scale model direct contact heat exchanger to be designed and constructed under this contract was to include trays to increase the heat transfer efficiency of the scale model direct contact heat exchanger. Tray design and spacing was to be evaluated to determine heat transfer rates, approach temperatures between hot aqueous liquids and working fluids, and working fluid losses. Backmixing and entrainment of working fluid was also to be evaluated.

Table 1-1

Objectives and Sub-Objectives with Associated Tasks

<u>Objectives</u>	<u>Tasks</u>
Assessment of Technical and Economic Feasibility	Feasibility analysis
Selection of best conceptual designs	Conceptual design
Experimental data for assessment	Design, construction, and operation of pilot scale DCHE
DCHE design procedures	Development of design procedures concurrent with conceptual design
Identification of DCHE engineering problems	Concurrent with above tasks
Program for follow-on R&D	Concurrent with above tasks

Experimental Data for Assessment of Feasibility. The direct contact heat exchanger was to be operated such that a working fluid in contact with a 250-350°F aqueous liquid in the direct contact heat exchanger would be heated at elevated pressure near the critical pressure to produce a suitable fluid for expansion through an expander-generator to produce electricity.

The experimental heat exchanger and test unit was to be installed at the ORC facilities in La Verne. The unit was to be debugged, calibrated, and operating procedures established. Then a series of 4 one-day runs were to be conducted using fresh water within a pressure range of 500 to 650 psia, a water temperature range of 200 to 300°F, isobutane to water ratios of 0.5 to 1.5, and water and isobutane flow rates of 1 to 15 gpm.

Then the test unit was to be transported to the East Mesa test site of the Bureau of Reclamation and installed. The experimental program there was to consist of three test runs. Each test run was to consist of (1) one day of operations at varying pressures and flow rates to obtain efficiency and capacity data, (2) continuous operation at near optimum conditions for 8 days to determine the scale deposition characteristics and (3) examination for scale deposition. The exact nature of the runs were to be determined after completion of prior work.

Development of Direct Contact Heat Exchanger Design Procedures

Design procedures for a direct contact heat exchanger were to be based on the use of the direct contact heat exchanger in a power cycle in which the heated working fluid is expanded through an expander-generator for the production of electricity and the working fluid cooled, condensed, and returned to the direct contact heat exchanger. The design procedures were to include

1. procedures to determine fluid circulation rates,
2. procedures to determine tray details and tray spacing, with consideration being given to heat transfer, flooding, and scaling,

3. procedures to determine the diameter required to avoid flooding and to maximize heat transfer at various levels in the exchanger,
4. procedures to determine the number of actual trays and column height,
5. procedures to determine the inlet distributor details, and
6. procedures to determine the exchanger design details required to minimize entrainment of geothermal brine in working fluid and working fluid in geothermal brine for the purpose of maintaining satisfactory thermodynamic efficiency.

Identification of Direct Contact Heat Exchanger Engineering Problems

The engineering problems associated with the design, construction and use of a DCHE of the type tested in this work were to be identified. The use to be considered was the recovery of thermal energy from geothermal brines for the purpose of generating electricity.

Program for Follow-on R&D

Additional development work relating to the use of a direct contact heat exchanger for the recovery of energy from geothermal brine was to be recommended. The recommendation was to include an estimate of the time and the costs for such developments.

DISCUSSION

2. Plan

The plan which was followed to reach the objectives is shown in Figure 2-1 and consisted of four phases: (1) conceptual design for selection of column type and internals to be evaluated, (2) design and construction of the experimental apparatus, (3) the experimental program to obtain performance data, and (4) process evaluation and feasibility analysis based on the final results of the R&D program.

Conceptual Design

In the conceptual design phase, various concepts for recovering thermal energy from geothermal brine with a working fluid near the critical region are evolved. To select the most likely candidates, for evaluation, a process analysis equivalent to that required for the feasibility analysis is needed to estimate the relative merits of the candidates. This evaluation requires an analysis of the column performance. Thus, the first phase of the project includes a preliminary process evaluation and feasibility analysis and so a portion of the fourth phase is completed simultaneously with the conceptual design phase as shown in Figure 2-1.

Design and Construction of the Experimental Apparatus

The design of the test apparatus is begun prior to completion of the previous phase so that the overall time for completion of the project can be minimized. The apparatus is to be designed for mobility and to operate with existing USBR facilities at East Mesa.

Experimental Program

The apparatus is to be tested at La Verne and limited operating data obtained using fresh water. Then the experimental program, which includes 8-day scaling tests, will be carried out using geothermal brine from East Mesa Well 6-2. Because of the success

Figure 2-1. Schematic representation of the program plan.

of the first two runs, the third scheduled scale test run is replaced with a partial parametric study of the sieve plate column.

Process Evaluation and Feasibility Analysis

This analysis is partially carried out concurrently with Phase 1 and is completed following the experimental program. This phase includes preparation of the design procedures, since they are required for a process evaluation. Also, as a byproduct of the feasibility analysis and process evaluation, recommendations for future work are generated.

Schedule

The schedule which was followed in this program is shown in Table 2-1. Two significant delays outside the control of ORC occurred causing delays totaling 10 months respectively. The program took 23 months to complete, which after subtracting the 10 months of delay meant the actual technical effort required 13 months.

Table 2-1
Program Schedule

Conceptual Design	July 1, 1975 to Sept. 15, 1975
Design and Construction of Experimental Apparatus	July 15, 1975 to Sept. 15, 1975 (See Note 1) Feb 1, 1976 to July 1, 1976
Experimental Tests, La Verne	July 1, 1976 to Aug 31, 1976
" " La Verne (extended)	Aug. 31, 1976 to Nov. 1, 1976 (See Note 2)
" " East Mesa	Jan 26, 1977 to March 4, 1977
Process Evaluation and Feasibility Analysis	March 4, 1977 to May 31, 1977

Note 1: Because of progress in the conceptual design phase resulting in larger throughputs in the column and because of unanticipated test apparatus costs, the budget was not adequate and the project was delayed From Sept 15, 1975 to approximately Feb. 1, 1976 until new budgeting was resolved.

Note 2: Because the East Mesa ERDA test pad facility was delayed on several occasions, extended testing at La Verne was executed and finally terminated until permission to operate was finally received from Lawrence Berkeley Laboratory, the ERDA test facility operators, on January 26, 1976.

DISCUSSION

3. Literature Survey

There have been a considerable number of papers, for example Cortez et. al. (1973), Kihara and Fukunaga (1975), Chou et. al. (1974), Sherwood (1976) and Sheinbaum (1975), in the recent geothermal literature discussing the performance, including costs, of a secondary working fluid based on estimated heat exchanger performance. Since the work reported herein is concerned with supercritical pressure direct contact heat exchange, the literature survey reported is limited to the literature dealing with liquid-liquid heat transfer in direct counter-current flow.

Markowitz and Bergles (1970) present data for heat transfer between dispersed oil and water phases throughout a counter-current spray column that substantiates a model proposed by Letan and Kehat (1965). The data of these researchers show that for a column operating with interacting dispersed liquid droplets in what they call the multiple drop mode, there is a large amount of heat transfer at either end of the column due to turbulence. This turbulent mixing of the two phases in the region of contact where there is substantial backmixing causes large irreversible temperature losses in the column operation. This is because the streams are quickly equilibrated in the contact dispersion region with respect to temperature so that there is little heat transfer throughout the rest of the column. Thus the column is in effect acting as a one to two equilibrium stage contactor.

Markowitz and Bergles (1970) used 30 gauge copper constantine thermocouples with a sensitive bead 0.02" in diameter inserted directly in the column to measure temperature of the oil droplets as they flowed over the sensitive thermocouple region. They measured volumetric heat transfer coefficients in

the neighborhood of 100-600 Btu/hr. ft³ °F. In their multiple drop tests, the drops were generated at the base of the column with the distributor plate consisting of 48 nozzles 1/8" OD by 0.061" ID protruding about 1/4" from the plate surface.

The former researchers comment that "the rate of heat transfer from a drop is higher for the single drop and flooded mode than for the multiple drop mode. However, the single drop mode is obviously inefficient for transferring considerable amounts of heat, and the flooded mode is difficult to control, with the result that the multiple drop mode is of the most practical interest." This means that attention must be given to achieving relatively large counter-current flow rates with relatively low backmixing so that true counter-current operation of the column can be achieved.

The difficulty in achieving counter-current flow in the spray column as described above could be due to one of, or a combination of, three causes. The most likely explanation is that the heat transfer rate in the middle of the column is so low relative to the heat transfer rate in the turbulent regions at either end that all of the heat is effectively transferred at either end of the column in two stages. Thus the column, for practical lengths, acts effectively as a two-stage column. This can be overcome by increasing the rate of heat transfer in the middle of the column as, for example, by mixing devices. Another possibility is that of backmixing due to the bulk phase liquid, that is the water, being carried up the column by the drag of the liquid droplets. The third is that the liquid droplets are not thermally homogeneous. That is, if the outer surface of the liquid droplet is at the same temperature as the liquid, then the heat will be transferred only by conduction. The result will be that the lower temperature portion of the liquid

droplet will be carried up the column and then absorb heat in a higher temperature region. This will have the same overall effect as backmixing, namely to cause irreversible losses by non-countercurrent operation.

Steinmeyer and Woodward (1969) ran a three foot ID carbon steel vessel with six feet between distributors to transfer heat between a chlorinated biphenyl and water. These tests confirmed the above described theories because the heat transfer was accomplished as a one or two stage device. Two-stage performance was obtained when an interface was maintained between the two distributors.

Fair (1972) gives design methods for estimating the performance of direct contact heat transfer spray columns. His equations for the volumetric heat transfer coefficient U_g for the gas phase is

$$U_g = 0.043 G^{0.8} L^{0.4} / Z_t^{0.5} \quad (3-1)$$

The heat transfer coefficient for pentane drops evaporating in water (Sideman and Taitel, 1964) was found to be 200-400 for 3-1/2mm drops and 400-700 for 2mm drops. The heat transfer coefficient for non-evaporating drops was determined to be 64. The velocity of evaporating drops was found to be about twice that of non-evaporating drops which had velocities of 13-17 cm/sec.

Simpson et. al. (1974) present experimental results on the evaporation of liquid butane droplets in water for the purpose of freezing water. The instantaneous local velocities varied from 25 cm/sec. at lower diameter ratios to a velocity which increased with the square root of the diameter ratio of the bubble size to the size of the initial liquid bubble. They reported heat transfer coefficients of 400 Btu/hr.ft³ °F.

DISCUSSION

4. Conceptual Design

To transfer heat efficiently a countercurrent device must be used to achieve as high a temperature of the contacting liquid as possible. More correctly, the temperature differences in the heat exchanger must be minimized since the ΔT represents an irreversible process, and to achieve maximum efficiency, the process must approach ideal, reversible conditions as closely as possible. To achieve this, countercurrent heat transfer is required.

Basic Phenomena

The conceptual design evolves from an analysis of the basic phenomena involved in countercurrent liquid-liquid heat exchange namely heat transfer between liquid phases and drop formation and settling. The basic phenomena consist of two elements:

1. Intimate and thorough mixing of the two fluids so as to transfer the maximum possible heat between the two contacted fluids. This is accomplished in an equilibrium stage if the temperatures of the two fluids are identical before they leave the stage.
2. Separating the two phases using the density difference where the lighter isobutane phase coalesces and flows upward and the continuous brine phase flows downward.

Since the thermal conductivity of liquids is low, the drops must be of small volume for reasonable heat transfer rates. Conversely, for good separation of the two liquid phases, large coalesced drops are required. The larger the drop size the higher the velocity of one liquid phase relative to the other. Since heat transfer requires small drops and countercurrent flow requires large drops, the two processes might be more effectively accomplished in two different regions. That is, the heat transferred in a mixing region and the phase separation achieved in a separate settling region.

DCHE Types

Liquid-liquid direct contact heat transfer is not a state of the art; that is, is not used in commercial processes. However liquid-liquid contacting is common in the chemical industry for mass transfer. Therefore the technical information available from mass transfer experience is useful in guiding the conceptual design and also in design work involving sizing equipment for liquid-liquid contacting applied to heat transfer.

There are three basic types of liquid-liquid contactors considered.

1. Mixer-Settlers
2. Gravity operated columns
3. Columns with agitation

Characteristics of these devices are summarized in Table 4-1.

Any contacting process which is based on selective distribution between phases must make allowances for intimate contacting of the phases and also separation of the phases. The mixer-settler accomplishes the contacting via a line containing a turbulence generator (valve, orifice, in-line mixer) and a vessel to provide a quiescent means for phase separation. A 5-stage mixer-settler system is shown in Figure 4-1.

Most columns with no moving parts do not have distinct regions for mixing and settling but instead provide for continuous countercurrent contacting of the phases as in a packed column. The phases never approach equilibrium; instead heat transfer occurs as a result of the displacement from equilibrium. Simple devices, such as spray towers or baffle towers, provide little more than one or two stage contacting.

Sieve plate towers do have distinct regions for dispersion and separation of the phases. They are like sieve plate distillation columns except that no outlet weir is used. As shown in Figure 4-2, one phase is dispersed by jetting through perforations into a continuum of the other phase, and the dispersed

Table 4-1

Advantages and Disadvantages of the Various Types of DCHE Contacting Devices

<u>Type of Contactor</u>	<u>Advantages</u>	<u>Disadvantages</u>
I. Mixer Settler	<ol style="list-style-type: none"> 1. Prevents backmixing. 2. Most reliable in absence of design data. 3. High throughput. 4. Easy maintenance. 5. Can supply any required power input for mixing. 	<ol style="list-style-type: none"> 1. Costly for a large number of stages.
II. Columns with Simple Gravity Flow		
A. Sieve Tray Column	<ol style="list-style-type: none"> 1. High throughput. 2. Minimizes backmixing. 3. Many theoretical stages. 	<ol style="list-style-type: none"> 1. Low power for mixing. 2. Costly maintenance in fouling service.
B. Packed Column	<ol style="list-style-type: none"> 1. High throughput. 2. Many theoretical stages. 	<ol style="list-style-type: none"> 1. Low power for mixing. 2. Not suitable for fouling service.
C. Spray Column	<ol style="list-style-type: none"> 1. Easy maintenance. 2. Simple design. 	<ol style="list-style-type: none"> 1. Substantial backmixing in commercial columns with only one or two theoretical stages. 2. Low throughput.
III. Columns with Stirrers (e.g. Rotating Disc Contactor, York-Scheibel Column)	<ol style="list-style-type: none"> 1. High power input for mixing. 2. Reliable scaleup. 3. Many theoretical stages. 	<ol style="list-style-type: none"> 1. Low throughput. 2. High cost. 3. Unsuitable if emulsifying tendencies.
IV. Centrifugal Devices	<ol style="list-style-type: none"> 1. Maximum separation for emulsifying conditions. 	<ol style="list-style-type: none"> 1. High cost. 2. Low throughput.

Figure 4-1. Five stage mixer-settler direct contact heat exchange.

Figure 4-2. Schematic drawing of a direct contact heat exchange column using sieve trays.

phase settles through the other until it coalesces within a continuum of the dispersed phase adjacent to the next plate. The continuous phase flows across the tower and down the downcomers.

The other classification of columns have mechanical or pumped agitation which is provided either by mechanical stirring or by a pulsed pumping of the liquid.

Throughput

To transfer heat from geothermal brines, a large volume of fluid must be handled, typically 10,000,000 lbs. per hour for a 50 megawatt power plant, and only a small number of stages, typically 4 to 6 theoretical, are required for transferring the heat with a sufficiently small temperature difference throughout the heat exchanger device. Since a large number of stages are not important, the most attractive type of contacting device is that which allows the highest throughput in a given size vessel. At high throughput, backmixing occurs when one phase is entrained in the other and flows back through the equipment towards its own feed point, diminishing the stagewise advantage of countercurrent flow. A mixer-settler system is designed to prevent backmixing, but in countercurrent columns, at high isobutane dispersed phase rates (for a constant continuous brine phase rate) the holdup increases and the cross sectional area available to the brine phase is decreased. Thus the velocity of the isobutane drops relative to the brine is increased and the drag on the drops becomes greater. Eventually the isobutane drops are carried in the direction of the flow of the brine. This is called the flooding point, and the isobutane will no longer enter the column but flows out with the brine. Flooding will also occur as the brine rate is increased at a constant isobutane rate. A column will flood at any flow rate of one phase if the rate of the other phase is large enough.

Because of the large capital cost of the DCHE, maximum throughput is the controlling design factor. The spray type column has a very low throughput in addition to providing only

one or two stages of contacting and so was not considered. The power mixed column also has a low throughput for the capital costs, because of the added costs of the mixing stages and because the layout provides relatively inefficient settling compared to the mixing region of the device. Classical estimating methods predict a maximum brine rate of only 2000 lb/hr-ft² for agitated columns and quotations from current vendors are not more than 2500 lb/hr-ft², compared to an estimated 6000-9000 lb/hr-ft² for packed columns and 12,000-15,000 lb/hr-ft² for sieve tray columns.

Conclusions

Sieve tray and packed columns both have high throughput as does a mixer-settler system. Consequently these were considered in the feasibility analysis in detail. The mixer-settler system described in Figure 4-1 will cost less than a sieve tray column for a 10,000,000 lb/hr brine processing unit in which 5 theoretical stages are sufficient. However, because the mixer-settler system was beyond the scope of the work statement and because ERDA requested no further work on the mixer-settler system, no work was done beyond the basic feasibility analysis, and the packed and sieve tray columns were selected for further technical evaluation, including experimental and economic feasibility.

DISCUSSION

5. Experimental Apparatus

The purpose of the experimental apparatus was to test sieve plate and packed columns using geothermal brines and isobutane at a pressure near the supercritical pressure of isobutane.

Figure 5-1 shows the process flow for the experimental apparatus set up for operation with a geothermal brine feed to the unit. A schematic of the control system for this arrangement is shown in Figure 5-2. Figure 5-3 is a photograph of the unit installed at East Mesa.

Brine from the well head flows to the brine pump P-1 where the pressure is raised to the operating pressure of the Direct Contact Heat Exchange column (DCHE) V-02, normally 650 psia controlled by PC-1. The brine flows downward through V-02 as the continuous phase, and cooled brine from V-02 is dumped or reinjected. Brine is maintained as the continuous phase by controlling the interface near the top of the column using the brine level controller LC-1.

Isobutane is stored in the Butane Accumulator V-01 and is pumped to V-02 pressure by the Butane Pump P-2. Heat added to the isobutane by P-2 is removed by the Butane Cooler HE-5. The cold, high-pressure isobutane flows upward through V-02 as the dispersed phase and is heated by the brine to at least 285°F, 10°F above critical. For the DCHE test unit, the supercritical isobutane is condensed and cooled in the Butane Condenser HE-1 and HE-2, and the cold isobutane is returned to V-01. Water vapor in equilibrium with the hot isobutane from V-02 is also condensed in HE-1 and 2 and collects in the bottom of V-01. The water is dumped by LC-2 which automatically controls the water-isobutane interface level in the bottom of V-01.

ORC DIRECT CONTACT HEAT EXCHANGE TEST UNIT PROCESS FLOW DIAGRAM

Figure 5-1. Process flow schematic of the ORC direct contact heat exchange test unit.

ORC DIRECT CONTACT HEAT EXCHANGE TEST UNIT CONTROL SYSTEM

Figure 5-2. Process control system for the ORC direct contact heat exchange test unit.

Figure 5-3. A view of the DCHE test unit installed at East Mesa.

Figure 5-4 shows the process flow for the apparatus modified to recirculate city water for instrument calibration and preliminary performance tests at La Verne. In this arrangement cooled water from V-02 is heated in the Butane Condensers, while cooling the hot overhead isobutane. Additional heat is added to the feed water by the Water Heater HE-4 to raise the water temperature to the desired value.

The isobutane in V-01 is stored at a temperature of about 180°F. Final cooling to the desired feed temperature is provided by HE-5.

Construction drawings for the apparatus are presented in Appendix F and describe the unit in detail.

Packed Column Section

An outline drawing of the DCHE V-02 with the packed column section installed is shown in Figure 5-5. Figure 5-6 is a photograph of a portion of the Goodloe packing. The column outline drawing shows the dimensions of the section and location of the instruments and sample points as installed in the field. Construction details are presented in Appendix F.

Sieve Tray Section

Figure 5-7 shows the DCHE column outline drawing for V-02 with the sieve tray section installed. The sieve tray layout is shown in Figure 5-8. Dimensions and instrument and sample point locations as installed in the field are shown on the column outline. Construction details are given in Appendix F.

Temperature and Flow Rate Measurements

Temperatures are measured using Platinum Resistance Thermometers (RTD) and a Rosemount Model 2501 Digital Temperature Readout instrument with $\pm 0.3^\circ\text{F}$ accuracy. Calibration details are discussed in Appendix D.

A Daniels honed flow section orifice meter run was used for both brine and isobutane flow rate measurement. The uncalibrated meter runs have an accuracy of $\pm 3/4\%$. The flow meter differential pressure transmitters were calibrated as described in Appendix D.

ORC DIRECT CONTACT HEAT EXCHANGE TEST UNIT

PROCESS FLOW DIAGRAM

LA VERNE, CALIF.

Figure 5-4. Process flow schematic of the ORC direct contact heat exchange test unit for operation with electrically heated water at the ORC La Verne laboratory.

Figure 5-5. Column outline for the packed column DCHE section.

Figure 5-6. A section of Goodloe packing.

Figure 5-7. Column outline for the sieve tray DCHE section.

Figure 5-8. Sieve tray layout for the 8-inch experimental DCHE column V-02.

DISCUSSION

6. Experimental Procedure

Prior to startup all utilities were checked - electricity, air and cooling water - the alarm system was checked, and isobutane level in the accumulator was checked and replenished as required. HCV-5 in the overhead line was opened halfway. About 7 gallons of isobutane was loaded to purge the air from the column and provide sufficient isobutane to fill the top of the column to keep brine from entering the overhead system. The seal water pump was started and brine was then pumped to the column. When the column was filled and at operating pressure, LC-1 was opened manually to maintain pressure and establish brine flow through the column. Isobutane was then fed to the column and the overhead pressure controllers and LC-1 were put on automatic control. The accumulator water level controller LC-2 was also set on automatic.

The flow controllers FC-1 and FC-2 were set for manual control, which gave more stable column operation and was satisfactory to maintain steady feed rates. The accumulator pressure was periodically checked and nitrogen added if necessary to maintain the pressure at 110-130 psig.

Operating data shown in Appendix H was taken every 30-60 minutes to monitor equipment operating conditions and provide a basis for adjusting the automatic controllers, pump bypass settings, seal water pump rate, etc. Overhead and bottom samples were taken frequently, usually every hour, to monitor water carryover and isobutane carryunder as a check that the column was operating properly.

Temperatures throughout the column were read periodically using the Platinum Resistance Thermometer (RTD) digital readout instrument. When steady state was reached and the temperatures

were essentially constant, a series of temperature, flow rate, and pressure readings were recorded for averaging to obtain the reported steady state temperature profile. Flow rates and pressures were maintained at constant values during this steady state period.

Calibration Runs

The RTD calibration runs were made at La Verne. Water only was circulated through the column and piping. After steady state was reached, a constant water temperature was obtained throughout the column and inlet and outlet piping. About one quarter of the water circulation rate was removed from the overhead butane line to ensure that the constant temperature water was in contact with the overhead system RTD's. Temperature readings for the column and for the inlet and outlet sensors were recorded and are summarized in Appendix D. The steady state value of the water temperature was obtained by averaging these readings and this value used to determine the temperature corrections. Details are given in Appendix D.

Scaling Test ES-1

Startup for this 76-hour test run using the packed section and East Mesa brine was as described above. The objective of this test was to run continuously and observe the amount and location of scale build up during the run. Flow rate set points were selected to provide stable supercritical operation for the packed column at about 50% of flooding. During this 76-hour run performance data was recorded each hour. Operating data was also recorded each hour to monitor equipment performance and overhead and bottoms samples were taken at intervals of 4 to 12 hours. Details of the operation for this run are given in Appendix G, and the data are summarized in Appendix H.

Scaling Test ES-2

This test ran continuously for 160 hours at East Mesa using the sieve tray section. Startup and general operating procedure is described above. As in run ES-1 the objective of the test was to observe the amount and location of scale buildup during a long continuous run. Performance and operating data were recorded hourly. Feed rates were set at about 50% of flooding to give stable operating conditions and minimize the isobutane losses in the outlet brine. For the first 18 hours of the run, the feed rates were about 70% of flooding with stable column operation but higher isobutane losses than desired. Consequently, the rates were reduced to 50% of flooding for the final 142 hours of the run. Operating details and performance data are summarized in Appendixes G & H, respectively.

Parametric Runs, EP-1 to EP-15

A series of 16 parametric runs were made to obtain performance data for the sieve tray section under varying operating conditions. The mass flow rate ratio of isobutane to brine was varied from 0.60 to 1.15. Isobutane inlet temperatures ranged from 85°F to 115°F. Column pressure was varied from 650 psia to 500 psia. Brine superficial mass velocity covered from 8300 lb/hr. ft² to 14,500 lb/hr. ft² at 650 psia at which point the column flooded by butane carryunder. Details of these runs are given in Appendices G and H.

DISCUSSION

7. Theoretical Analysis

Column Thermodynamics

The temperatures and mass flow rates of the two entering and two leaving streams from each tray of a sieve plate column Figure 7-1 are related by mass and energy balances, and the tray efficiency. Referring to Figure 7-1 the energy balance is

$$m_{W,i} h_{W,i} + m_B h_{B,i} + m_{S,i} h_{S,i} = m_{W,i-1} h_{W,i-1} + m_{S,i+1} h_{S,i+1} + m_B h_{B,i+1} \quad (7-1)$$

where the solubility of isobutane in water has been neglected and entrainment in both the overhead and underflow is assumed to be negligible. The material balance is

$$m_{W,i} - m_{W,i-1} = m_{S,i} - m_{S,i+1} \quad (7-2)$$

The tray efficiency E_i is defined by

$$E_i = (T_{W,i-1} - T_{W,i}) / [T_{W,i-1} - (T_{W,i} + T_{B,i})/2] \quad (7-3)$$

This is the ratio of the temperature drop $T_{W,i-1} - T_{W,i}$ of the brine to the difference between the entering brine temperature $T_{W,i-1}$ and the equilibrium temperature of tray i which is approximately halfway between the leaving isobutane and brine temperatures. A derivation of Equation 7-3 is given in Appendix C. At the limit of maximum heat transfer, $T_{W,i}$ will equal $T_{B,i}$ and E_i will be 1.0.

The flowrate $m_{S,1}$ of water, i.e. steam, in the isobutane outlet stream is proportional to the flowrate of isobutane according to the equation

$$m_{S,1} = [M_W p_{W,1} / M_B (P - p_{W,1})] m_B \quad (7-4)$$

where $p_{W,1}$ is assumed to be equal to the vapor pressure of water at the temperature of the outlet isobutane phase.

Figure 7-1. Schematic representation of the i and adjacent plates in a sieve plate column showing flowrate and temperatures

The enthalpies in Equation 7-1 are all related to the four temperatures by the equations of state. The equation of state for brine is obtained from the equation of state for water plus the corrections developed by Wahl (Wahl, 1977). For a given mass flow rate ratio, the above set of equations contain as variables four temperatures, the plate efficiency, and the water and steam entering and leaving flow rates. Since there are four equations and nine unknowns, five additional variables must be fixed.

Each additional stage adds four equations but only five additional unknowns, namely the tray efficiency, and the two temperatures and the water and steam flow rates at the further interface. Thus, if five variables plus one for each stage are fixed the rest are determined by the above set of equations. In addition the flowrate $m_{S,N+1}$ of water entering the bottom stage in the isobutane phase is zero and the brine entering the top stage is $m_{W,0}$. Consequently, fixing three variables plus one for each stage or a total of $N+3$ will allow solution of the above set of equations for the remaining variables.

Experimental plate efficiency. For the experimental sieve plate column tested in this program, the brine temperature $T_{W,i}$ leaving each plate was measured. In addition the mass flow rate ratio r is determined by the experimental flow rates. Also the isobutane temperature entering and leaving the column, and the brine temperature entering the column were determined experimentally. Thus in addition to one variable $T_{W,i}$ for each stage, four other variables were measured. The isobutane inlet temperature was used with an overall energy balance to check the consistency of the data as described in Appendix I.

Using the remaining three experimental variables and the temperatures $T_{W,i}$ in the above set of equations, the isobutane temperatures $T_{B,i}$ and the tray efficiency for each plate was determined as described in Appendix I and shown in Table I-4.

The results are summarized in Table 7-1, wherein the plate efficiencies of the individual plates were arithmetically averaged to indicate the general trend. Examination of this table shows that the top tray operates with an efficiency of 1.1 and the remaining trays generally operate with an efficiency of 0.7. In some cases the efficiency varies considerably from this value but this may be due to errors in calculating the isobutane temperature.

Another method of evaluating column performance is by the number $N_{100\%}$ of 100% efficient plates required to achieve the heat transfer accomplished experimentally by the 12 plate column. As shown in Table 7-1, the number of equilibrium stages range from 5 to greater than 14 with 8 to 11 being most common. The latter represent average plate efficiencies of 0.7 to 0.9.

Examination of Table 7-1 shows that plate efficiency and $N_{100\%}$ tend to increase with increasing ratio of brine to isobutane flowrates although other factors are obviously involved. The result of comparing actual and predicted performance is

1. plate efficiency typically varies from 0.4 to 1.3 although higher and lower numbers occur
2. the number of 100% stages achieved by the 12 plate column varies from 5 to greater than 14 with 8 to 11 being most representative

Thus, for predicting performance of sieve plate columns an efficiency of 1.1 at the top plate and 0.7 elsewhere can be assumed. Alternatively, the number of actual stages can be taken equal to 1.3 times the number $N_{100\%}$ of 100% efficient stages which represents an average plate efficiency of 75%.

Table 7-1

Sieve Plate Column Experimental Efficiencies

Tabulated Versus Increasing Brine Flowrate

Run No.	Pressure Psia	T _{E,9} °F	Flowrates, lbs/hr		Efficiencies for Plates				N _{100%}
			Brine	Isobutane	1	2 to 5	6 to 9	10 to 12	
16	530	226.	2197	2150	1.31	0.40	0.45	0.48	6.2
15	360	186	2241	2070	1.60	0.18	0.46	0.39	6.1
18	638	287	2384	2035	1.10	0.73	0.57	0.45	7.8
17	638	287	2593	2165	1.09	0.76	0.71	0.60	8.7
19	650	282	2909	2550	1.09	0.48	0.61	0.58	7.7
30	550	268	3068	3800	1.07	0.25	0.41	0.50	5.0
20	650	287	3347	2520	1.06	0.25	0.74	0.61	8.1
21	650	284	3390	3070	1.11	0.55	0.62	0.62	7.9
23	650	286	3643	3640	1.11	0.70	0.71	0.66	8.8
32	500	273	3719	3190	1.13	2.55	1.35	0.80	>>14
29	650	283	3827	4305	1.13	0.80	0.82	0.67	10.1
31	550	281	3946	3105	1.12	0.89	0.97	0.74	>>14
28	650	286	4050	4075	1.12	0.67	0.72	0.65	8.7
24	650	288	4204	3700	1.11	0.66	0.75	0.69	9.0
33	500	303	4289	2440	1.30	2.36	1.15	0.76	>>14
22	650	307	4656	2795	1.16	0.65	1.04	0.70	10.8
27	650	288	4693	4100	1.11	0.67	0.69	0.70	8.7
26	650	302	5023	3090	1.22	0.52	1.11	0.77	11.8
34	650	308	5082	3080	1.17	0.63	1.16	0.53	11.5
25	650	304	5155	3220	1.23	0.60	1.30	0.77	>>14

Estimating performance. For design purposes, the inlet brine temperature is the wellhead temperature and the isobutane inlet temperature is the condenser temperature. As discussed above, the plate efficiency for each stage is 75% which fixes $N+2$ variables. From these and the above set of equations, the isobutane outlet temperature can be estimated as a function of the mass flow rate ratio r and the number N of stages.

Since the mass flow rate ratio r of isobutane to brine is a basic design variable, it is convenient to express Equations 7-1, 7-2, and 7-4 in terms of r by dividing them by the brine flow rate m_W . The resulting equations on a basis of a unit mass of brine entering the column are

$$X_i h_{W,i} + r h_{B,i} + Y_i h_{S,i} = X_{i-1} h_{W,i-1} + Y_{i+1} h_{S,i+1} + r h_{B,i+1} \quad (7-5)$$

$$X_i - X_{i-1} = Y_i - Y_{i+1} \quad (7-6)$$

$$Y_i = [M_W p_{W,i} / M_B (P - p_{W,i})] r \quad (7-7)$$

Equation 7-7 applies only when isobutane is a vapor phase (e.g. above the critical temperature). Otherwise Y_i is zero for liquid isobutane.

Packed Column. Assuming that a packed column is homogeneous, then an energy balance around a differential section of the column, Figure 7-2, gives

$$m_W dh_W + (h_W + h_S) dm_W + m_B dh_B = 0 \quad (7-8)$$

Figure 7-2. Schematic Representation of a Differential Section of a Packed Column Showing Flowrates and Temperatures

or

$$Xdh_W + (h_W + h_S)dX + r dh_B = 0 \quad (7-9)$$

where use has been made of the material balance relationship $dm_S = -dm_W$ or $dY_i = -dX_i$. If this differential section is written as a finite element, then

$$X_i (h_{W,i} - h_{W,i-1}) + (h_{W,i} + h_{S,i})(X_i - X_{i-1}) + r(h_{B,i} - h_{B,i+1}) = 0 \quad (7-10)$$

Since $X_i - X_{i-1} = Y_i$, this equation is identical to Equation 7-5 for the sieve plate column. Thus all the proceeding comments on the sieve plate column also apply to the packed column except that the solution is for the limit of an infinite number of stages, $N \rightarrow \infty$. Ten to fifteen stages approach sufficiently closely to $N = \infty$.

In the case of the packed column the efficiency E will vary throughout the packing. Non homogenous behavior, such as channeling can be accounted for by the efficiency term.

Column Hydraulics

For the case of isobutane working fluid and a sieve plate column, the design procedure for maximum brine throughput consists of selecting the continuous phase, designing the sieve plate configuration, and determining the number of plates. Continuous Phase Fluid. Estimates of the optimum relative isobutane to brine flow rates are discussed above. Isobutane has the larger volumetric flow rate.

The interfacial area per unit volume of both fluids is greater when the fluid with the larger volumetric flow is dispersed into the other, and the temperatures of each phase will more rapidly approach equilibrium. Therefore tray efficiency is higher if isobutane is the dispersed phase.

Using the fluid with the smaller volumetric flow rate as the continuous phase gives maximum throughput for a given downcomer velocity corresponding to a given tendency to entrain drops of the dispersed phase.

Thus for both highest efficiency and throughput, brine is the continuous phase and isobutane the dispersed phase. Sieve Plate Configuration. Determination of the optimum sieve plate layout is accomplished by establishing the required downcomer area and the required bubbling area for the given set of brine and isobutane flow rates.

Downcomer Area. The terminal velocity of an isobutane drop in brine is the velocity at which the net gravitational forces are just balanced by the drag forces and is given by the general relation

$$u_t = \sqrt{\frac{4 g D_p (\rho_w - \rho_B)}{3 \rho_w C}} \quad (7-11)$$

The terminal velocity may be better obtained for this purpose using the Hu-Kinter correlation for liquid drops in liquid media (Hu-Kinter, 1955) using the dimensionless

groups $N_{Re}/P^{0.15}$ which contains the terminal velocity and $CN_{We} P^{0.15}$ where CN_{We} equals $4 D_p^2 (\rho_W - \rho_B) / 3\sigma$ and is the ratio of gravitational to drag forces.

The velocity u_W of the brine in the downcomer must be equal to or less than u_t to prevent isobutane entrainment and flooding. Therefore, for maximum brine throughput, $u_W = u_t$ and the required downcomer area A_D , which is the volumetric throughput $m_W/3600\rho_W$ of brine divided by the brine velocity u_t , is

$$A_D = m_W/3600\rho_W u_t \quad (7-12)$$

Experimental results from this study discussed in Section 8 show that Equations 7-11 and 7-12 with a drop diameter D_p equal to 1/32 inch predict the flooding point of the column.

Bubbling Area. For fluids with high interfacial tension as in the brine-isobutane system, small holes are needed to give good dispersion. Hole diameters of 1/8 to 1/4 inch are common in the industry (Perry, 1973) and in the absence of more detailed studies, 1/4-inch holes are recommended. The holes are arranged on an equilateral triangle or square diamond pattern with the total open hole area between 15 and 25% of the bubbling area.

To approximate minimum cost, the ratio m_W/A_T of column throughput per unit column area is maximized subject to practical constraints on the dimensions and velocities. For the purposes of this work, rather than maximizing this ratio, the isobutane velocity u_H through the holes is estimated.

The isobutane velocity through the holes should be such that the isobutane streams through and breaks up a slight distance above the plate. The minimum recommended velocity is 0.5 ft/sec (Perry, 1973). Consequently u_H is estimated as 1.5 ft/sec. This gives a reasonable bubbling area and depth of coalesced isobutane.

The relative weight ratio r of isobutane to brine is determined from thermodynamic analyses discussed in System Performance, below. The isobutane volumetric flow rate Q_B is then

$$Q_B = r m_W / 3600 \rho_B \quad (7-13)$$

and the required hole area is

$$A_H = Q_B / u_H \quad (7-14)$$

The relationship of bubbling area A_B and hole area A_H is

$$A_B = A_H / a_H \quad (7-15)$$

where a_H is between .15 and .25 as discussed above. Referring to Figure 7-3, the column cross sectional area A_T is the sum of twice the downcomer area A_D , the bubbling area A_B , and a 1% allowance for downcomer plate thickness, support rings, clamps, etc. Thus

$$A_T = 1.01 (2 A_D + A_B) \quad (7-16)$$

The above 6 equations relate the 7 variables A_D , A_H , A_B , A_T , u_t , m_W , Q_B . As a design basis either column area A_T or brine throughput m_W is set and these equations solved stepwise for the remaining 6 variables.

Ideally these equations should be solved for each tray and in a practical cylindrical column, one tray will have a limiting throughput. This will be the design throughput for the column. With experience the limiting tray can be determined from physical properties and in the case of supercritical isobutane, this is the bottom tray. An example calculation is given in Appendix K for a 12-foot diameter tray. The results are shown in Figure 7-3.

CENTER DOWNCOMER AREA = SIDE DOWNCOMER AREA = 34.1 FT²
 ACTIVE BUBBLING AREA = 43.8 FT²
 HOLE AREA = 9.6 FT² (22% OF BUBBLING AREA)
 HOLE DIAMETER = 1/4 IN. ON 1/2 IN. EQUILATERAL TRIANGLE PATTERN
 TRAY SPACING = 24 INCHES
 DOWNCOMER VOLUME = 11.4 FT³ (BASED ON 20 INCH CLEARANCE BETWEEN
 DOWNCOMER AND TRAY BELOW)

Figure 7-3. Tray layout for a 12 ft. sieve tray column.

Column Throughput. The above 6 equations can be solved to give the ratio of total column area to brine flow rate

$$\frac{A_T}{m_W} = \frac{(1.01)}{3600\rho_W} \left(\frac{r\rho_W}{u_H a_H \rho_B} + \frac{2}{u_t} \right) \quad (7-17)$$

where u_t is given by Equation 7-11. Substituting the numerical values $D_p = 0.0026$ ft, $C = 1.23$, $u_H = 1.5$ ft/sec, $A_H = 0.22$, which correspond to the example calculation and tray design in Figure 7-3, Equation 7-17 becomes

$$\frac{A_T}{m_W} = \frac{r}{1176 \rho_B} + \frac{1}{535 \sqrt{\rho_W (\rho_W - \rho_B)}} \quad (7-18)$$

This gives the ratio of column area to brine flow rate as a function of densities and the ratio of mass flow rates of isobutane to brine. Because the densities of the fluids vary from tray to tray, the limiting tray can be found using Equation 7-18.

Tray Spacing. The minimum tray spacing is the sum of the height Z_B of the coalesced isobutane layer plus the height Z_W required to accomodate flow of the brine across the tray

$$Z = Z_B + Z_W \quad (7-19)$$

Referring to Figure 4-2, page 24, the isobutane phase coalesces and collects below each plate. The pressure must be higher below the plate than above to force the isobutane through the holes. This pressure difference is the height of the coalesced isobutane layer times the density difference of the phases.

The depth of z_B of this coalesced isobutane layer is given by Treybal (Treybal, 1963)

$$z_B = \frac{u_H^2 [1 - (A_H/A_B)^2] \rho_B}{2g_C (0.67)^2 (\rho_W - \rho_B)} + \frac{6\sigma}{D_{P,0.1} g_C (\rho_W - \rho_B)} + \frac{4.5 u_W^2 \rho_W}{2g_C (\rho_W - \rho_B)} \quad (7-20)$$

in which the terms represent respectively the pressure drop due to contraction-expansion through the holes, interfacial tension, and brine blow through the downcomer. The second term is negligible at orifice velocities greater than 1 ft/sec (Treybal, 1963). Neglecting this term and simplifying, Equation 7-20 becomes

$$z_B = \frac{u_H^2 \rho_B}{30.1 (\rho_W - \rho_B)} + \frac{4.5 u_W^2 \rho_W}{2g_C (\rho_W - \rho_B)} \quad (7-21)$$

The height z_W required by the linear velocity of the brine, is obtained from the geometry of the tray. Setting the linear velocity of brine flowing across the tray equal to the downcomer velocity

$$z_W = 0.5 A_D/W_L \quad (7-22)$$

where W_L is the chord length of the side downcomer shown in Figure 7-3.

A sample tray spacing calculation is included in the example calculation given in Appendix K.

Number of Plates. The number of plates is determined by balancing the cost of the column, which increases with increasing number of plates, against the increased work output so as to minimize the cost of electricity. The increase dW in work output

is a result of the decrease in ΔT throughout the column due to the increase in the number N of plates and can be obtained from the work loss term by

$$dW = - \frac{d}{dN} \left[\int (c_p \Delta T / T_z) dT_z \right] dN \quad (7-23)$$

The cost analysis shows that the minimum cost of electricity occurs in the range of 5 to 8 stages which corresponds to 7 to 11 plates for a tray efficiency of 75% as was obtained in the experimental column.

System Performance and Costs

A simplified schematic for a process system utilizing a direct contact heat exchanger is shown in Figure 7-4. For the purposes of this study, isobutane is assumed to be the working fluid and the temperature of isobutane $T_{B,N+1}$ entering the DCHE is assumed to be equal to the condensing temperature T_{BC} . Variables to be optimized are the column pressure P , the ratio r of isobutane to brine mass flowrates, the column type j , and the number of stages. Feasibility is determined by the cost C_e^* of electricity produced for the optimum, that is minimum cost C_e^* , system.

As shown in Figure 7-5, a process system, such as that of Figure 7-4, for converting a geothermal resource into electricity consists of three elements. These are the wells, the fluid handling plant for collecting the wellhead geothermal fluid and driving a turbine, and an electrical plant for converting the turbine power to transmittable electricity.

Costs. The time reference for all costs is 2nd quarter, 1975. The cost of electricity from a geothermal powered system is the cost of maintenance of the wells and of capital plus a small amount due to operating and other maintenance costs. Referring to Figure 7-5 the annual cost is given by

$$C_A^\# = a_W C_W + a_F C_F + a_E C_E + C_B^\# \quad (7-24)$$

where the coefficient a_k is the cost of capital plus maintenance and replacement costs per unit capital cost and the subscripts W, F and E refer to wells, fluid handling plant and electricity plant respectively. Since well costs and therefore system costs vary considerably from reservoir to reservoir, a simple estimate of the component system costs are sufficient for assessing costs and feasibility in this analysis. Future work should include a detailed study of the cost of the working fluid loop and then a more detailed cost analysis will be useful.

Figure 7-4. Direct contact heat exchange process system for producing electricity.

Figure 7-5. Capital elements for a geothermal powered electricity generating process system.

To study various DCHE concepts, it is convenient to break the fluid handling costs into two parts, one the installed cost of the DCHE and the other the installed cost of all other fluid handling components of the process system. This latter term will be proportional to the fluid mass flow rate to the 0.6 power. Thus the plant cost will be

$$C_F = C_D + C_P \quad (7-25)$$

where the D and P subscripts refer to the DCHE and other process plant components respectively. Thus the annual cost of electricity is

$$C_A^\# = a_W C_W + a_F (C_D + C_P) + a_E C_E + C_B^\# \quad (7-26)$$

Electricity Produced. Taking one hour as a basis, the net electricity W_e produced is the generator efficiency times the net work W_n , that is the turbine work less the pump requirements of the system.

$$W_e = e_g (W_t - W_1 - W_2 + W_3) \quad (7-27)$$

Expressed in megawatts, the power or work per hour is given by

$$W_e^* = 0.293 \times 10^{-6} W_e \quad (7-28)$$

The turbine work is turbine efficiency times the work produced by the isentropic expansion of $m_B + m_S$ lbs. of vapor from the DCHE pressure P and temperature T_{B1} to the pressure corresponding to the condenser temperature T_{BC} . In general, part of the steam condenses to liquid water during the expansion and the turbine should be designed for this condition. The temperature T_{B1} is calculated by estimating the performance of the DCHE by solution of Equations 7-5 through 7-7 as described previously. The pump and expander works are the reversible work requirements times the appropriate efficiency. The set of equations and calculational procedures for these work terms are given in Appendix A.

Cost of Electricity. The cost C_e of the hourly electricity W_e produced is

$$C_e = C_A^{\#} / 8760 W_e \quad (7-29)$$

The cost C_e^* in mils/kw or equivalently \$/Mw is

$$C_e^* = 3.34 \times 10^{-11} C_A^{\#} / W_e \quad (7-30)$$

or

$$C_e^* = 0.293 \times 10^{-6} C_A^{\#} / W_e^* \quad (7-31)$$

Procedure. The cost C_e^* for this system along with other useful information, Figure 7-6, was calculated as described above and detailed in Appendix A by the computer program given in Appendix B for a range of the above cited variables and fixed parameters as listed in Table 7-2.

The annual cost factor for the wells is 20% for capital plus 10% for replacement and effluent disposal, for the fluid plant is 15% for capital plus 5% for operation and maintenance and for the electric plant is 12% for capital plus 3% for operation and maintenance.

Since the objective of these computations is to find the optimum design condition for the DCHE, compare different DCHE types, and assess feasibility, simple lump sum terms for the other costs are sufficient. The capital cost of the wells including the collection system is taken as $\$17 \times 10^6$ for typical 6000 ft wells producing 500,000 lb/hr of brine per well (Wahl, 1977). The capital costs of the electric plant as derived from data on geothermal steam flash electric generating plants (Bloomistan and Cohn, 1976) is

$$C_E = (0.2 + 0.4 W_e^{*0.6}) 10^6 \quad (7-32)$$

Figure 7-6. Analytic scheme for evaluation of direct contact heat exchangers in an electric generating process.

Table 7-2

Values of Fixed and Variable Parameters used to Execute
the Computer System Analysis on Basis of 10×10^6 lbs/hr brine

<u>Parameter</u> <u>Name, abbreviated</u>	<u>Symbol</u>	<u>Type</u>	<u>Range or</u> <u>Value</u>	<u>Units</u>
Brine mass flowrate	m_W	fixed	10×10^6	lbs/hr
Cost factor, Wells	a_W	"	.30	-
Cost factor, Fluid	a_F	"	.20	-
Cost factor, Electric	a_E	"	.15	-
Capital Cost, Wells	C_W	"	17×10^6	\$
Capital Cost, Fluid	C_P	"	Eq. 7-33	\$
Capital Cost, Electric	C_E	"	Eq. 7-32	\$
Generator efficiency	e_g	"	0.98	-
Cost of Butane Loss	C_B	"	0.4×10^6	\$/yr
Turbine efficiency	e_t	"	0.82	-
Brine pump efficiency	e_1	fixed	0.70	-
Butane pump efficiency	e_2	"	0.70	-
Recovery turbine efficiency	e_3	"	0.70	-
Sieve plate efficiency	E	fixed	0.75	-
Packed column HETP	HETP	fixed	0.70	ft
Column pressure	P	variable	550-700	psia
Wellhead temperature	T_{WH}	variable	400-600	°F
Condenser temperature	T_{BC}	fixed	100	°F
Number of plates	N	variable	3 to 12	-
Mass flowrates ratio	r	variable	0.4 to 2.0	-

The fluid handling component installed capital cost will vary with 0.6 power of the fluid flowrate m_B . Using estimates of capital costs for the components from Bloomistan and Cohn (1976), the fluid handling process plant cost C_P is given by

$$C_P = \{2 + [2 + 0.01(P - P_{BC})][m_B/10^6]^{0.6}\}10^6 \quad (7-33)$$

To this must be added the annual cost $C_B^\#$ of working fluid losses. The concentration of isobutane in the brine at 50 psia exhaust will be no greater than the solubility, i.e. 50 ppm. In practice, the losses should be less and might be 0.1 ppm (Sheinbaum, 1975). At \$0.10 per pound of isobutane and 50 ppm loss in the brine, this cost is $\$0.4 \times 10^6$ per year. Using these values as shown in Table 7-2, Equation 7-24 becomes

$$C_A^\# = 10^6 \{6 + 0.4W_e^{*0.6} + (m_B/10^6)^{0.6}[2 + 0.01(P - P_{BC})]\} + 0.2C_D \quad (7-34)$$

DISCUSSION

8. Experimental Results

Scaling

Experimental runs were made using the packed section and sieve tray section to evaluate scaling in the DCHE column. Run ES-1 (Packed Column Scaling Test). The small deposit formed on the inlet section of column wall, on the bottom distributor, and in the outlet valve was very low in calcium carbonate and silicates but high in iron in the oxide form (Table H-12, Appendix H). This deposition may be because of leakage of seal water through the mechanical seal of the brine pump, because of startup and shutdown oxygen contamination in the system, or because of other startup and shutdown factors. In following tests oxygen contamination was minimized during start-up and shutdown to establish whether or not this deposition was due to that source of oxygen. No scale formed in subsequent runs. Consequently, oxygen contamination during startup and shutdown is assumed to have caused the scale found in Run ES-1

Very little scale was formed during this 76-hour run and was completely absent in the column itself. Of the scale that was present, only a negligible amount which had formed on the top distributor plate was calcium carbonate and silicate scale.

Run ES-2 (Sieve Tray Column Scaling Test). After 160 hours of continuous operation with the sieve tray section, inspection of the unit showed the only scale was a small amount of deposit on the carbon steel column wall extension in the vapor space inlet region above the exchanger column. The scale that did form occurred in small gray-black spots that wiped off easily. Throughout the unit, there was not enough scale to get a sample for analysis. The butane overhead and column bonnet section appeared the same as at the start of the run. Rust present in the sieve tray section at the start of the run was removed

during the run and the sieve trays and column wall were substantially cleaner at the end of the 160-hour run than before the run.

These runs are described in greater detail in Appendix G and visual inspection and sample data are discussed in Appendix H.

Throughput

Packed Column Section. There were no flooding tests on the packed column section at supercritical conditions. However several runs were made at La Verne at lower temperatures which resulted in flooding. At a water rate of 4800 lb/hr ft², the column flooded at an isobutane rate of 5500 lb/hr ft². For an inlet water temperature of 240°F, the column flooded by carrying water overhead with the isobutane, while an inlet water temperature of 185°F caused the column to flood by isobutane carryunder with the outlet water. Later supercritical runs at East Mesa using 340°F brine gave stable column operation with a brine rate of 7500 lb/hr ft² and an isobutane rate of 4400 lb/hr ft².

Sieve Tray Section. Column operation was stable at a brine rate of over 12,000 lb/hr-ft² for six of the parametric runs (EP-3,5,6,7,8, and 14). Isobutane rates for these runs varied from 7,000 lb/hr-ft² at 500 psia to 12,500 lb/hr-ft² at 650 psia. Runs EP-6 and 7 showed stable operation at a brine rate of 14,000 lb/hr ft² with an isobutane rate of 9000 lb/hr ft². During run EP-15 the sieve tray section flooded by carrying isobutane out the bottom of the column after 22 minutes of operation with a brine rate of 14,500 lb/hr-ft² and an isobutane rate of 9000 lb/hr-ft² at 650 psia.

Plate Efficiency

The plate efficiencies were determined from the experimental brine temperatures and calculated isobutane temperatures as described

in Section 7. Theoretical Analysis - Column Thermodynamics - Experimental plate efficiency. The result is

1. Plate efficiency typically varies from 0.4 to 1.3 although higher and lower numbers occur.
2. The number of 100% stages achieved by the 12 plate column varies from 5 to greater than 14 with 8 to 11 being most representative.

Thus, for predicting performance of sieve plate columns an efficiency of 1.1 at the top plate and 0.7 elsewhere can be assumed. Alternatively, the number of actual stages can be taken equal to 1.3 times the number $N_{100\%}$ of 100% efficient stages which represents an average tray efficiency of 75%.

Packed Column Efficiency. The effectiveness of the packed column section was determined at the measured flowrates by computing the number of theoretical plates N (i.e. 100% stages) corresponding to the experimental brine and isobutane temperatures. From this value of N and the known packing dimensions, the height of packing equivalent to a theoretical plate HETP is calculated. These results are summarized in Table 8-1. The HETP varies from 1.3 to 2.7 feet with 1.7 to 2.3 feet being most representative. The average HETP value of 2 feet satisfactorily represents the performance of the Goodloe packing for this service.

Table 8-1
Experimental Packed Column Performance

Run No.	Pressure Psia	T _{E,9} °F	<u>Flowrates, lbs/hr</u>		Flowrate Ratio, r	HETP ft
			<u>Calculated Brine</u>	<u>Experimental Isobutane</u>		
1	595	275	1255	1180	0.940	2.3
2	605	166	968	1910	1.97	2.2
3	615	275	1025	1010	0.985	2.7
4	515	219	1502	1910	1.27	1.5
5	615	271	891	1050	1.18	1.8
6	615	283	1469	680	0.463	2.3
7	632	301	2195	970	0.442	1.3
8	600	302	2358	970	0.411	1.3
9	548	292	2570	1540	0.599	2.4
10	557	293	2903	1520	0.524	1.5
11	565	293	1895	965	0.509	1.8
12	565	287	1934	1075	0.556	1.8
13	565	291	2018	1075	0.533	1.7
14	565	295	1950	1065	0.546	1.7

DISCUSSION

9. Theoretical Results

Literature Survey

Experimental tests of direct contact heat exchange columns have used spray-type columns without packing or internal components in which the liquid droplets were formed by a distributor at the top and the bottom of the column. The liquid drops thus formed in the continuous phase flowed counter-current to the continuous phase. These experimental tests showed that all of the heat transfer takes place in a relatively well-mixed zone at either end of the column, so that these columns operate essentially as one or two well-mixed tanks. That is, they are a single, or at most a two-stage counter-current device.

Conceptual Design

The implication of the literature survey is that the heat transfer rate in the center section of the column is too low relative to the rate at either end of the column where the drops are formed. This means that some sort of internal arrangement must be used to increase the rate of heat transfer in the column. This can be accomplished by sieve trays, by packing, or by using mechanical rotating devices to increase the rate of heat transfer in analogy with common practice for mass transfer columns. Another approach is to run a series of mixer settlers as a system to provide a countercurrent direct contact heat exchange device. As discussed in this report, this latter approach is in certain instances the most economical because of the capital cost of the equipment for processing a sufficiently large amount of brine.

The results of the conceptual design phase are integral with Section 4, Conceptual Design.

Design Procedure

There are 7 principal variables in the sieve tray layout design. These are downcomer area A_D , active hole area A_H , bubbling area A_B , column cross sectional area A_T , terminal velocity of an isobutane drop in brine u_t , brine mass flow rate m_w , and isobutane volumetric rate Q_B .

The 6 equations relating these 7 variables are

$$u_t = \sqrt{\frac{4g D_p (\rho_W - \rho_B)}{3\rho_W C}} \quad (9-1)$$

$$A_D = m_W / 3600 \rho_W u_t \quad (9-2)$$

$$Q_B = r m_W / 3600 \rho_B \quad (9-3)$$

$$A_H = Q_B / u_H \quad (9-4)$$

$$A_B = A_H / a_H \quad (9-5)$$

$$A_T = 1.01 (2A_D + A_B) \quad (9-6)$$

These equations can be combined to give the ratio of column area to brine flow rate assuming the conditions discussed in Section 7.

$$\frac{A_T}{m_W} = \frac{r}{1176 \rho_B} + \frac{1}{535 \sqrt{\rho_W (\rho_W - \rho_B)}} \quad (9-7)$$

For the layout design, either column area A_T or brine throughput m_W is set in Equation 9-7 and Equations 9-1 through 9-6 solved stepwise for the remaining 6 variables.

Minimum tray spacing is given by

$$Z = Z_B + Z_W \quad (9-8)$$

where

$$Z_B = \frac{u_H^2 \rho_B}{30.1 (\rho_W - \rho_B)} + \frac{4.5 u_W^2 \rho_W}{2gc (\rho_W - \rho_B)} \quad (9-9)$$

and

$$Z_W = 0.5 A_D / W_L \quad (9-10)$$

These relationships are described in detail in Section 7 and an example calculation for tray layout design and tray spacing is given in Appendix K.

Feasibility Analysis

Starting with a given brine flow rate and temperature and a given condenser temperature, column thermodynamic and hydraulic calculations were made for a range of isobutane flow rates and number of stages. The cost of electricity was computed for each of the three DCHE types selected in the conceptual design phase, namely a packed column, a sieve tray column and a mixer-settler cascade. However since the cost of the DCHE is only 5-10% of the cost of the system, the DCHE type does not significantly effect the cost of electricity, the variation in general being less than 0.3 mils/KW.

Number of Stages. The optimum number of stages to minimize the cost of electricity is 6 to 8 as shown in Figure 9-1. As discussed previously, more stages result in a higher capital cost for the DCHE, but do not significantly increase the work output from the system. That is, the efficiency of recovering heat from a geothermal brine using a supercritical working fluid is not significantly increased by using additional stages beyond 7 or 8.

Flow Rate Ratio. As shown in Figures 9-2 and 9-3, the optimum mass flow rate ratio of isobutane to brine for the supercritical isobutane system is 0.7 to 1.0, depending on the inlet brine temperature and operating pressure of the column. The cost of electricity can be considered to be a minimum in general for flow rate ratios between 0.8 and 1.0.

Figure 9-1. Cost of electricity vs. number of stages for equal brine and isobutane mass flow rates for a well head temperature of 400°F and a DCHE operating pressure of 650 psia.

Figure 9-2. Cost of electricity vs. the isobutane to brine mass flow rate ratio r for seven contracting stages and a DCHE operating pressure of 650 psia.

Figure 9-3. Cost of electricity vs. the isobutane to brine mass flow rate ratio r for seven contracting stages and a well head temperature of 400°F.

Operating Pressure. Figure 9-3 shows the effect of column operating pressure on system performance. The optimum flow rate ratio and operating pressure are strongly related as shown in Figure 9-3, since the system ΔT is a function of both the flow rate ratio and the shape of the isobutane heat content curve which is strongly pressure dependent. In general, lower pressures (above critical) have the lowest cost of electricity.

Unsolved Engineering Problems

There are no unsolved engineering problems that are critical to the design of a commercial scale DCHE plant. However, in the event of commercialization of the process, further knowledge as discussed below for refinement of the process design should be completed since for the large scale equipment involved, small improvements save significant capital investment.

Hole Diameter. The effect of hole diameter on column capacity and efficiency should be studied to find an optimum value. The holes in the 8-in. diameter experimental column were 1/8-in. diameter. Larger holes may increase column throughput for a given degree of downcomer backmixing by giving a larger average isobutane drop size, and consequently less tendency to be swept back down the column with the brine. In commercial scale liquid-liquid extraction columns, the holes are typically 1/8- to 1/4- in. diameter.

The thermal conductivity of water is 0.4 and for other liquids typically 0.1 Btu/hr. ft²°F. Mass diffusion coefficients on the other hand, are 1 to 6 x 10⁵ ft²/hr. For typical heat transfer processes and mass transfer processes, this means that the transfer rate for a thermal process is about 20 times that of a mass diffusion process. Consequently, even though larger

diameter holes give less effective dispersion of the isobutane drops in the brine, the heat transfer efficiency may not be greatly affected because of the high transfer rate for a thermal process and column throughput may be increased because of the lower tendency to entrain the larger diameter isobutane drops with the brine.

Minimum Bubbling Area. The effect of high isobutane hole velocity on tray efficiency and tray spacing requires study. A maximum velocity u_H of isobutane through the holes gives the minimum bubbling area A_B by Equations 7-14 and 7-15. For a given column cross sectional area, using the minimum bubbling area results in the maximum downcomer area and therefore the maximum brine throughput since brine velocity in the downcomer is the limiting design criteria. However, if a smaller isobutane drop diameter D_p resulted from higher velocity, increased backmixing may counteract the advantage of more downcomer area.

Isobutane Recovery from the Cooled Brine. Because of the large volumes handled in a commercial plant, a small percentage loss of working fluid is costly. Consequently design of the system to minimize isobutane losses should be thoroughly studied.

Standard chemical and petroleum industry practice for good separation of two liquid phases (Clarke and Davidson, 1962) is that the terminal velocity u_t of an isobutane drop, under Stoke's law conditions corresponds to a drop size D_p of 0.005 inches or 127 microns. This is the criteria used in this project to define the size of the column bottoms surge section and the final cold stage of the mixer-settler cascade.

During the experimental run ES-2, where isobutane losses were accurately determined, the Stoke's law terminal velocity in the bottom surge section of the experimental column corresponded to a calculated isobutane drop size of 120 microns. As

discussed in Appendix H, the experimental losses were 0.00125 mole fraction isobutane in the brine, whereas the best available solubility data predicts 0.00023 mole fraction. Because of the time required to reach equilibrium, the amount of isobutane dissolved in the brine should be less than the equilibrium solubility (Boehm, 1977). However for the criteria $D_p = 0.005$ inches, experimental results show isobutane losses are about 5 times the equilibrium solubility value. Consequently, standard chemical industry criteria does not appear to provide a satisfactory separation for the DCHE system application and further work is needed in the brine-isobutane separation system.

Sieve Plate Tray Efficiency. To accurately determine the tray efficiencies, it is necessary to know the temperature of the isobutane leaving the tray as well as the brine temperature from each tray. Instrumentation can be designed to measure the temperatures of both phases, but was beyond the scope of the current project. Further, the effect of scale-up on tray efficiency is not known. Consequently future studies should include using a column with a minimum diameter of 3 ft. and a tray layout as described in Section 7.

Mixer-Settler System. The mixer-settler system should be studied because it may be the most economic and reliable system for direct contact heat exchange of geothermal brines. The main question to be resolved with respect to the mixer-settler is the size of the settling device. The theoretical considerations to estimate this performance are rather involved and require an analysis of the statistical distribution of eddies within the system. This can be done but is more time consuming and less reliable than a direct experimental measurement of a single mixer-settler stage. This can be accomplished with simple modification to the ORC DCHE unit.

Follow-on R&D. The most important follow-on phase is the demonstration of a complete geothermal powered secondary fluid system including the turbine. The first step would be selection of a suitable reservoir, probably in the 400-500°F temperature range. The reactions of complex brine chemistry are impossible to predict with certainty, so the DCHE test unit should be operated using the reservoir fluid at the site selected to ensure that there are no scaling problems and that the system handles the non-condensibles with no venting facilities required.

The next step would be the design of the system to minimize the cost of electricity, including selection of the working fluid, optimization of the DCHE operating pressure, design of the DCHE and selection of the optimum stripper pressure in the working fluid recovery system. An overall economic evaluation should be included to confirm the value of a secondary working fluid system.

Another important follow-on phase is the development of a general procedure for the total system design described above. Other recommended work would result in improvement of the system after its successful demonstration in the above phases as shown in Table 9-1.

Table 9-1
Recommended Follow-on R&D

	<u>Estimated Time</u>	<u>Estimated Cost, \$</u>
I. Demonstrate a 5 to 50 MW Prototype Electricity Generating Plant, Excluding Reservoir Development.		
Phase I - Selection of the geothermal reservoir site and - Design of the complete plant	2 years	750,000
Phase II - Construction and testing of the secondary working fluid system (excluding turbine)	2 years	5,000,000
Phase III - Addition of the turbine and related facilities and testing of the complete electricity generating plant	2 years	40,000,000
II. Develop Total System Design Procedures Including (1) Selection of Working Fluid, (2) DCHE Operating Pressure, and (3) Working Fluid Recovery System Optimization	6 months	95,000
III. Perform Experimental Mixer-Settler System Studies Using a Modification of the Present ORC DCHE Test Unit.	1 year	120,000
IV. Test the Scaling Characteristics of Two Other Geothermal Reservoirs Using the ORC DCHE Test Unit.	1 year	90,000

DISCUSSION

10. Discussion of Results

Assessment of the technical feasibility of direct contact heat exchange based on the pilot plant tests together with theoretical interpretation shows that there are no unusual difficulties in operating such a system. Assessment of the feasibility of the entire system based on economics and conventional chemical processing technology shows that the system is essentially state-of-the-art and that the capital cost of the direct contact heat exchange portions of the system are a small portion of the total capital cost. This is to be contrasted with secondary working fluid systems that use conventional heat exchangers and for which the exchanger costs are about 50% of the cost of the process plant. A commercial size column would probably be 10 or 12 feet in diameter because this is the largest transportable vessel diameter and because the optimum commercial size vessel excluding transport considerations would be larger than this.

The estimated plate efficiency is typically 75% but varies from 30% to greater than 100%. Since the number of plates in a column does not substantially affect cost, feasibility is sufficiently well determined by an approximation of 75% for the design of a first commercial size unit. Subsequent refinement on a commercial scale would be implemented on following designs.

The conceptual design and feasibility analysis showed that the capital cost of conventional direct contact heat exchange units are a significant portion of the capital cost of the plant and consequently the highest throughput per unit capital cost should be obtained. As a result of optimization based on the conceptual design and feasibility analysis so as to achieve a maximum throughput, the capital cost of the sieve

tray column became a relatively small portion of the total capital cost of the plant so that further improvement in the capital cost of the direct contact heat exchangers have a relatively small effect on the ultimate cost of the electricity produced. Thus, the cost of electricity obtained from the feasibility analysis phase will not be significantly altered by further refinements in the DCHE design.

Since the design procedures that were developed for the sieve tray column adequately predict the thermodynamic and hydraulic performance of the experimental column, no further pilot scale data is required for the design of a commercial size unit. It should be noted, however, that increasing the diameter of a column from the 8-inch diameter pilot scale column to a 12-foot diameter commercial column will affect the efficiency and maximum throughput.

Thus, the next step in the research and development of direct contact heat exchange for generating electricity would be the construction and operation of a 10 or 12-foot diameter column in a complete secondary working fluid system. For this design the efficiency of the plates could be estimated but the error in prediction would not alter the performance of the system by more than one or two plates which amounts to a change in temperature of the overhead isobutane of about 5°F and would not alter the performance of the system significantly for evaluation. This size column corresponds to an electrical output of 6 to 7 megawatts depending on the temperature of the geothermal reservoir being used.

Since operation of the pilot scale column and test unit at East Mesa showed that no scale deposition occurred in the column or anywhere in the test unit, it would be expected that any geothermal reservoir containing calcium carbonate and/or silica concentrations similar to that of the East Mesa Well 6-2 brines would be satisfactorily handled by this type of system. Since this includes a large percentage of geothermal reservoirs,

the direct contact heat exchange system should have a wide applicability. How the system would behave on a very high total dissolved solids fluid such as the Salton Sea Brine is unknown and can best be determined by an experimental study. Thus, operation of the test unit using other geothermal reservoirs would demonstrate the breadth of applicability of this type of direct contact heat exchange column and system.

No gas release was necessary from the accumulator tank which followed the condenser in the test unit, thus no requirement for removal of noncondensibles is anticipated for the condenser in an electric generating system. This means that the net power produced from the system would be as much as 5% higher compared with systems that require removal of noncondensibles such as a flash steam plant. This is a function of the pressure at which the condenser is operated and this particular characteristic should be optimized in a design study and the results verified in a complete operating system of sufficient size for proper evaluation, such as a 5 megawatt or larger system. In addition, no environmental problems would exist for this system because of discharge of evolved gas to the atmosphere.

Since only 6 theoretical stages or less are needed for efficient heat recovery, a mixer-settler system competes effectively with a sieve tray column. The feasibility analysis shows that the capital cost of a mixer-settler DCHE system is less than that of a sieve tray column system in which 5 theoretical stages are used. Consequently a mixer-settler system should be further investigated to determine its value compared with that of a sieve tray column.

NOMENCLATURE

a_k	Annual cost factor; a_W for wells, a_F for fluid process plant; a_E for electric plant
A	Area, ft^2
c	Heat capacity, BTU/lb °F; corrosion allowance, inches
C	Drag coefficient, dimensionless; cost, \$
C_e	Cost of electricity, \$/BTU
C_e^*	Cost of electricity, \$/MW or mils/kw
$C^\#$	Annual cost, \$/year
D	Diameter, ft
e	Machine efficiency
E	Efficiency
$f()$	Function
$f_B()$	Equation of state for isobutane
$f_W()$	Equation of state for water
g	Local acceleration due to gravity, 32.17 ft/sec^2
g_c	Dimensional constant, $32.17 (\text{ft})(\text{lb})/(\text{lb-force})(\text{sec}^2)$
G	Mass velocity, lb/hr ft^2
h	Specific enthalpy, BTU/lb; differential pressure, in. of water
H	Vessel height or length, ft
J	TI number
K	Conversion factor for units in work equations
L	Liquid mass velocity, lb/hr-ft^2
m	Mass flow rate, lbs/hr
M	Molecular weight
n	Polytropic expansion coefficient
N	Number of plates, dimensionless number
p	Partial pressure, psia

P	Column pressure, psia; P_{BC} condenser pressure; P_{WH} wellhead pressure; dimensionless group, $g_c^3 \rho_w^2 \sigma_i^3 / g \mu_w^4 (\rho_w - \rho_B)$
Q	Heat flow, BTU/hr; volumetric flow rate, ft^3/sec
r	Ratio of mass flow rates of isobutane to brine
R	Volume fraction isobutane in column
s	Specific entropy, BTU/lb-°R
S	Allowable working stress, psi
t	Vessel wall thickness, inches; time, sec.
T	Temperature, °F
T*	Temperature, °R
T'	Temperature, °K
TI	Abbreviation for temperature indicator
u	Velocity, ft/sec; u_t terminal velocity of droplet, ft/sec
U	Overall heat transfer coefficient, $\text{BTU/hr-ft}^2\text{-°F}$
V	Volume, ft^3
W	Side downcomer chord length, ft; weight, lbs; work, BTU/hr
W*	Work, megawatts
w_t	Weight percent of total dissolved solids
x	Liquid phase mole fraction; equivalent thickness, inches
X	Mass fraction of entering brine as liquid
y	Vapor phase mole fraction
Y	Mass fraction of entering brine as steam
z	Compressibility factor = PV/RT
Z	Height, ft.

Subscripts

A	Annual
B	Working fluid, e.g. isobutane; bubbling
BC	Isobutane condenser
C	Critical
CAL	Calibration factor
D	Dissolved and entrained i-C ₄ in outlet brine; downcomer; DCHE
e	Net electrical
E	Experimental corrected, electrical
F	Fluid process plant
g	Generator
H	Hole
i	Stage number, component
I	Indicated, i.e. experimental uncorrected; internals
j	Type of DCHE
J	TI number
k	Type of plant
K	Column internals
L	Losses; liquid; length; packed column segment
M	Mixer, mixer-settler
n	Net
o	Orifice
p	Dispersed phase particle; packing; polytropic
P	Plant
R	Reversible
Re	Reynolds
s	Sieve tray

S	Water (steam) in isobutane phase
t	Terminal; turbine
T	Total; tray
th	Thermodynamic
V	Vessel, vapor
W	Water or brine; wells; work
We	Weber
WH	Well head

Greek Symbols

ρ	Density, lb/ft ³
ρ'	Density, gm/cc
σ	Interfacial tension, lb-force/ft
τ	Temperature function = 647.11 - T'
μ	Viscosity, lb/ft. sec.
μ'	Viscosity, centipoises

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Appendix A

Description of Computer Program for System Performance Analyses

The program consists of three parts, column temperature calculations, cost analysis, and power output and related efficiencies, as shown in Figure A-1. This program and the system equations described herein are on the basis of m_W pounds per hour of brine. All computations were for m_W equal to 10 million pounds per hour.

Column Thermodynamics

This section of the computer program is on the basis of one pound of brine so that the results are expressed as a function of r . Determination of the temperature at each stage of the column requires the solution of the four nonlinear equations 7-3, 7-5, 7-6, and 7-7. These are solved by an iteration technique. Then, a trial and error technique is used to find the isobutane outlet temperature for N stages that results in the calculated condenser temperature matching the given condenser temperature T_{BC} . The computer flow diagram for this sequence is given in Figure A-2.

Stage Calculations. The mass balance Equation 7-6 on the i th stage, Figure 7-1, page 41 is solved to give the water flow rate X_i leaving the stage

$$X_i = Y_i - Y_{i+1} + X_{i-1} \quad (A-1)$$

The energy balance Equation 7-5 on the i th stage is solved to give the enthalpy of isobutane entering the stage

$$h_{B,i+1} = (X_i h_{W,i} + r h_{B,i} + Y_i h_{S,i} - X_{i-1} h_{W,i} - Y_{i+1} h_{S,i+1})/r \quad (A-2)$$

Figure A-1. Flow Schematic for Sequence of Computer Computations.

Figure A-2. Subroutine Finish

The temperature $T_{B,i+1}$ and enthalpy $h_{B,i+1}$ of isobutane entering the i th stage are related by the equation of state

$$T_{B,i+1} = f_B(h_{B,i+1}, P) \quad (A-3)$$

The water vapor pressure p_W is calculated using

$$P_W = f_W(T_{B,i+1}) \quad (A-4)$$

assuming that the water vapor is in equilibrium with water liquid at the temperature of the leaving isobutane. The rate Y_{i+1} of steam entering the i th stage with the isobutane is given by Equation 7-7 reproduced here as

$$Y_{i+1} = M_W p_{W,i+1} / M_B (P - p_{W,i+1}) \quad (A-5)$$

The isobutane and water temperatures leaving a given stage are related by Equation 7-3 for the tray efficiency which is solved to give

$$T_{W,i} = [T_{W,i-1} - E(T_{W,i-1} - T_{B,i}/2)] / (1 - E/2) \quad (A-6)$$

The enthalpy $h_{W,i}$ of water leaving the i th stage is then given by the equation of state for brine water.

$$h_{W,i} = f_W(T_{W,i}, P) \quad (A-7)$$

Typical values of the equation of state used in this program are give in Tables A-1 and A-2.

The entering isobutane enthalpy $h_{B,i+1}$, entering steam rate Y_{i+1} , and leaving water enthalpy $h_{W,i}$ of the first, $i=1$, and subsequent stages are then calculated by iteration of the above equations.

Table A-1

Properties of Brine and Water Calculated from the Equation of
State Used in the Computer Programs

Solids	Temperature	Enthalpy	Density
<u>Wgt. %</u>	<u>°F</u>	<u>Btu/lb</u>	<u>gms/cc</u>
0.0	50.0	19.2	.987
	100.0	69.0	.983
	150.0	118.7	.972
	200.0	168.7	.956
	250.0	219.0	.937
	300.0	271.3	.913
	350.0	319.8	.887
	400.0	373.0	.856
	450.0	430.8	.822
	500.0	490.0	.782
	550.0	549.0	.735
10.0	600.0	612.2	.677
	50.0	18.3	1.060
	100.0	64.1	1.056
	150.0	109.9	1.045
	200.0	155.8	1.029
	250.0	202.1	1.010
	300.0	250.2	.986
	350.0	294.8	.960
	400.0	343.7	.929
	450.0	396.7	.895
	500.0	451.0	.855
	550.0	505.1	.808
	600.0	562.9	.750

Table A-2

Thermodynamic Properties of Isobutane Calculated from the
Equation of State Used in the Computer Program

Pressure	Temperature	Enthalpy	Density
<u>Psia</u>	<u>°F</u>	<u>Btu/lb</u>	<u>gm/cc</u>
600.00	0.00	103.42	.6166
	50.00	130.12	.5848
	100.00	158.98	.5477
	150.00	190.45	.5032
	200.00	225.45	.4472
	250.00	266.85	.3670
	300.00	344.73	.1550
	350.00	389.33	.1038
	400.00	424.92	.0863
	450.00	459.42	.0757
	500.00	494.05	.0682
	550.00	529.25	.0625
	600.00	565.18	.0580
700.00	0.00	103.71	.6178
	50.00	130.35	.5865
	100.00	159.09	.5502
	150.00	190.34	.5071
	200.00	224.86	.4538
	250.00	264.63	.3818
	300.00	319.86	.2519
	350.00	380.03	.1353
	400.00	418.84	.1067
	450.00	454.76	.0917
	500.00	490.24	.0817
	550.00	526.02	.0743
	600.00	562.39	.0686

Step 1. The isobutane temperature $T_{B,1}$ leaving the top stage of the column is estimated by the equation

$$T_{B,1} = [T_{WH} - 50(T_{WH}/325)^3/E + 0.8(N-5)]/r^{0.2} \quad (A-8)$$

This estimate is based on a 50°F temperature difference between $T_{B,1}$ and T_{WH} for $T_{WH} = 325^\circ\text{F}$ at 100% stage efficiency of a 5 plate column operated at $r = 1.0$. The second term corrects for the variation of T_{WH} and stage efficiency, the third term for the number of stages and the denominator for the variation in flow rate ratio. The steam rate Y_1 , leaving the column is determined by solution of Equations A-4 and A-5 for $(i+1)=1$.

Step 2. The stage calculations involve the iterative solution of Equations A-1 through A-7. The unknowns are X_i , $T_{B,i+1}$, $h_{B,i+1}$, $P_{W,i+1}$, Y_{i+1} , $T_{W,i}$, and $h_{W,i}$.

When the isobutane is in a vapor phase, steam is present and for this case the heat balance is computed by trial and error. The entering isobutane enthalpy $h_{B,i+1}$ is estimated by assuming the isobutane temperature using the equation

$$T_{B,i+1} = T_{B,i} - 0.8 (T_{W,i-1} - T_{B,i}) \quad (A-9)$$

The steam rate is then given by Equation A-5, the water rate by the material balance Equation A-1, and the isobutane enthalpy $h_{B,i+1}$ is calculated by the heat balance Equation A-2. The estimated temperature $T_{B,i+1}$ is incremented by one half the delta given by Equation A-9 and the process is repeated until the isobutane enthalpy converges to within 0.1 BTU/lb. The stage information is then stored.

When isobutane is in a liquid phase, or if the temperature is below the critical temperature when operating at supercritical pressures, no steam is present. That is, Y_{i+1} is zero. For this case the entering isobutane enthalpy $h_{B,i+1}$ is calculated directly by assuming there is no vaporization in the system so that Equation A-2 becomes

$$h_{B,i+1} = (X_{i-1}/r)(h_{W,i} - h_{W,i-1}) + h_{B,i} \quad (A-10)$$

The corresponding temperature of the entering hydrocarbon stream is calculated by Equation A-3. Equation A-3 is solved by an iterative solution of the equation of state in the form

$$h(T,P) = \text{const.} \quad (A-11)$$

The stage information is then stored.

Step 3. The stage calculations, Step 2, are repeated for N stages to determine the isobutane inlet temperature $T_{B,N+1}$ corresponding to the assumed isobutane outlet temperature $T_{B,1}$.

Step 4. A new isobutane temperature $T_{B,1}$ is then estimated. If the calculated isobutane inlet temperature $T_{B,N+1}$ differs from the condenser temperature by more than 100°F, the new value assigned to $T_{B,1}$ is determined from the previous assumed value of $T_{B,1}$ by the equation

$$T_{B,1}(\text{new}) = T_{B,1}(\text{assumed}) + 0.5(T_{WH} - T_{B,1}) \quad (A-12)$$

If the difference between the condenser temperature and calculated inlet temperature $T_{B,N+1}$ is within 100°F, previous history is used to determine the change in outlet temperature with respect to the change in calculated inlet temperature $\partial T_{B,1}/\partial T_{B,N+1}$ and a new value of $T_{B,1}$ is estimated using the equation

$$T_{B,1}(\text{new}) = T_{B,1}(\text{assumed}) + \partial T_{B,1}/\partial T_{B,N+1} (T_{B,N+1} - T_C) \quad (\text{A-13})$$

Then Steps 1 through 4 are repeated until calculated isobutane temperature $T_{B,N+1}$ entering the bottom of the column matches the desired temperature, the condenser temperature T_{BC} .

Despite the several interrelated iteration procedures, the numerical stability of the calculations is quite good. The guess of the outlet temperature of the isobutane is the most important starting parameter.

Power Output

The net power output is the turbine power minus the pump power requirements.

Turbine Work. The turbine model used is an isentropic expansion to give the reversible work W_R for the expansion of isobutane and steam from the DCHE overhead conditions to the condenser pressure P_{BC} .

The entropy of the isobutane-steam mixture at the turbine inlet is calculated by

$$s_{IN} = \sum (Y_i s_{V,i} - y_i \ln y_i) \quad (\text{A-14})$$

where $s_{V,i}$ is the component real gas entropy evaluated at the temperature of the stream and the component partial pressure.

The turbine outlet fluid temperature T_t at P_{BC} is calculated by trial and error to give $s_{OUT} = s_{IN}$ where

$$s_{OUT} = \sum (Y_i s_{v,i} - Y_i \ln Y_i) (m_B + m_{S,t}) / (m_B + m_{S,l}) + s_{L,W} (m_{W,t}) / (m_B + m_{S,l}) \quad (A-15)$$

At the turbine discharge conditions, in general some of the steam has condensed resulting in $m_{W,t}$ pounds of liquid water plus $m_{S,t}$ pounds of steam remaining in the vapor phase.

The reversible work is then given by

$$W_R = (h_{IN} - h_t) (m_B + m_{S,l}) \quad (A-16)$$

where the turbine inlet enthalpy h_{IN} is evaluated at the DCHE overhead conditions, and the turbine discharge enthalpy is evaluated for the equilibrium fluid at the condenser pressure P_{BC} and the isentropic discharge temperature T_t .

The total actual turbine work W_t produced is the reversible work multiplied by the polytropic machine efficiency e_p . Thus

$$W_t = e_p W_R \quad (A-17)$$

Pump Work. The pump work W_1 requirement for the brine is

$$W_1 = K m_W (P - P_{WH}) / W e_1 \quad (A-18)$$

and for the isobutane is

$$W_2 = K m_B (P - P_{BC}) / B e_2 \quad (A-19)$$

where P_{WH} and P_{BC} are the wellhead and condenser pressures respectively.

Some work is also recovered by expansion of the cool brine before reinjection. The model used is

$$W_3 = K m_{W,N} (P - P_{IN}) e_3 / \rho_W \quad (A-20)$$

where $m_{W,N}$ is the water mass flow out of the last stage and P_{IN} is the reinjection pressure.

In the mixer settler system there are also the pump losses of the pump for each stage. These are accounted for by subtracting W_M per stage which is estimated to be 72 KW for 10×10^6 lbs/hour of brine when making the net power calculations.

Net Work. The net work W_n produced is

$$W_n = W_t - W_1 - W_2 + W_3 - NW_M \quad (A-21)$$

The thermal efficiency E_{th} is obtained by dividing the net work output by the heat available from the brine. Thus

$$E_{th} = W_n / m_W (h_{W,WH} - h_{W,N}) \quad (A-22)$$

A more meaningful efficiency is the ratio of the net work output to the available work above the condenser temperature. This ratio E_W is given by

$$E_W = W_n / [h_{W,WH} - h_{W,WC} - T_{BC} (s_{W,WH} - s_{W,WC})] m_W \quad (A-23)$$

where the "WC" subscript refers to water at the condenser temperature and its saturation pressure.

Efficiency Terminology. The efficiency of the system calculated by Equation A-22 is termed "thermal efficiency" and that calculated by Equation A-23 is termed "work efficiency". The term "thermodynamic efficiency" has been avoided because of its ambiguity.

Efficiency refers generally to the ratio of an actual process value to the maximum possible value of an ideal process. The term "thermodynamic efficiency" specifically refers to the ratio of work produced to some maximum possible value. Strictly speaking, the thermodynamic efficiency of the DCHE system is defined as the ratio of the net work output of the system to the total heat absorbed by the system, that is, the heat given up by the brine as it is cooled from the wellhead temperature to the DCHE column outlet temperature as given by Equation A-22. This ratio is also called "thermal efficiency" in thermodynamics, and this latter definition is used throughout this report.

Common usage of thermodynamic efficiency for a work producing system is to refer the net work output of the system to the maximum possible work output of an ideal system as given by Equation A-23. However, as noted above this definition is not strictly correct for the DCHE system. Therefore, the efficiency given by Equation A-20 is referred to as the "work efficiency".

One further possible ambiguity exists. The term thermal efficiency is most commonly used in connection with conventional shell and tube heat exchanges and is the ratio of the actual temperature change of the shell side fluid to the maximum possible temperature change. This definition clearly does not apply to the case of DCHE and thus, no confusion should arise.

Cost.

A detailed development of the cost equations used in this analysis is given in Appendix L. The time reference for all costs is 2nd quarter, 1975.

In modeling the plant costs both fixed and variable cost have been included. as described in Section 7. The fixed costs are the well costs, generator costs, plant buildings, etc. The variable costs are the heat exchanger and pump costs.

The most important factor in the heat exchanger cost is the size of the vessel needed. The basic equation for the cost of a single vessel is

$$C_v = 21.08 [136D(H+0.8D) (PD/1948.+0.205)]^{0.66} \quad (A-24)$$

This is Equation L-5 discussed in Appendix L. The cost of column internals or auxilizary equipment must be added to C_v to obtain the total DCHE direct materials cost

$$C_j = C_v + C_{I,j} \quad (A-25)$$

C_v is a function of height H and diameter D. The expressions relating H and D for the sieve tray, mixer settler, and packed columns are given below.

1. Sieve tray. The equations for the cross sectional area given full treatment in Section 7 of this report. In the region where liquid isobutane is flowing counter-currently with brine (below the critical point) the ratio of minimum total column area to brine low rate is given by Equation 7-18.

$$\frac{A_T}{m_W} = \frac{r}{1176\rho_B} + \frac{1}{535\sqrt{\rho_W(\rho_W - \rho_B)}} \quad (A-26)$$

Above the critical temperature, isobutane is in a vapor phase and water is in a liquid phase. Using vapor-liquid contacting column sizing criteria (Perry, 1973) the ratio of total column area to brine flow rate is

$$\frac{A_T}{m_W} = \frac{r}{890\sqrt{\rho_B(\rho_W - \rho_B)}} + \frac{1}{1390\rho_W} \quad (A-27)$$

To obtain the total cross sectional area of the sieve tray column the areas of all the stages are computed and the largest used. The height of the column, for 2-foot tray spacing is calculated using the equation

$$H = 2N/E + 15 \quad (A-28)$$

where N is the number of plates. It should be noted that the practical diameter of the vessel, due to transportation difficulties and practical shell thickness, is limited to 12 ft. Consequently, for large flow rates an equivalent number of 12 ft. diameter vessels is calculated. The values of D and H are then used in Equation A-24 for vessel cost. To this vessel price is added the cost, $7.83 ND^2$, of the column internals (Equation L-6) such as the trays and downcomers so that the total cost C_s of the sieve plate column is

$$C_s = (C_v + 7.83 ND^2) \times (\text{No. of columns}) \quad (\text{A-29})$$

2. Mixer-Settler. Theory of settling gives a relationship between the diameter and height of each stage as described in Appendix M.

Last stage (cool)

$$D H = \frac{450 m_W \left(\frac{r}{\rho_B} + \frac{1}{\rho_W} \right) \mu_W}{(\rho_W - \rho_B)} \quad (\text{A-30})$$

2nd Last Stage to critical point of isobutane

$$D H = \frac{m_W \left(r \frac{\rho_W}{\rho_B} + 1 \right)}{2100 \sqrt{\rho_W (\rho_W - \rho_B)}} \quad (\text{A-31})$$

Above the critical point

$$D = \frac{0.036 \sqrt{r m_W}}{[\rho_B (\rho_W - \rho_B)]^{0.256}} \quad (\text{A-32})$$

The diameter of each stage below the critical temperature is obtained using a fixed velocity criterion. The procedure is to set D initially at zero. Then the diameter is incremented and the velocity u calculated

$$u = \frac{m_W (1/\rho_W + r/\rho_B)}{900 \pi D^2} \quad (\text{A-33})$$

When u is less than 5 ft/sec the information is stored. The velocity is a function of temperature (via density) and therefore the diameter can vary somewhat from stage to stage. Once the diameter is established for each stage the length H of each stage can be calculated from Equations A-30 and A-31. Above the critical point the diameter is calculated from Equation A-32 and the vessel height H is 12 feet as discussed in Appendix M. Unlike the other heat exchange systems, the mixer settler system uses different size vessels for each stage based on the properties of each stage. Consequently, Equation A-24 is used to determine the cost of each stage separately. To these costs, pump costs and mixer costs are added, as discussed in Appendix L. The total cost for the mixer settler system as given by Equation L-11 is

$$C_M = \sum_{i=1}^N C_{v,i} + 24,000 N \quad (A-34)$$

3. Packed Column. The limiting condition on the packed column is the brine mass velocity 6000 lb/hr ft² for just below flooding conditions. The diameter satisfying this condition is used as the diameter of the vessel. The height is, assuming a height equivalent to a stage of 0.7 ft. for Goodloe-packing,

$$H = 0.7 N + 15 \quad (A-35)$$

The values of H and D are used in Equation A-24 to determine the cost of the vessel shell. Cost of the Goodloe packing at \$70 per ft.³ is $38.5ND^2$. Added to vessel cost, the total cost of the packed column heat exchanger is

$$C_p = [C_v + 38.5ND^2] \times (\text{No. of Columns}) \quad (A-36)$$

Total Cost. The installed cost of the heat exchange system is obtained by multiplying the exchanger cost from Equation A-29, A-34, or A-36 by an installation factor of 3.0. That is, installed cost C_D given by

$$C_D = 3.0C_j \quad (A-37)$$

where j refers to exchanger type. To the installed cost of the heat exchange system is added the plant costs C_P , the well costs C_W , the electric plant costs C_E , and the annual cost of the isobutane losses to give

$$C_A^\# = a_F(3.0C_j + C_P) + a_W C_W + a_E C_E + C_B^\# \quad (A-38)$$

Where the electric plant costs are given by

$$C_E = (0.2 + 0.4 W_e^{*0.6}) 10^6 \quad (7-32)$$

The fluid handling process plant cost is given by

$$C_P = \left\{ 2 + [2 + 0.01 (P - P_{BC})] [m_B/10^6]^{0.6} \right\} 10^6 \quad (7-33)$$

The well costs C_W are $\$17 \times 10^6$ and the isobutane costs $C_B^\#$ are $\$0.4 \times 10^6$ per year. The net cost of electricity is then

$$C_e = C_A^\# / 8760 W_e \quad (7-29)$$

The total capital cost is

$$C_T = C_D + C_P + C_W + C_E \quad (A-39)$$

Appendix B
System Analysis Computer Program

Program Description

Main Routine. The primary function of the Main routine is to enter data and provide the user with options to enter number of stages, flow rates, etc. The data entered are

on line 1:

- brine inlet temperature, °F, $T_{W,1}$
- estimated temperature difference at the bottom tray, °F, PREV
- condenser temperature, °F, T_{BC}
- brine inlet flow rate, lbs/hr, MA

on line 2:

- print control parameter IH; IH=1: work and cost analysis calculated and printed; IH=0: not calculated or printed

on line 3:

- data control parameter IG: IG=0: default; IG=1: stage number read in; IG=2: flow ratio read in; IG=3: both flow ratio and number of stages read in

on line 4:

- number of stages: initial (NA), final (NB), increment (NC)

on line 5:

- flow rate ratio, 100r: initial (IA), final (IB), increment (ID) on line 6:
- output control parameter OUTCNTL; OUTCNTL = 1: terminal output; OUTCNTL = 2: line printer output

on line 7:

- maximum iterations, MAXI, allowed for calculations of temperature from given enthalpy
- convergence criteria, EP (BTU/lb), for calculation of temperature from given enthalpy

- relaxation parameter, QZ, for calculation of temperature from given enthalpy
- weight percent of dissolved solids in brine, WP
- maximum iterations, MAT, allowed for convergence to condenser temperature

on line 8:

- column pressure, psia
- maximum iteration allowed for convergence to condenser temperature, NT
- convergence parameter, EPS(°F), for condenser temperature
- maximum iterations, MAXIT, allowed for stage properties convergence

on line 9:

- stage efficiencies

A sample input is given in Figure B-1.

This routine also manages the order in which all of the following subroutines are called and computes the cost of generating electricity according to Equation 7-29.

Subroutine FINISH. This subroutine computes the thermodynamic properties down the column as shown in Figure A-2. An explanation of the calculation has been given in the program description Appendix A. Figure B-2 shows the stage identification used in this subroutine. The objective of the routine is to calculate the temperatures, viscosities, densities, steam and water rates through the column.

Subroutine TEFF. The pump work terms Equations A-18 through A-20 are calculated and the results combined with the work output for the isobutane turbine from Subroutine Work according to Equation A-21. The efficiencies given by Equations A-22 and A-23 are calculated.

Subroutine WORK. The subroutine calculates the work output from the isobutane turbine as described in Appendix A. The turbine model consists of an isentropic expansion to the condenser pressure using Equations A-14 through A-17.

Subroutine ETHB. This subroutine calculates the enthalpy of isobutane. Typical values of the enthalpy are shown in Table A-2.

400., 21., 100., 1.E07

1

3

3, 12, 1

100, 100, 100

2

175, 0.1, 1., 3., 20

650., 80, 2., 175

1., 1., 1., 1., 1., 1., 1., 1., 1., 1., 1., 1., 1.

Figure B-1. Sample Computer Input.

WATER

ISOBUTANE

Figure B-2. Computer nomenclature for I^{th} stage of DCHE corresponding to scientific nomenclature of Figure 7-1.

Subroutine ENTB. This subroutine calculates the entropy of isobutane using the API correlation.

Subroutine PHASEB. This subroutine determines isobutane is in the liquid or vapor phase at the given conditions.

Subroutine COST. This subroutine sizes the vessels and computes column cost and the total plant cost. It uses other subroutines described below. The development of the cost equation is described in detail in the program description section and a flow chart is given in Figure B-3.

Subroutine VESCT. This subroutine calculates the minimum cost from the length to diameter product for the mixer-settler system. The procedure starts by incrementing diameters until one just large enough to satisfy the flow velocity criteria of 5 ft/sec is found. For this diameter the length is calculated in the function subroutine HT1 below. The vessel cost is then calculated in Function Subroutine VE.

Function HT1. This subroutine calculates the length requirement H for the mixer-settler system using Equations A-30 and A-31 or at 12 ft. as described in Appendix A. The diameter is calculated using Equation A-32 or A-33 and finally the capital cost C_M from Equation A-34.

Function DENS. This subroutine calculates the density of water and corrects it for dissolved solids using Equations D-8 and D-9 from Wahl (1977). Typical values are given in Table A-1.

Function VISC. This subroutine calculates the viscosity of water as a function of temperature and corrects for dissolved solids using Equations D-6 and D-7 which are taken from Wahl (1977). Typical values are given in Table A-1.

Function DENB. This subroutine calculates the density of isobutane from the equation of state. Typical values of density from this are given in Table A-2.

Function VE. This subroutine calculates the cost of a pressure vessel according to Equation A-24 as discussed in program description.

Figure B-3. Subroutine Cost

Subroutine WRTOUT. This subroutine prints the results in an appropriate format. The following are also calculated in this subroutine:

1. The net work produced from one pound of brine in BTU/lb which is W_n/m_w .
2. The capital cost per unit of net power output in \$/BTU which is C_T/W_n .
3. The cost of electricity produced in mils/kw, C_e^* .

A sample of the output is shown in Figure B-4.

Nomenclature

ACD	Cross section area of column, test parameter
ADCTG	Fixed plant cost (generator)
ADCTW	Fixed well cost
ADIF	Convergence parameter
AM	Brine flow rate (moles/hr)
AMEG	Power output in megawatts
AN	Isobutane flow rate (moles/hr), also coefficient used in column calc.
AR	Cross section area column
ARI	Cross sectiona area 12 ft diameter column
ATEM	Convergence parameter
B1	Term in Shultz expression for turbine power
B2	Term in Shultz expression for turbine power
B3	Term in Shultz expression for turbine power
BTLB	Output (BTU's per pound of brine)
CA	Cost of extras (mixer-settler)
CC	Installed cost of DCHE, C_D , \$
CCF	Ratio of Installed to Capital Cost
CCKW	Installed cost of exchanger per kilowatt
CDUTY	Condenser duty
CID	Identifier for thermodynamic properties subroutine
CMI	Cost of mixers in mixer-settler system
CMAT	Cost coefficient
CMBTU	Cost of plant per BTU

```

DIRECT CONTACT HEAT EXCHANGER
-----
WORKING FLUID: ISOBUTANE
-----
FLOW RATES - BRINE: 10000000.00  BUTANE: 10000000.00  FACTOR: 1.00
-----
INLET TEMPERATURE BRINE: 400.00  OUTLET TEMPERATURE BUTANE: 293.00
-----
EXCHANGER PRESSURE: 450.00
-----
NUMBER OF STAGES: 3
-----
CALCULATED CONDENSER TEMPERATURE: 99.92
-----

STAGE      1      2      3      4      5
-----
STD EFF      1.00      1.00      1.00      0.00      0.00
TEMP BUTANE  293.00  244.01  186.35  99.92    0.00
-----
TEMP WATER   293.00  244.01  186.35    0.00    0.00
-----
BUTANE
ENTHALPY      338.43  262.43  215.35    0.00    0.00
-----
STEAM
ENTHALPY     1176.27  1140.95  1138.95    0.00    0.00
-----
WATER
ENTHALPY      259.60  210.48  152.04    0.00    0.00
-----
STEAM RATE   319039.04  97.52   17.78    0.00    0.00
-----
WATER VAPOR
PRESSURE      40.45    0.00    0.00    0.00    0.00
-----

TURBINE INLET      TURBINE OUTLET
-----
TEMPERATURE, DEG F: 293.8      170.7
PRESSURE, PSIA :    450.0      75.0
FLOW RATE, 1000 LB/HR
ISOBUTANE :        10000.0      10000.0
STEAM :            319.0      187.0
WATER :             0.0      132.1

ISENTROPIC OUTLET TEMPERATURE, DEG F: 154.4

CONDENSER
-----
TEMPERATURE, DEG F: 100.0
PRESSURE, PSIA:    72.0
DUTY, MMH:        1904.5
TYPE OF COLUMN: 1
-----
1 - STEW TRAY
2 - MIXER SETTLER
3 - PACKED COLUMN

WORK REQUIREMENTS AND OUTPUTS (MM BTU/HR)
-----
BRINE PUMP: 19.37  BUTANE PUMP: 44.52  MIXER PUMP: 0.00
BUTANE TURBINE: 244.61  BRINE TURBINE: 15.09
THERMAL EFFICIENCY: 9.39  WORK EFFICIENCY 31.50  MEGAWATTS: 51.62

COST ANALYSIS
-----
EQUIVALENT 12FT. DIAMETERS: 7.27  HEIGHT: 23.57
INSTALLED COST OF EXCHANGER- 1759031.07
TOTAL CAPITAL COST- 33055787.21
ANNUAL COST- 8887984.64
MILS/KLW- 19.4548543
INSTALLED COST EXCHANGER/KLW- 34.08
BTU/LB- 19.58
COST/BTU- .0453892

```

Figure B-4. Sample output from the system analysis computer program

COF	Coefficient in sieve tray cost analysis
COMEGA	Eccentric factor for isobutane
COSGR	Factor for plant cost scale up based on electric power
CP	Heat capacity of brine
CRB	Heat capacity ratio of isobutane
CRW	Heat capacity ratio for water
DBC	Density isobutane
DEB 1	Density isobutane (some average)
DEB 2	Density isobutane (some average)
DEB 3	Density isobutane (some average)
DELT	Iteration parameter
DELTB	Iteration parameter
DIFTEM	Temperature difference (iteration monitor)
DQ	Stage diameter storage
DV1DWW	Density brine
EF1	Efficiency brine pump
EF2	Efficiency butane pump
EF3	Efficiency brine expander
EFFW	Work efficiency
EFGEN	Electric generator efficiency
EFT	Efficiency butane turbine
EFSC	Ratio of equilibrium to actual plates for sieve tray column
ELOUT	Electric power output
EP	Iteration accuracy parameter
EPS	Iteration accuracy parameter
EQU12	Equivalent 12 ft diameter
EQUIMIX	Stage efficiency
FAC	Mass flowrate ratio, isobutane to brine
IA	Flow rate isobutane (minimum studied)
IB	Flow rate isobutane (maximum studied)
IC	Flow rate increment
JC	Parameter used in thermodynamic calculations

H	Height of column
HB	Enthalpy isobutane
HCD	Heat capacity combination term
HETP	Height of an equilibrium stage, packed column, ft
HSF	Output butane enthalpy
HW	Enthalpy brine
MA	Brine flow rate
MAT	Iteration parameter
MAZI	Iteration parameter
MAXIT	Iteration parameter
MB	Butane flow rate
MOB	Molecular weight isobutane
MOLW	Molecular weight isobutane
MOW	Molecular weight water
N	Number of equilibrium stages
NA	Minimum number of stages studied
NB	Maximum number of stages studied
NC	Stage increment
NGC	Number of substances in hydrocarbon stream
NP	Number of pumps per stage
NT	Iteration parameter
PAC	Pressure (some average)
PACP	Pressure (some average)
PB	Isobutane partial pressure
PC	Condenser pressure, also critical pressure of isobutane
PCR	Critical pressure of isobutane
PCl	Water condenser pressure
PL	Power requirement for mixer pumps (kilowatts)
PPL	Price of packing (packed column)
PR	Power requirement for mixer pumps (BTU)
PRA	Power requirement for mixer pump (Kw)
PREV	Temperature difference last stage
PSS	Pressure storage term
PT	Turbine discharge pressure, psia
PY	Design pressure of column

QZ	Iteration parameter
R	Gas constant
RHC	Heat capacity ratio
STC	Extra internal cost for the sieve tray
T	Stage temperature
Tl	Butane outlet temperature ($^{\circ}\text{R}$)
TA	Brine inlet temperature
TAV	Temperature (average)
TAVP	Temperature (average)
TAX	Brine inlet temperature
TB	Butane column exit temperature
TC	Condenser temperature, also total cost of vessel
TCT	Total Installed capital cost, \$
TY	Critical temperature of working fluid, = TC in SSI subroutine
TD	Temperature
TEM	Assumed stage temperature
TEN	Temperature storage term
THEFF	Thermal Efficiency, $100 E_{th}$, %
TOUT2	Turbine discharge temperature
TS1	Capital cost, C_F , \$
TSFF	Cost of electricity, \$/MW or mils/kw
TSS	Annual cost, C_A , \$
VISCOS	Viscosity of brine
VL	Limiting velocity mixer-settler system
W	Turbine work output (BTU)
W1	Brine pump work requirement (BTU)
W2	Isobutane pump requirement (BTU)
W3	Brine expander power output (BTU)
WA	Brine expander power output (BTU)
W1A	Brine pump work requirement (10^6 BTU)
W2A	Isobutane pump work requirement (10^6 BTU)
W3A	Brine expander power output (10^6 BTU)

WNET	Net power output (BTU)
WP	Percent dissolved solids
X	Liquid of water fraction
XNDR	Fraction hydrocarbon isobutane
Y	Steam fraction of water
YSF	Final steam rate
ZB	Compressibility of isobutane

Appendix C

Derivation of Tray Efficiency Equation

Referring to Figure 7-1, the efficiency of the i^{th} heat transfer tray is defined as the approach of the brine temperature $T_{W,i}$ to the equilibrium tray temperature, analogous to mass transfer tray efficiency as the approach to the equilibrium composition. Thus, the heat transfer efficiency E_i of the i^{th} plate is given by

$$E_i = (T_{W,i-1} - T_{W,i}) / (T_{W,i-1} - T_{*,i}) \quad (\text{C-1})$$

where $T_{*,i}$ is the equilibrium temperature of the i^{th} plate.

At equilibrium, the temperature of both the isobutane and brine leaving the tray will be equal to the equilibrium temperature $T_{*,i}$. A heat balance around the i^{th} plate gives

$$m_W c_W (T_{W,i-1} - T_{*,i}) = m_B c_B (T_{*,i} - T_{B,i+1}) \quad (\text{C-2})$$

At equal mass flow rates, $m_W c_W$ is approximately equal to $m_B c_B$. With this assumption, Equation C-2 can be simplified and solved for an approximate value of the equilibrium temperature

$$T_{*,i} \doteq (T_{W,i-1} + T_{B,i+1}) / 2 \quad (\text{C-3})$$

A heat balance around plate i for the non-equilibrium condition can be written

$$m_W c_W (T_{W,i-1} - T_{W,i}) = m_B c_B (T_{B,i} - T_{B,i+1}) \quad (\text{C-4})$$

Again assuming $m_W c_W$ is approximately equal to $m_B c_B$, Equation C-4 becomes

$$T_{W,i-1} + T_{B,i+1} \doteq T_{W,i} + T_{B,i} \quad (\text{C-5})$$

Thus by substituting Equation C-5, Equation C-3 can also be expressed as

$$T_{*,i} \doteq (T_{W,i} + T_{B,i})/2 \quad (C-6)$$

Using the approximate value of the equilibrium temperature $T_{*,i}$ given by Equation C-6, the tray efficiency defined by Equation C-1 becomes

$$E_i = (T_{W,i-1} - T_{W,i})/[T_{W,i-1} - (T_{W,i} + T_{B,i})/2] \quad (C-7)$$

Appendix D

Miscellaneous Experimental Results

1. Temperature Calibrations

Several runs were made in which only water was circulated through the column section and associated piping. Data was recorded after steady state temperatures had been reached. Since no heat transfer occurred except for ambient losses, a constant water temperature was obtained throughout the column and inlet and outlet piping. A bleed stream (about one quarter of the water circulation rate) was removed from the overhead butane line to ensure that the constant temperature water was in contact with the overhead system RTD's. The temperature readings for the column and inlet and outlet sensors were averaged to obtain the steady state value of the constant water temperature.

Data was obtained for steady state nominal temperatures of 120°F, 150°F and 200°F for the packed column and 130°F and 180°F for the sieve tray column. The arithmetic average of the temperature indicator (TI) readings was computed as the true value of the water temperature for each run. Additive corrections were calculated for each TI as the difference of the TI reading and the computed true water temperature.

The raw temperature data, computed true water temperature, and the temperature correction for each TI is presented in the following tables in the form $TE(J) = TI(J) + TCAL(J)$, corresponding to the computer program nomenclature used in Appendix J.

Where

$TE(J)$ is the corrected experimental temperature, °F

$TI(J)$ is the actual TI reading, °F

$TCAL(J)$ is the correction, °F

Tables D-1, D-2 and D-3 summarize the data for the packed column runs and Tables D-4 and D-5 present the sieve tray data.

Table D-6 presents the adjusted final temperature corrections used in this study for the packed column section and Table D-7 presents the adjusted corrections for the sieve tray column.

Table D-1

Packed Column Data

Nominal 120°F Run

<u>J = TI No</u>	<u>TI (J)</u>	<u>TE (J)</u>	<u>TCAL (J)</u>
7	124.5	123.8	-0.7
8	122.5	123.8	1.3
9	124.0	123.8	-0.2
18	123.7	123.8	0.1
19	124.4	123.8	-0.6
20	123.4	123.8	0.4
21	123.5	123.8	0.3
22	122.8	123.8	1.0
23	124.2	123.8	-0.4

Table D-2

Packed Column Data

Nominal 150°F Run

<u>J = TI No.</u>	<u>TI (J)</u>	<u>TE (J)</u>	<u>TCAL (J)</u>
7	155.2	153.9	-1.3
8	152.8	153.9	1.1
9	154.1	153.9	-0.2
18	153.9	153.9	0.0
19	154.5	153.9	-0.6
20	153.6	153.9	0.3
21	153.8	153.9	0.1
22	153.0	153.9	0.9
23	154.0	153.9	-0.1

Table D-3

Packed Column Data

Nominal 200°F Run

<u>J = TI No.</u>	<u>TI (J)</u>	<u>TE (J)</u>	<u>TCAL (J)</u>
7	197.7	196.3	-1.4
8	194.9	196.3	1.4
9	196.4	196.3	-0.1
18	196.5	196.3	-0.2
19	197.0	196.3	-0.7
20	196.1	196.3	0.2
21	196.4	196.3	-0.1
22	195.4	196.3	0.9
23	196.1	196.3	0.2

Table D-4

Sieve Tray Data

Nominal 130°F Run

<u>J = TI No.</u>	<u>TI (J)</u>	<u>TE (J)</u>	<u>TCAL (J)</u>
7	134.0	133.1	-0.9
8	130.7	133.1	2.4
9	133.2	133.1	-0.1
10	133.3	133.1	-0.2
11	133.2	133.1	-0.1
12	135.0	133.1	-1.9
13	134.1	133.1	-1.0
14	133.2	133.1	-0.1
15	133.2	133.1	-0.1
16	133.0	133.1	0.1
17	132.9	133.1	0.2
18	133.0	133.1	0.1
19	133.1	133.1	0
20	133.0	133.1	0.1
21	133.0	133.1	0.1
22	131.8	133.1	1.3
23	133.3	133.1	-0.2

Table D-5

Sieve Tray Data

Nominal 180°F Run

<u>J = TI No.</u>	<u>TI (J)</u>	<u>TE (J)</u>	<u>TCAL (J)</u>
7	184.2	182.7	-1.5
8	179.2	182.7	3.5
9	182.6	182.7	0.1
10	183.4	182.7	-0.7
11	182.8	182.7	-0.1
12	185.0	182.7	-2.3
13	184.1	182.7	-1.4
14	183.1	182.7	-0.4
15	183.0	182.7	-0.3
16	182.7	182.7	0
17	182.4	182.7	0.3
18	182.5	182.7	0.2
19	182.5	182.7	0.2
20	182.4	182.7	0.3
21	182.2	182.7	0.5
22	181.0	182.7	1.7
23	182.3	182.7	0.4

Table D-6

Packed Column

Temperature Corrections

<u>J = TI No.</u>	<u>TCAL (J)</u>
7	-0.7
8	1.5
9	-0.1
18	0.0
19	-0.6
20	0.3
21	0.1
22	0.9
23	0.2

Table D-7

Sieve Tray Column

Temperature Corrections

<u>J = TI No.</u>	<u>TCAL (J)</u>
7	-0.7
8	1.5
9	-0.1
10	-0.5
11	-0.1
12	-2.1
13	-1.2
14	-0.3
15	-0.2
16	0.0
17	0.2
18	0.1
19	0.1
20	0.2
21	0.3
22	0.9
23	0.2

Table D-6 presents a composite set of temperature corrections obtained from the three temperature level runs given on Tables D-1, D-2 and D-3 for the packed section. Table D-7 presents the results from Tables D-4 and D-5 for the sieve tray section. Consideration was also given to column heat losses, discussed in Appendix D-4.

The inlet temperature sensor, TI-7, appears to read about 0.7°F high after including estimated heat losses. Estimated column inlet heat losses account for about $0.7\text{-}0.8^{\circ}\text{F}$ on the inlet brine stream at an operating temperature of $250\text{-}300^{\circ}\text{F}$ in the column. Thus the larger difference between measured value and computed constant temperature for TI-7 shown in Tables D-1 through D-5 can be attributed to this heat loss. All other TI readings, however, are assumed to be reading the computed true temperature, and the temperature corrections are accordingly averaged for the runs. Using these correction factors gives a value of corrected temperature $\pm 0.1^{\circ}\text{F}$.

The remaining temperature sensors throughout the system (those not shown above) are important to the operation of the unit, but are not involved in the precise calculations required for determining the performance of the DCHE column. Therefore a calibration of these other sensors was not made and accuracy of $\pm 1^{\circ}\text{F}$ was assumed for their use.

2. Flow Measurement Equations and Calibration

The most commonly used of the flow measurement devices by differential pressure is the orifice meter. The orifice meter when properly and carefully installed and when using a design within specified tolerances, can obtain an overall accuracy of measurement of ± 0.75 percent. With flow calibration, accuracies of ± 0.25 percent can be achieved.

The orifice meter consists of a differential-pressure measuring device connected to an orifice flange by suitable piping. The orifice meter tube, called a meter run, consists of upstream and downstream sections of pipe whose diameter is determined by flow rate. The length and tolerances of the tube must be within specifications. These specifications are given in the American Gas Association Gas Measurement Committee Report No. 3, and in the ASME Fluid Meters Reference Book Parts 1 and 2.

The orifice plate is held by the flanges in a plane 90° to the direction of the flow. Bore, circumference, edge, and tolerances must all meet specifications. Purchase of a meter run from a reliable company such as Daniel Industries to meet these specifications is sufficient.

The flow of any liquid through an orifice is determined by the following equation:

$$w = \frac{CaF_a n}{1 - \beta^4} \sqrt{2 gh} \quad (D-1)$$

With the units in American practice, the equation can be written:

$$w_h = 359 K d^2 F_a \sqrt{h_w \rho} \quad (D-2)$$

w_h = rate of flow, lb/hr

K = flow coefficient

a = orifice area, in²

d = orifice diameter, in.

D = pipe inside diameter, in.

β = diameter ratio, d/D

h = differential pressure

h_w = differential pressure, inches of 68°F H₂O

ρ = density of flowing fluid at the inlet of the orifice, lb/ft³

F_a = thermal expansion factor of the orifice plate

A 1-inch Daniel orifice flange honed flow section is used for both brine (FC-1) and isobutane (FC-2). These units in Schedule 80 tubes need no flow calibration to assure accuracy of ± 0.75 percent. (With calibration, Daniel honed flow sections could achieve flow coefficient accuracy of ± 0.25 percent or better.) The flow coefficient, K, is computed by the equation:

$$K = 0.5991 + \frac{0.0044}{D} + (0.2155 + \frac{0.0175}{D}) (\beta^4 + 2\beta^{16}) \\ + [\frac{0.00052}{D} - 0.000192 + (0.01648 - \frac{0.00116}{D}) (\beta^4 + 4\beta^{16})] \cdot \frac{1000}{\sqrt{R_D}} \quad (D-3)$$

where K = flow coefficient

D = inside diameter of meter tube, in.

β = orifice to tube diameter ratio d/D

R_D = meter tube Reynolds number

This equation has been found to give coefficients within ± 0.75 percent of the values obtained from a calibration where pressures are measured from corner grooves and where D is between 0.5- and 1.5-in., β is between 0.1 and 0.8, and R_D is greater than 1000.

Flow Meter Calibration. A calibration was made of the differential pressure transmitters for FC-1 and FC-2 using a monometer and manifold system to apply an accurately known differential pressure from 0 to 30- inches of water to the transmitters. These pressures were recorded vs. the pot setting reading of the panel mounted controllers.

Table D-8 presents the raw data for FC-1; and Table D-9 presents the data for FC-2. A plot of the data is shown in Figure D-1. Smoothed results are given in Tables D-10, and D-11. Tables D-10 and D-11 also include a correction for water density from the measured temperature to 68°F.

Table D-8

FC-1 (Brine Flow) Calibration Data

<u>In. H₂O (at 91°F)</u>	<u>Digital Knob - Pot Setting</u>
0	22
1	48
2	86
4	157
6	223
8	293
10	358
15	528
20	695
25	865
27	928

Table D-9

FC-2 (i-C₄ Flow) Calibration Data

<u>In. H₂O (at 92°F)</u>	<u>Digital Knob - Pot Setting</u>
0	10
1	19
2	62
4	123
6	189
8	253
10	312
15	472
20	635
24	760
28	890
30	950

- D10 -

Figure D-1. Flow meter calibration. Pot setting vs. in. of H₂O.

Table D-10

Adjusted Results, FC-1 (Brine Flow)

<u>In. H₂O (at 68°F)</u>	<u>Digital Knob - Pot Setting</u>
0	20
1	54
2	88
4	155
6	223
8	290
10	358
15	527
20	696
25	865
27	932
29	1000

Table D-11

<u>In. H₂O (at 68°F)</u>	<u>Digital Knob - Pot Setting</u>
0	0
1	31
2	63
4	126
6	190
8	254
10	317
15	477
20	636
24	763
28	891
30	954
31.4	1000

The adjusted values of the digital knob-pot setting agree with the raw data values well within the accuracy of the manometer readings. The adjusted values shown on Tables D-10 and D-11 may therefore be taken as an accurate indication of the differential pressure, and when these calibrated readings are combined with Equations D-2 and D-3 will give calculated flow rates with an accuracy of ± 0.75 percent.

PHYSICAL PROPERTIES OF BRINE AND ISOBUTANE USED IN FLOW EQUATIONS

Isobutane. Viscosity and density were estimated from data in the API Technical Data Book -- Petroleum Refining. Data is summarized in Table D-12 below, and fitted to Equations D-4 and D-5. Plots of these data are presented on Figures D-2 and D-3.

Brine. Viscosity and density data for geothermal brines were obtained from Reference 1 and are represented by Equations D-6 through D-9.

At 650 psig both the density and viscosity of isobutane show corrections of about 3-5 percent from atmospheric pressure data. Table D-12 and Equations D-4 and D-5 below, represent viscosity and density data, respectively, at 650 psig.

Table D-12

Isobutane Density (ρ) and Viscosity (μ) at 650 psia

<u>Temperature, °F</u>	<u>ρ, lb/ft³</u>	<u>μ, centipoise</u>
75	35.16	0.165
125	33.05	0.128
175	30.37	0.096
225	27.13	0.070
275 (critical)	18.18	0.027

DENSITY
LB/FT³

Figure D-2. Isobutane density at 650 psia.

Figure D-3. Isobutane viscosity at 650 psia.

Isobutane viscosity at 650 psig, for $60^{\circ}\text{F} < T < 250^{\circ}\text{F}$:

$$\mu_{\beta} (\text{lb/ft-sec}) = \left(\frac{231.6}{T^* (^{\circ}\text{R})} - 0.268 \right) (6.72 \times 10^{-4}) \quad (\text{D-4})$$

Isobutane density at 650 psig, for $60^{\circ}\text{F} < T < 225^{\circ}\text{F}$:

$$\begin{aligned} \rho_{\beta} (\text{lb/ft}^3) = & 35.16 - 1.822 \left(\frac{T-75}{50} \right) - 0.29 \left(\frac{T-75}{50} \right)^2 \\ & + 0.00167 \left(\frac{T-75}{50} \right)^3 \end{aligned} \quad (\text{D-5})$$

Brine viscosity:

$$\log \mu' = -2.03 + 1008/T^* \quad (\text{D-6})$$

$$\mu'_w = \mu' (1 + 0.021 w_t + 0.0027 w_t^2) \quad (\text{D-7})$$

where

μ' = viscosity of water, centipoise

μ'_w = brine viscosity, centipoise

T = temperature, $^{\circ}\text{F}$

T^* = temperature, $^{\circ}\text{R}$

w_t = weight percent of total dissolved solids

Brine density:

$$\rho' = (1 + d\tau^{1/3} + e\tau) / (V_c + a\tau^{1/3} + b\tau + c^4) \quad (\text{D-8})$$

$$\rho'_w = \rho' + 0.0073 [1 + 1.6 \times 10^{-6} (T-273)^2] w_t \quad (\text{D-9})$$

where

ρ' = density of water, g/cm^3

ρ'_w = density of brine, g/cm^3

$\tau = 647.11 - T$

T' = temperature, $^{\circ}\text{K}$

$V_c = 3.1975 \text{ cm}^3/\text{g}$

$a = -0.315148$

$b = -1.203374 \times 10^{-3}$

$c = +7.48908 \times 10^{-13}$

$$d = 0.132489$$

$$e = 3.946263 \times 10^{-3}$$

$$w_t = \text{weight percent of total dissolved solids}$$

The error in isobutane density is less than 2 percent at 225°F. At lower temperatures Eq. D-5 is within 1% of experimental data. Eq. D-4 represents pure isobutane viscosity within ± 5 percent. Brine density as given by Eq. D-9 is less than 2 percent in error for typical geothermal brines as concentrated as 3.5 weight percent. Brine viscosity Eq. D-7 is based on typical geothermal brine analyses. Modifications are required for abnormally high concentrations of calcium chloride as in the Salton Sea region. (Wahl, 1977).

3. Pressure Transmitter Calibrations

The pressure indicator controllers PC-1 (DCHE pressure) and PC-3 (iso-butane condenser pressure) were calibrated on site by Rosemount engineers using a precision gage with an accuracy of ± 2 psi. Readings were taken first starting with low pressure and increasing, then starting at high pressure and decreasing. Tables D-13 and D-14 summarize the results of the calibration readings, and Figure D-4 presents a plot of these results.

Table D-13

PC-1 Data - DCHE Column Pressure

<u>Pressure, psig</u>	<u>Digital Knob - Pot Setting</u>
260	0
360	240
460	485
560	727
660	970

Table D-14

PC-3 Data - Iso-butane Condenser Pressure

<u>Pressure, psig</u>	<u>Digital Knob - Pot Setting</u>
100	100
200	219
300	329
400	440
500	550
600	661
700	772

Figure D-4. Digital knob-pot settling vs. control pressure for PC-1 and PC-3.

4. Column Heat Loss Estimation

Table D-15 presents a summary of the data for two similar isothermal runs, circulating water only through the packed column DCHE at a nominal 4 gpm and 200°F. The ambient temperature was 85°F, and there was a moderate wind of about 5-10 knots.

Under these conditions, the data, TE for J of 18 through 23, indicate nil heat loss from the column. However TE (7) shows there is a small loss associated with the brine inlet arrangement. From the data in Table D-15, UA for this section is estimated as 25 BTU/°F-Hr, where the temperature difference is TE (7) - T (ambient). Thus the heat loss for this section (Q_L) is

$$Q_L = 25 [TE(7) - T (\text{ambient})] \text{ BTU/HR} \quad (\text{D-10})$$

This heat loss applies only to the data from the LaVerne test runs. When the unit was moved to East Mesa, the inlet section was thoroughly insulated and the heat loss became negligible.

Table D-15
Isothermal Data

Run No. 1			Run No. 2	
J	TI(J), °F	TE(J), °F	TI(J), °F	TE(J), °F
7	198.5	197.8	198.7	198.0
18	196.5	196.5	196.7	196.7
19	197.0	196.4	197.4	196.8
20	196.1	196.4	197.4	196.8
21	196.4	196.5	196.7	196.8
22	195.5	196.4	195.9	196.8
23	196.1	196.3	196.6	196.8
P, psig		645		648
Water rate, lb/hr		2,146		2,169
T(ambient), °F		85		85
Wind, knots		5-10		5-10

Appendix E

Heat Transfer at Isobutane Inlet

The isobutane inlet piping configuration is shown in Figure E-1. This surface must be taken into account in the heat transfer calculations because of the sensitivity of the DCHE efficiency to the inlet and outlet temperatures at the bottom of the packing (or at the bottom sieve tray).

For $r=1$ the brine is cooled 1°F before it leaves the column, and the isobutane is heated 2°F before it reaches the bottom of the packing (or bottom tray).

Clean overall heat transfer coefficients were estimated as a function of the brine flow rate for equal ($\pm 30\%$) mass flow rates of brine and isobutane. These coefficients are based on the outside area of the isobutane inlet piping and are summarized in Table E-1. The outside surface area of the inlet pipe is 1.30 ft^2 and the distributor adds 1.185 ft^2 . Thus the area for the sieve tray operation is 1.30 ft^2 and for the packed column operation, 2.485 ft^2 .

NOTES:

1. ISOBUTANE INLET PIPE IS 1" SCH 40 C. S.
2. INLET AND DISTRIBUTOR SHOWN FOR PACKED COLUMN INSTALLATION. WHEN SIEVE TRAYS ARE IN PLACE, THE DISTRIBUTOR IS REMOVED.

Figure E-1. Bottom column section showing isobutane inlet piping details.

Table E-1

Estimated Heat Transfer Coefficients for Clean Isobutane Inlet Piping

Brine Flow Rate, m_w (lb/hr)	Packed Column Coefficient, U_p For Inlet Pipe Plus Distributor (BTU/hr-ft ² -°F)	Sieve Tray Coefficient, U_s For Inlet Pipe Only (BTU/hr-ft ² -°F)
1500	19.4	31.6
3000	34.1	54.0
4500	48.0	75.0

Combining these coefficients with the appropriate area and fitting to a polynomial results in the following equations for the UA product:

Packed Column

$$(UA)_p = 57.3 \left(\frac{m_w}{1500}\right) - 1.85 \left(\frac{m_w}{1500}\right)^2 + 1.6 \left(\frac{m_w}{1500}\right)^3 \quad (E-1)$$

Sieve Trays

$$(UA)_s = 50.4 \left(\frac{m_w}{1500}\right) - 11.0 \left(\frac{m_w}{1500}\right)^2 + 1.7 \left(\frac{m_w}{1500}\right)^3 \quad (E-2)$$

where

$$UA = \frac{Q}{\Delta T} \quad (E-3)$$

In these equations

- U = overall heat transfer coefficient, BTU/hr-ft²-°F
- A = outside surface area, ft²
- m_w = brine flowrate, lb/hr
- Q = heat transferred, BTU/hr
- ΔT = TE(22) - TE(24), temperature difference of the outlet brine and inlet isobutane
- TE(J) = corrected experimental temperature, °F

Appendix F

Experimental Apparatus

Figure F-1 shows the piping and instrument diagram (P&ID) for the experimental apparatus. Details of the piping plan and elevation are shown in Figures F-2 and F-3. Electrical connections and instrument console details are shown on the electrical single-line diagram Figure F-4.

Process flow and control diagrams for the unit are given in Section 5.

Packed Column Section, Bottom Surge, and Overhead Section, V-02

Construction details for the packed column are shown in Figure F-5. This drawing also shows the details of the bottom surge section and column supports and the overhead feed and vapor separation section. These latter sections are used for both the packed column and sieve tray column installations.

The bottom surge compartment provides 2 minutes residence time for a typical brine rate of 5 gpm, and the downward velocity will not entrain isobutane droplets larger than 120 microns so that good separation of the brine and isobutane is achieved.

The overhead section is designed to minimize entrainment of water and is also provided with a demister to remove any droplets which may have been carried upward in the isobutane stream.

Sieve Tray Section V-02

Figures F-6 and F-7 show the construction drawings for the sieve tray section. Connections for the RTD's are located so that the installed RTD is in the downcomer outlet and measures the temperature of the brine flowing from the tray above through the downcomer. The sieve tray layout is designed to handle up to 9 gpm of brine by the procedure described in Section 7.

The top and bottom sections shown in Figure F-5 are also used for the sieve tray column installation.

Butane Accumulator V-01

Details of the isobutane accumulator are shown in Figure F-8. This vessel has 115 gallons of surge for the circulating isobutane and is designed to thoroughly separate condensed water from the isobutane and remove the water from the bottom of the vessel.

Figure F-1. Piping and Instrument Diagram.

PLAN
SCALE 1/4" = 1'-0"

TYPICAL (BC) SAMPLE CONNECTION
DETAIL
NO SCALE

ELEVATION DETAIL AT HE 1 AND HE 2

PIPING MATERIAL SPECIFICATION

5" SCH 40 (5.31 IN) E & B (1.31 IN)
 PIPE
 FITTING: 1/2" 16 SW-70S FLARE
 3/4" SCH 40 BW
 FLANGES: 1/2" 16 SW-70S
 GASKETS: 1/2" 16 SW-70S
 O-RINGS: 1/2" 16 SW-70S
 VALVES: 1/2" 16 SW-70S
 GATE 3" 5000# FLG 1/2" 16 SW-70S
 CHECK 1/2" 16 SW-70S
 NOTE: THREADED JOINTS TO BE 55AL WELDED

SPEC B (BRINE)

PIPE - SCH 80 SMLS TR ASTM A-55
FITTINGS - 1/2" 3000 PS SCREWED
- 5/4" SCH 80 B.W.
UNIONS - 1/2" 3000 PS SCREWED STL/STL SEAT
FLANGES - 1/2" 600 RF SCREWED
- 5" 600 RF 90 (UNLESS NOTED)
REDUCTIONS - 1/2" FULL SIZE R BUSH. (EXCEPT AS NOTED)
GASKETS - FLEXISTAL STYLE - CAP
VALVES - GATE VOLT 5-8855 THRU 5-8888 OR 10
- GLOBE (WV) - 855 THRU 858 OR 10
- CHECK - 705 THRU 708 OR 10

SPEC. CW (COOLING WATER)

PIPE - SCH 40 BLK TB ASTM A-53
FITTINGS - 150# W/ SCREWED
UNIONS - 150# W/ SCREWED OR JT
REDUCTIONS-REDUCING FEE
VALVE - GATE POWELL FIG 580 SCED OR 58
- CHECK - FIG 116 SCED OR 10

SPDC 1A (INST. AIR)

PIPE - SCH 40 GALV T.E. ASTM A-53
FITTINGS - 150# MI SCED GALV
REDUCTIONS - REDUCING TEES

REVISION	DATE	DESCRIPTION	APPROVED
1	4-12-76	GEN. DIVISION

OCCEIDENTAL RESEARCH CORPORATION
A SUBSIDIARY OF OCCEIDENTAL PETROLEUM CORPORATION

PROJECT "854-DI OZDA DIRECT CONTACT HEAT EXCHANGER"

PIPING PLAN & DETAILS		
Drawn by A.T.	Checked by	Checked by
Date 9-18-76	Approved	554-H-001-3

Figure F-2. Piping Plan and Details.

4-13-76	GEN. REVISION
ORV OROCESTRAL RESEARCH CORPORATION <small>10000 N. 10th St., Suite 100, Phoenix, AZ 85020</small>	
PROJECT: SEA-ON KEDA DIRECT CONTACT HEAT EXCHANGER	
FILE: PIPING SECTIONS	
DATE: 3-15-76	DRAWN BY: 554-H-002-3

Figure F-3. Piping Sections.

ALARM SYSTEM AND MANUAL DISCONNECT WHEN ACTIVATED OPENS CONTACTS IN MAIN FEEDER UNIT, MANUALLY RESTARTED BY THE SYSTEM RESTART BUTTON.

- NOTES:
1. IS AREA TO BE CLASSIFIED - CLASS 1, DIVISION 2, GROUP D (HAZARDOUS AREA)
 2. ELECTRICAL INSTALLATION TO CONFORM TO CURRENT NATIONAL ELECTRICAL CODE
 3. AIR CONDITIONER, INSTRUMENT CONSOLE, CONTROL HOUSE LIGHTING AND OUTLET TO BE SUPPLIED BY G.R.C.
 4. LA VERNE INSTALLATION GENERATOR AND HEATER TO BE SUPPLIED BY G.R.C.

ERDA DCHE CONTROL HOUSE BASIC FLOOR PLAN - TOP VIEW

LA VERNE INSTALLATION

3	4/7/78	ADD NOTES 3 & 4	
REVISION	DATE	DESCRIPTION	APPROVED
2	3/31/78	REDRAW
OCCIDENTAL RESEARCH CORPORATION <small>A DIVISION OF OXYCORP, INC. (A DIVISION OF OXYCORP, INC.)</small> <small>LA VERNE, CALIFORNIA 91760</small>			
FOR	PROJECT 554-01 ERDA DIRECT CONTACT HEAT EXCHANGER		
TITLE	ELECTRICAL SINGLE-LINE DIAGRAM AND CONTROL HOUSE LAYOUT		
DRAWN BY	E.J.R.	DESIGNED BY	DRAWING NUMBER
DATE	3/31/78	APPROVAL	554-N-001-2

Figure F-4. Electrical Single-Line Diagram and Inst. Console.

Figure F-5. Packed Column for Occidental Research Corporation.

Figure F-6. 8" Diameter Column V-02 Assembly and Details.

Figure F-7. Mini Tray Assembly and Details for 8"-
Sch 40 Column V-02.

Figure F-8. V-01 Isobutane Accumulator.

Appendix G

Experimental Operations

Preliminary Runs with Isobutane and Water

A series of runs were made at La Verne from August through November, 1976, to test the system, calibrate the instruments and obtain performance and flooding data for the DCHE using isobutane and city water.

Figure 5-3 shows the process flow for the apparatus while operating at La Verne.

A temperature calibration run and two shakedown performance runs were made with the sieve tray section. The remainder of the runs were made using the packed column section. Three supercritical runs were made with varying isobutane to water flow rate ratios. The capacity of the water heaters HE-4 limited the flow rate at supercritical conditions to 3500 lb/hr-ft^2 of water. Consequently supercritical flooding runs were not possible. However a series of runs were made at lower water inlet temperatures ($200\text{-}240^\circ\text{F}$) and water rates up to 4800 lb/hr-ft^2 . At the 4800 lb/hr-ft^2 water rate the packed column flooded at an isobutane rate of 5500 lb/hr-ft^2 . For an inlet water temperature of 240°F , the column flooded by carrying water overhead with the isobutane, while an inlet water temperature of 185°F caused the column to flood by isobutane carryunder with the outlet water. Later supercritical runs at East Mesa using 340°F brine gave stable column operation with a brine rate of 7500 lb/hr-ft^2 and an isobutane rate of 4400 lb/hr-ft^2 .

Runs with Isobutane and East Mesa Well 6-2 Brine

Figure 5-1 shows the process flow diagram for East Mesa operation. A plot plan is presented in Figure G-1. A 72-hour run ES-1 was made using the packed column section to observe scaling characteristics of continuous operation. Run ES-2 was a similar 160-hour scale deposition test using the sieve tray section. Following these continuous operation tests a series

Figure G-1. Plot plan for the DCHE test unit installation at the USBR East Mesa site.

of sixteen short parametric runs were made using the sieve tray section to obtain performance data at different isobutane/brine flow rates, varying inlet temperature, and at a flooding point.

Packed Column Section

Run ES-1 was the first test of the direct contact heat exchange unit with geothermal brine. This test was conducted at East Mesa Geothermal Well 6-2 from January 19 to January 29, 1977. The objective was to observe operation of the unit with the Goodloe stainless steel mesh packed heat exchange column, using geothermal brine.

Scale buildup was the main concern of this run, but very little scale formed in the unit during the test. Most of the scale that did form was iron oxide scale caused by oxygen leaking into the unit. Only a trace of carbonate silicate scale was found.

Run ES-1 consisted of four separate operating periods. Three of these periods were short and were not a significant part of Run ES-1 except that some scale deposition may have occurred during these times. A fourth 72-hour test was made where smooth operation was achieved. During the 72-hour test, the column was operated at 548 psig with a flow rate of 1070 lbs/hr isobutane and 1830 lbs/hr brine. Brine and isobutane temperatures in and out of the column are plotted in Figure G-2. Input brine temperatures ranged from 330°F to 340°F with brine exit temperatures ranging from 205°F to 215°F. Isobutane input temperatures ranged from 95°F to 102°F. Isobutane exit temperatures ranged from 287°F to 298°F.

Brine was supplied to the unit from Well 6-2 as diagrammed in Figure G-1. The total well flow rate was approximately 90 gpm. Well supply pressure was 141 psig for the first 48 hours of the 72-hour test and then dropped to 132 psig. Well supply temperature was 330°F for 51 hours and then rose to 340°F. There was a 7°F brine temperature drop from where well supply temperature was measured to the test unit.

Figure G-2. Inlet and outlet packed column temperatures for test run ES-1.

At the unit, brine was pumped to 800 psig. From this 800 psig pressure head, 1830 lbs/hr of brine was fed, through manually controlled valve FC-1, to the top of the heat exchange column operating at 550 psig. The brine level in the column was controlled automatically at 9 in. below the top of the packing.

Isobutane was supplied to the bottom of the column from the accumulator. Isobutane was pumped to between 570 and 660 psig during the test. A pressure of 610 psig was necessary to maintain the 1070 lbs/hr isobutane flow into the column. The isobutane pump was capable to increasing pressure by 500 psi. Therefore, it was necessary to have at least 110 psig pressure in the accumulator. Nitrogen was supplied to the accumulator to maintain the pressure.

Operating Problems - 72 Hour Portion of Run

Shut-off. During ES-1 the brine supply was shut off accidentally. This occurred 29 hours into the test. The brine flow was reestablished and the unit put back on stream. Downtime was around one hour.

Controls. Automatic control of water level in the accumulator was troublesome. During the day isobutane in the dP cell LT-2 leg was heated above the accumulator temperature by the sun. Consequently vapor formed in these lines and LCV-2 opened draining not only water but isobutane. To prevent this, LCV-2 was manually operated during the day.

Sieve Tray Section

Run ES-2 was a scale deposition test of the direct contact heat exchange sieve tray unit. The test was conducted from 15:45, February 11, to 8:30, February 18, 1977. The objective of this test was to observe scaling characteristics during continuous operation of the sieve tray heat exchange column using geothermal brine. Only traces of scale were found after the test. By visual inspection the unit was cleaner after the test than before the test was begun.

Run ES-2 lasted 160 hours. The column was operated at 623 psig. Brine and isobutane flows were varied somewhat during the run. These flows are plotted in Figure G-3. Brine flow was 2345 lbs/hr for most of the run. Isobutane flow was 2000 lbs/hr $\pm$ 100 lbs/hr throughout the run. Brine and isobutane temperatures in and out of the column are plotted in Figure G-4. Input and output brine temperatures averaged 100°F and 287°F, respectively. The total flowrate from Well 6-2 was approximately 90 gpm. Well supply pressure was 138 to 145 psig with supply temperature between 332 and 336°F. There was a 7°F temperature drop from the well supply line to the DCHE unit.

At the unit, brine was pumped to 800 psig, and brine was fed to the top of the column through manually set valve FC-1. Brine flowed down the column over 12 sieve trays. The brine level was controlled at the midpoint between trays 2 and 3 by automatically controlling brine flow out of the column through LC-1.

Isobutane to the bottom of the column from the accumulator was pumped to between 670 and 730 psig during the test. Nitrogen was used to maintain accumulator pressure between 110 and 130 psig.

Operating problems. The run was started at 15:45, February 11, 1977. Five hours into the run the brine pump was cavitating very badly and the unit was shut down for 5 minutes. Upon restarting cavitation continued, although not as badly.

Fourteen hours into the run the unit was shutdown because cooling water pressure was lost. No damage was done, but the unit was down for one hour.

Sixteen hours into the run a small gas bleed was installed at the 6-2 well head. This was sufficient to stop the brine pump cavitation.

The test ran with no further problems for a total run time of 160 hours.

Figure G-3. Brine and isobutane flowrates for test run ES-2.

Figure G-4. Inlet and outlet sieve tray column temperatures for test run ES-2.

Appendix H

Experimental Data

Thermal Data

The data presented in the following tables summarize all the steady state runs made using the DCHE packed column and sieve tray column. The steady state duration of each run used for averaging and the number of temperature readings taken during this period for the average are given in the tables. The column temperatures reported are these averaged values.

Although the maximum deviation of the readings from the average temperatures reported are frequently as high as 3°F, with occasional maximum deviations for a run as great as 8°F, the average deviation of the temperature readings is less than 2°F, and the temperatures cycle with a period of 5 to 20 minutes when plotted against time, so that a steady state cycling of $\pm 2^\circ\text{F}$ occurs.

The column pressure was automatically controlled at the constant value shown on the tables. No pressure variation from operation of the controller was observed during the runs. Flow control valves were manually set and flowrates generally remained constant with the exception of the East Mesa sieve tray parametric runs EP - 8, 9, and 10, when the brine pump was cavitating thus causing some fluctuation and pulsing of the brine flow. Preliminary Runs with Isobutane and Water. A series of runs were made at LaVerne from August through November, 1976, to test the system, calibrate the instruments, and obtain performance and flooding data for the DCHE using isobutane and city water. Figure 5-3 shows the process flow for the apparatus while operating at LaVerne. The calibration data is treated in Appendix D, and the performance data obtained during this period is summarized in Tables H-1 and H-2.

Runs with Isobutane and East Mesa Well 6-2 Brine. The process flow for the East Mesa runs is shown in Figure 5-1. Packed column data is presented on Tables H-3 and H-4. Table H-3 summarizes performance data obtained during four start up runs over a

Table H-1

Packed Column Preliminary Runs

Using Isobutane and Water

Date:	<u>9/21/76</u>	<u>10/19/76</u>	<u>10/20/76</u>	<u>10/25/76</u>	<u>10/26/76</u>		
(Computer Run No.)	(1)	(2)	(3)	(4)	(5)	(6)	
<u>Temperatures (°F)</u>							
TI	7	320.1	186.5	315.9	241.5	319.1	300.0
	8	273.1	164.7	272.9	217.3	268.0	280.5
	9	275.5	166.3	275.2	218.7	271.1	282.6
	18	275.0	157.8	277.2	215.7	268.5	282.3 ⁽¹⁾
	19	238.3	125.4	219.0	193.0	212.1 ⁽¹⁾	266.5
	20	230.1	124.8 ⁽¹⁾	210.8 ⁽¹⁾	187.5 ⁽¹⁾	202.8	260.2
	21	193.5	119.4	179.7	154.5 ⁽¹⁾	172.7	246.4
	22	182.2	121.1	167.4	156.6	167.8	226.4
	23	180.8	121.7	167.5	157.1	167.2	220.0
	24	138.3	117.5	126.3	128.3	130.1	129.2
Length of Run After Steady State (Min.)	23	15	25	34	15	20	
No. of Readings	3	3	6	6	2	3	
(1) Max. Devia- tion (°F)	8.0	1.5	7.3	8.2	3.0	4.2	
<u>Flow Rates</u>							
IB Meter	90	230*	64	230**	70	30	
Lb/Hr	1180	1910	1010	1910	1050	680	
Brine Meter	190	530*	160	500**	180	290	
Lb/Hr	965	1690	875	1635	930	1210	
<u>Col. Pressure</u>							
PC-1 Pot Setting	770	800	820	580	820	820	
PSIA	595	605	615	515	615	615	

* Flooded after 15 minutes - water carryover in isobutane overhead. Water = 50 LV%

** Flooded after 25 minutes - iC₄ carried out column bottom. IC₄ = 65 LV%.

Table H-2

Sieve Tray Preliminary Runs Using Isobutane and Water

Date:	<u>11/15/76</u>	<u>11/8/76</u>
(Computer Run No.)	(15)	(16)
<u>Temperatures (°F)</u>		
TI	7	234.1
	8	223.5
	9	226.6
	10	227.5
	11	220.3
	12	219.9
	13	216.8
	14	212.5 (1)
	15	209.4
	16	204.7
	17	200.8
	18	195.6
	19	191.1 (1)
	20	186.2
	21	177.6
	22	167.3
	23	168.0
	24	137.0
Length of Run, after Steady State (min.)	12	10
No. of Readings	2	3
(1) Max. Deviation (°F)	1.0	1.7
<u>Flow Rates</u>		
IB Meter	270	290
Lb/Hr	2070	2150
Brine Meter	830	995
Lb/Hr	2145	2325
<u>Col. Pressure</u>		
PC-1 Setting	380	620
PSIA	360	530

Table H-3

Packed Column Preliminary East Mesa Runs

(Start-Up Runs)

Date:	<u>1/19/77</u>			<u>1/20/77</u>
(Computer Run No.)	(7)	(8)	(9)	(10)
<u>Temperatures (°F)</u>				
TI 7	336.0	335.0	337.7	339.1
8	296.0	301.0	290.6	290.5
9	301.0	302.0	292.1	292.6
18	303.0	311.7 ⁽¹⁾	283.3	312.8
19	298.7	298.5	249.7 ⁽¹⁾	292.5
20	288.1 ⁽¹⁾	286.0	237.5	277.6 ⁽¹⁾
21	280.5 ⁽¹⁾	280.0	171.5	205.8 ⁽¹⁾
22	228.6	230.6	185.5	207.9
23	227.5	229.7	185.3	207.4
24	87.1	86.4	79.5	81.8
Length of Run				
After Steady				
State (Min.)	15	15	30	15
No. of Readings	4	3	6	2
(1) Max. Deviation (°F)	4.0	2.3	1.5	10
<u>Flowrates</u>				
IB Meter	55	55	139	135
Lb/Hr	970	970	1540	1520
Brine Meter	620*	620*	1010*	550**
Lb/Hr	1840	1840	2345	2585
<u>Col. Pressure</u>				
PC-1 Pot Setting	865	783	660	682
PSIA	632	600	548	447

* $\beta = .522$

** $\beta = .625$

TABLE H-4

Packed Column Scaling Run, ES-1

Time From Start of Run

(Hours)		<u>19-23</u>	<u>26-28</u>	<u>32-48</u>	<u>49-72</u>
(Computer Run No.)		(11)	(12)	(13)	(14)
<u>Temperatures (°F)</u>					
TI	7	330.6	330.5	333.1	340.3
	8	290.9	284.8	288.3	292.7
	9	293.3	287.3	290.7	295.0
	18	306.8	303.7 (1)	307.5	312.8
	19	287.8 (1)	283.2	293.5 (1)	296.5
	20	276.0	271.4	280.1	283.6 (1)
	21	253.2	248.5	252.6	256.0
	22	208.2	204.7	207.7	206.9
	23	207.3	203.7	206.8	205.8
	24	94.1	100.9	96.0	93.0

Length of Run

after Steady

State (Min)

No. of Readings

(1) Max. Deviation (°F)

120	100	480	660
3	3	9	15
2.2	1.9	2.0	2.3

Flow Rates

IB Meter

Lb/Hr

56	70	70	68
965	1075	1075	1065
275	275	285	275
1830	1830	1865	1825

Brine Meter

Lb/Hr

Col. Pressure

PC-1 Pot Setting

PSIA

700	700	700	700
565	565	565	565

two day period. Typical steady state performance data from the 76 hour scaling test run ES-1 are summarized in Table H-4. Run ES-1 temperature data are plotted in Figure H-1.

Table H-5 presents data for the 7 day scaling test run ES-2 using the sieve tray column. Figure H-2 is a plot of these data. Following the scaling test, sixteen runs were made under varying conditions of flow rate, temperature, and column pressure. These data are summarized in Table H-6.

Figure H-1. Packed column temperature profile for test run ES-1.

TABLE H-5

Sieve Tray Scaling Run, ES-2

Time From Start of Run (Hour)	<u>0-18</u>	<u>20-160</u>
(Computer Run No.)	(17)	(18)
<u>Temperature (°F)</u>		
TI 7	333.0	339.5
8	285.5	286.1
9	286.8	287.4
10	300.5	300.0
11	277.5	275.5
12	273.4	272.2
13	268.0 (1)	261.8
14	260.3	254.0
15	255.1	247.9
16	245.9	240.6
17	238.3	231.9 (1)
18	228.8	221.7
19	218.9	210.3
20	206.2 (1)	196.2
21	189.2	179.1 (1)
22	167.7	168.9
23	167.9	169.1
24	100.0	102.5
Length of Run After Steady State (Hours)	7	50
No. Of Readings	8	23
(1) Max. Deviation (°F)	3.8	5.8
<u>Flow Rates</u>		
IB Meter	170	150
Lb/Hr	2165	2035
Brine Meter	390	210
Lb/Hr	3200	2345
<u>Col. Pressure</u>		
PC-1 Pot Setting	880	880
PSIA	638	638

Figure H-2. Sieve tray temperature profile for test run ES-2.

Table H-6

Sieve Tray Parametric Runs, EP-1 through EP-15

(Computer Run No.) Exp. Parametric Run No.		(19) EP-1	(20) EP-1a	(21) EP-2	(22) EP-3	(23) EP-4	(24) EP-5
<u>Temperatures (°F)</u>							
TI	7	319.8	323.0	328.6	338.7	340.2	336.1
	8	280.8	286.2	283.0	304.6	284.9	287.2
	9	281.8	287.5	284.4	307.0	285.9	288.1
	10	292.2	298.3	298.4	315.2	301.8	300.4
	11	274.1	283.0	273.1	294.7	272.0	275.8
	12	255.0	265.1 ⁽¹⁾	260.7 ⁽¹⁾	283.8	269.4	273.0
	13	268.5	273.1	265.3	282.7	262.4	270.1
	14	257.0	267.7	260.1	275.0	252.5	261.8 ⁽¹⁾
	15	251.5	263.8	251.8	271.8	245.1	255.9
	16	243.9	257.4	243.5	267.1	236.8 ⁽¹⁾	248.2
	17	235.8	251.2	235.8	261.8	226.5	240.6
	18	225.7 ⁽¹⁾	242.7	225.3	254.2	213.8	229.7
	19	215.2 ⁽¹⁾	234.2	215.7	247.6	202.9	218.0
	20	202.5	225.1	203.4	238.1 ⁽¹⁾	191.8	204.4
	21	184.1	207.5	188.7	220.1	172.6	185.1
	22	162.1	184.5	161.6	190.2	151.5	159.1
	23	162.3	184.6	161.8	190.2	151.6	159.4
	24	93.3	114.7	98.5	94.7	100.0	93.8
Length of Run After Steady State (Min.)		240	45	25	30	10	20
No. of Readings		10	5	4	3	3	3
(1) Max. Deviation (°F)		3.8	3.7	3.4	3.6	2.6	3.0
<u>Flow Rates</u>							
IB Meter		235	235	340	280	480	490
lb/Hr		2550	2520	3070	2795	3640	3700
Brine Meter		390	390	390	500	380	500
lb/Hr		3210	3210	3210	3620	3160	3620
<u>Col. Pressure</u>							
PC-1 Pot Setting		910	910	910	910	910	910
PSIA		650	650	650	650	650	650

Table H-6 (Continued)

Sieve Tray Parametric Runs, EP-1 through EP-15

(Computer Run No.) Exp. Parametric Run No.		(25) EP-6	(26) EP-7	(27) EP-8**	(28) EP-9	(29) EP-10	(30) EP-11
<u>Temperatures (°F)</u>							
TI	7	336.5	334.2	335.9	341.3	340.8	334.0
	8	301.9	299.7	287.2	284.5	281.8	267.4
	9	304.2	302.2	288.1	286.0	283.0	268.5
	10	305.7	302.4	297.6	299.7	292.8	287.5
	11	285.0	284.1	276.3	271.2	265.2	258.0
	12	280.1	276.5	272.8	267.8	259.2	253.2
	13	281.4	281.6	270.1	262.9	253.3	250.9
	14	274.9	275.7	261.6	253.7	241.5	241.8
	15	271.7	272.1	255.1	245.6	230.1	234.7
	16	267.1	267.6	249.2	236.3 ⁽¹⁾	219.2	225.6
	17	260.8	261.9	240.8	226.6	207.5 ⁽¹⁾	214.8
	18	253.6	254.2 ⁽¹⁾	231.0 ⁽¹⁾	214.8	194.0 ⁽¹⁾	201.7 ⁽¹⁾
	19	245.2	248.0	220.1	202.7	181.8	190.4
	20	234.8 ⁽¹⁾	237.9	206.7	190.7	168.5	178.5
	21	218.9	221.6	185.4	171.6	152.2	160.3
	22	186.7	186.2	159.4	152.5	138.8	138.5
	23	186.8	186.4	159.6	152.8	139.1	138.6
	24	98.5	91.4	92.1	102.5	101.4	85.6
Length of Run After Steady State (Min.)		20	40	60	40	30	30
No. of Readings		3	5	6	4	5	3
(1) Max. Deviation (°F)		2.5	3.1	8.1	2.9	6.0	1.3
<u>Flow Rates</u>							
IB Meter		375	340	600	600	670	510
Lb/Hr		3220	2090	4100	4075	4305	3800
Brine Meter		670	600	600*	640*	670*	520
Lb/Hr		4195	3970	3970	4095	4195	3695
<u>Col. Pressure</u>							
PC-1 Pot Setting		910	910	910	910	910	665
PSIA		650	650	650	650	650	550

* Brine pump cavitating

** Brine orifice taps plugged - cleaned out after EP-8.

Table H-6 (Continued)

Sieve Tray Parametric Runs, EP-1 through EP-15

(Computer Run No.)	(31)	(32)	(33)	(34)
Exp. Parametric				
Run No.	EP-12	EP-13	EP-14	EP-15

Temperatures (°F)

TI	7	337.4	339.7	339.5	339.9
	8	277.1	270.2	299.8	304.0
	9	280.6	273.3	303.4	307.9
	10	282.7	273.1	307.9	315.5
	11	264.8	253.5	272.4 ⁽¹⁾	295.1
	12	258.1	250.0	261.6	288.3
	13	257.5	243.5	261.5	286.1
	14	251.1	235.7	256.8	278.3
	15	245.2	229.0	254.6	273.8
	16	240.0	222.8 ⁽¹⁾	253.1	268.7
	17	232.0 ⁽¹⁾	214.2	247.9	259.5 ⁽¹⁾
	18	223.3	204.0	243.3	250.1 ⁽¹⁾
	19	213.0	193.7	237.4	238.2
	20	202.1	184.3	230.1	227.5
	21	183.3	167.5	215.3	212.0
	22	156.1	145.0	184.8	189.5
	23	155.8	145.1	185.0	189.9
	24	86.3	94.7	97.2	96.4

Length of Run
After Steady
State (Min.)

20	60	40	10
----	----	----	----

No. of Readings

6	7	5	3
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(1) Max. Deviation
(°F)

2.1	7.6	4.7	4.6
-----	-----	-----	-----

Flow Rates

IB Meter	340	365	215	340
Lb/Hr	3105	3190	2440	3080

Brine Meter	635	640	645	850*
Lb/Hr	4085	4095	4110	4720

Col. Pressure

PC-1 Pot Setting	665	545	545	910
PSIA	550	500	500	650

* Flooded after 22 minutes. IC₄ carried out with column bottoms.

Isobutane Carryunder.

Loss of isobutane in the outlet brine is determined from the loss of inventory in the Accumulator V-1 over a 5-day period during the scaling test ES-2 on the sieve plate column. Figure H-3 is the vessel outline drawing for V-1. Figures H-4 through H-7 show the Accumulator isobutane level meter readings (LI-1) vs. time. Over the 5-day period, the isobutane level in the Accumulator dropped at a constant rate of 1.3%/hr.

The DP cell transmitter (LT-3) for LI-1 is calibrated for 0 to 30 in. of water and the low pressure leg of the DP cell is 46 inches of 75°F isobutane. Thus when V-2 is full of isobutane

$$\begin{array}{l} \text{(LI-1)} = 0.0\% \\ \text{Full} \end{array} \quad \text{(H-1)}$$

and when empty

$$\begin{array}{l} \text{(LI-1)} = 84.7\% \\ \text{Empty} \end{array} \quad \text{(H-2)}$$

From the dimensions of V-01, the volume is 2.91 gal/inch. Combining with equation H-2 the volumetric meter reading is 1.58 gal/%.

Since the isobutane level dropped 1.3%/hr, the isobutane losses are 2.06 gal/hr, and

$$m_D = 9.51 \text{ lb/hr} \quad \text{(H-3)}$$

The brine rate m_W for run ES-2 was 2345 lb/hr. Therefore, isobutane carryover in the outlet brine can be expressed as

$$m_D/m_W = 0.0041 \text{ weight fraction} \quad \text{(H-4)}$$

or on a volume basis as

$$(m_D/m_W) (\rho_W/\rho_B) = 0.0081 \text{ LV fraction} \quad \text{(H-4a)}$$

Figure H-3. Vessel outline drawing for the Isobutane Accumulator V-01.

Figure H-4. Isobutane loss during test run ES-2, meter reading vs. time of day.

Figure H-5. Isobutane loss during test run ES-2, meter reading vs. time of day.

Figure H-6. Isobutane loss during test run ES-2, meter reading vs. time of day.

Figure H-7. Isobutane loss during test run ES-2, meter reading vs. time of day.

and on a mole basis as

$$(m_D/m_W) (M_W/M_B) = 0.00125 \text{ mole fraction} \quad (\text{H-4b})$$

Sampling data for the outlet brine during the ES-1 run is given on Table H-7. The volume of isobutane flashed from the brine sample is masked by the air present in the sample bomb and by the evolution of CO_2 or other volatiles as the brine is flashed to atmospheric pressure. The ES-2 data given on Table H-8 shows a 3-6 LV% loss on flashing, whereas the actual isobutane losses, given by Eq. H-4a, are 0.8 LV%. The data for the ES-1 run, Table H-7, also shows 3-6 LV% loss of volatiles.

Because of this interference of other volatile material in the brine samples or in the sample bomb the sampling data must be considered inconclusive for estimating isobutane losses and the isobutane losses represented by Eq. H-4 are used throughout.

Isobutane Solubility in Water. At 175°F and 200 psia the solubility of isobutane in water is 0.000075 mole fraction (API Technical Data Book). No data is available for higher pressures, but generally the pressure effect is small for a liquid-liquid system (API). Data of isobutane vapor solubility in water (Reed and McKetta, 1959) up to 320 psia and 220°F shows an increase in solubility of 2.7×10^{-7} mole fraction per psi. Applying this rate of increase to the above API data gives a solubility of isobutane in water at 175°F and 650 psia of 0.0002 mole fraction. From the correlation of Boehm (1977) the predicted solubility of isobutane in water at 175°F, 650 psia is 0.00023 mole fraction.

Table H-7

Outlet Brine Sample Data

for Run ES-1

<u>Date</u>	<u>Time</u>	<u>Volatiles Flashed</u>		<u>T(23), °F</u>	<u>Brine date, lb/hr</u>
		<u>cc</u>	<u>LV%</u>		
1/26	11:50	9	5.2	220	1830
	14:00	6	3.4	215	1830
	20:40	12	6.9	213	1830
	23:30	10	5.7	209	1815
1/27	7:00	12	6.9	207	1830
	20:20	8	4.6	209	1860
1/28	7:30	6	3.4	207	1860
	14:50	12	6.9	207	1830
1/29	6:30	6	3.4	206	1830
1/20	13:30	15	8.5	300	2590

Table H-8

Outlet Brine Sample Data for Run ES-2

<u>Date</u>	<u>Time</u>	<u>Volatiles Flashed</u>		<u>T(23), °F</u>	<u>Brine Rate, lb/hr</u>
		<u>cc</u>	<u>LV%</u>		
2/11	18:00	12	6.9	172	3200
	21:30	12	6.9	170	3160
2/12	5:20	5	2.9	169	3200
	18:10	9	5.2	164	2290
	21:00	9	5.2	168	2290
2/13	6:00	10	5.7	169	2190
	19:00	9	5.2	172	2290
2/14	18:00	7	4.0	173	2350
2/15	3:15	8	4.6	173	2405
	20:00	6	3.4	163	2350
2/16	5:00	5	2.9	157	2350
2/17	5:00	5	2.9	160	2350
2/18	5:10	5	2.9	158	2350

These estimates compared to the experimental losses of 0.00125 mole fraction given by Equation H-4b indicate that there is a small loss of isobutane by entrainment in the brine. Consequently in the analysis of the data, Equation H-4 losses are used to represent isobutane present in the outlet brine for the apparatus.

Water Carryover.

Sample data for water carryover is summarized on Tables H-9 and H-10 for the scaling test runs ES-1 and ES-2 respectively.

The equilibrium water content of the overhead isobutane stream can be calculated from the vapor pressure of water at 285°F and 650 psia

$$\begin{aligned} y_w &= 53.25 \text{ psia}/650 \text{ psia} \\ &= 0.082 \end{aligned} \quad (\text{H-5})$$

Using data from the NGSMA Engineering Data Book, water content of the gas at 285°F and 650 psia is 3950 lb/10⁶ std. cu. ft. Then, from this value

$$y_w = 0.083 \quad (\text{H-6})$$

From the generalized fugacity correlation of Li and McKetta (1963)

$$y_w = 0.063 \quad (\text{H-7})$$

As shown in Table H-10, data at this temperature and pressure are in this order of magnitude.

Table H-9

Water in Overhead Samples for ES-1

<u>Date</u>	<u>Time</u>	<u>ccH₂O</u>	<u>TI-9, °F</u>	<u>IB Rate LB/HR</u>	<u>Mole Fract. H₂O</u>
1/26	11:50	1	292	895	0.098
	20:45	4	296	995	0.38
	23:40	3	295	965	0.292
1/27	7:10	1	293	965	0.08
	20:30	2	293	1075	0.195
1/28	7:40	1	291	1085	0.08
	15:00	4	294	1065	0.38
1/29	6:40	1	294	1065	0.098

Table H-10

<u>Water in Overhead Samples for ES-2</u>						
<u>Date</u>	<u>Time</u>	<u>cc H₂O</u>	<u>TI-9, °F</u>	<u>IB Rate</u>		<u>Mole fract. H₂O</u>
				<u>Meter</u>	<u>lb/hr</u>	
2/11	21:30	2	285	170	2,165	0.195
2/12	18:15	3	285	130	1,895	0.292
	21:05	3	285	130	1,895	0.292
2/13	6:50	1	285	110	1,743	0.098
	19:00	2.5	286	130	1,895	0.244
2/14	18:10	3	288	140	1,968	0.292
2/15	3:30	1.5	288	120	1,822	0.171
	20:10	3.5	285	160	2,104	0.342
2/16	5:15	1	284	160	2,104	0.098
2/17	5:10	3	286	150	2,035	0.292
2/18	5:15	3	285	150	2,035	0.292

Sample of Internal Fluid Mixture

During the preliminary packed column tests at La Verne using isobutane and water, several samples were taken from near the mid point of the packing (Figure 5-5). Isobutane hold-up during non-flooding operation measured from 3 LV% to 35 LV%. The average sample value was 15 LV% isobutane in the liquid-liquid mixture at 170°F near the mid point of the packing.

At East Mesa, samples were taken of the fluid under the downcomer between trays 7 and 8 of the sieve tray column during test run ES-2. The samples were taken over several days and ranged from 10 LV% isobutane to 26 LV% with an average value of 15 LV% isobutane at 220°F. These sample data are summarized in Table H-11.

Column Pressure Drop

Column pressure drop over the 76-hour run ES-1 remained constant. Pressure drop across the column was 1.0 psi, measured by DP-2. Pressure drop across the bottom distributor plate was 0.4 psi measured by DP-1.

During run ES-2, the column pressure drop measured by DP-2 was constant at 1.1 psi. The pressure drop across tray 6 measured by DP-1 remained constant throughout the run at 0.14 psi.

Scaling

Run ES-1. Upon shutdown, after 76 hours of operation, an examination of the unit was made for scaling. Very little scale was observed and there was no scale in the column itself. Even the largest amount of scale formed on the wall of the column above the packing was miniscule. This scale was black-brown in color and formed flakes about 1/16" thick. There appeared to be several grams of this scale formed. This scale appeared to flake off and fall into the distribution tray on top of packing. No scale formed on the stainless steel packing.

Table H-11

Internal Liquid Samples

I. Packed Column at La Verne

Date:	<u>10/21</u>		<u>10/25</u>	
ml. volatiles	67	35	11	10
LV% volatiles	38.2	20.2	6.3	5.8
Est. air, LV%	2.8	2.8	2.8	2.8
Est. IC ₄ , LV%	35.4	17.4	3.5	3.0
Temp., °F	170	168	183	188
Flow Rates, lb/hr				
Brine	1,105		1,330	
Isobutane	1,298		1,690	

Average IC₄ = 15 LV%

II. Sieve Tray Column at East Mesa, Run ES-2

Date:	<u>2/12</u>		<u>2/13</u>		<u>2/15</u>		<u>2/16</u>	
ml. volatiles	27	30	31	39	17	17	45	45
LV% volatiles	15.4	17.3	17.7	22.5	9.8	9.7	26.0	25.7
Est. air, LV%	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0
Est. IC ₄ , LV%	12.4	14.3	14.7	19.5	6.8	6.7	23.0	22.7
Temp., °F	212	210	218	220	222	220	204	203
Flow rates, lb/hr								
Brine	2,310		2,310		2,420		2,345	
Isobutane	2,050		2,050		2,010		2,035	

Average IC₄ = 15 LV%

Scale also formed on the weirs of the distribution tray. This scale was greenish-gray in color, very thin (0.01") and delicate. Not much scale formed here. Some scale also formed on the bottom distribution tray similar to scale formed on the top but with a more greenish tinge and harder to scrape off.

There was a thin black non-flakey greenish film on LC-1. This scale was much less than a gram in weight. The dump line valve was disassembled and inspected. The packing was clean except for some black on the valve seat head. This black was observed before the start of ES1, so no scale formed here. The expansion chamber following the dump line valve had a greenish thin film above the water filled part.

Scale formed during the first six hours of operation (the first three startups) may have been redissolved, because 1/10th Molar HCl was injected into the brine at something less than 10 gal/day for 5 hours of the third portion. This was done to clean the brine bypass valves at the inlet which had scaled because of the variable operation and/or frequent startups. This injection cleaned the valves and solved the operating problems. However during the final fourth 72-hour portion of the test nothing was added to the brine. Therefore the observed scale buildup represents at least 72 hours of operation but not more than 76 hours.

Samples of the scale buildup on the column wall, on top the stainless steel packing and on the top distributor plate were analyzed for calcium, carbonate, silicate and total iron. Results of these analyses are listed in Table H-12. Scale formed on the column wall and that found on top of the column packing are very similar in analysis. This scale is

Table H-12

Composition of Scale Formed in Run ES-1

<u>Location of scale Sampled</u>	<u>Ca, %</u>	<u>CO₃⁻², %</u>	<u>SiO₂, %</u>	<u>Fe, Total</u>
From column wall (top part of column)	2.02	5.33	2.9	65.8
Distribution weir	21.8	36.3	20.0	10.6
Top of stainless steel packing	1.26	2.5	2.3	62.0
From column wall (bottom part of column)	0.13	7.88	3.6	59.4

very low in calcium carbonate and silicates but high in iron. This iron was in the oxide form and was caused by some oxygen contamination in the brine during the initial startups and shutdowns.

Scale formed on the top distributor plate was calcium carbonate and silicate scale, which was expected from the brine. Only a negligible amount of this scale was formed in the unit.

Run ES-2. Upon shutdown, the unit was inspected for scale. Only traces of scale were found. Scale that did form in the column formed in gray black spots that wiped off with fingers. The rust that was on the column at the start was gone. The brine and butane overhead and column bonnet were the same as after ES-1. The dumpline valve had a very thin greenish scale (0.01"). In the whole unit there was not enough scale to get a gram sample.

Appendix I Data Reduction

Consistency of Data.

As discussed in Appendix D.1 corrected temperature measurements are accurate to 0.1 °F. The flow rates should be accurate to 3/4%, but this has not been established experimentally. The accuracy of flowrate measurement can be checked by an overall energy balance on the column.

The overall energy balance is

$$m_W h_{W7} - (m_W - m_S) h_{W23} - m_D h_{B23} = (m_B - m_D) h_{B9} - m_B h_{B24} + m_S h_{S9} \quad (I-1)$$

The flowrate m_S of water overhead in the isobutane outlet stream is proportional to the flowrate of isobutane according to the equation

$$m_{S1} = [M_W p_W / M_B (P - p_W)] m_B \quad (I-2)$$

where p_W is assumed to be equal to the vapor pressure of water at the temperature T_{E9} of the isobutane phase. The flowrate m_D of isobutane in the exhaust brine stream is proportional to the brine flowrate m_W as discussed in Appendix H so that

$$m_D = 0.0041 m_W \quad (I-3)$$

The heat Q_W extracted from the water phase which is given by

$$Q_W = m_W h_{W7} - (m_W - m_S) h_{W23} \quad (I-4)$$

should equal the heat Q_B absorbed by the isobutane phase given by

$$Q_B = (m_B - m_D) h_{B9} - m_B h_{B24} + m_S h_{S9} \quad (I-5)$$

These two heat flows calculated according to the computer enthalpy computations in Appendix J are given in Table I-1. The enthalpies were evaluated as follows:

h_{W7} at T_{E7} and P
 h_{W23} at T_{E23} and P
 h_{B9} at T_{E9} and $P-p_W$
 h_{B24} at T_{E24} and P
 h_{S9} at T_{E9} and p_W

Alternatively one flowrate can be calculated from the other by rearranging Equation I-1. Thus

$$m_W = \frac{(m_B - m_D)h_{B9} - m_B h_{B24} + m_S h_{S9} - m_S h_{W23}}{h_{W7} - h_{W23} + 0.0041(h_{B9} - h_{B23})} \quad (I-6)$$

and

$$m_B = \frac{m_W h_{W7} - m_W h_{W23} + m_D h_{B9} - m_D h_{B23}}{h_{B9} - h_{B24} + (p_W/P)(h_{S9} - h_{W23})} \quad (I-7)$$

The flowrates were calculated from the above equations using the computer program in Appendix J. The results are given together with the experimentally measured flowrates in Table I-2.

Since temperatures are experimentally confirmed, Tables I-1 and I-2 show that the measured flowrates are not consistent. The brine contains dissolved and gaseous carbon dioxide. Since the pump reduces the pressure at the inlet, gas bubbles are likely formed which do not redissolve before passing through the orifice. Consequently, the brine flowrate measurements are considered unreliable. There is no reason to suspect the butane flowrate to be incorrect, consequently subsequent analysis uses the experimentally measured isobutane flowrate and a brine flowrate calculated from an overall energy balance, Equation I-7.

Table I-1
 Energy Balance Consistency Demonstrated by
 Heat Flows Calculated from Experimental
Temperatures Using Equations I-4 and I-5

	Compu ^t er Run No.	Thermal Energy Transferred, Btu/hr	
		Out of Brine, Q_w	Into Isobutane, Q_p
Packed Column	1	135701.	173808.
	2	105216.	58885.
	3	130324.	150916.
	4	132778.	120405.
	5	139921.	133064.
	6	97054.	115277.
	7	196208.	229288.
	8	190983.	238788.
	9	353547.	382621.
	10	336213.	371710.
	11	222644.	227476.
	12	229321.	239145.
	13	232667.	248233.
	14	242515.	255696.
Sieve Plate Column	15	91522.	93726.
	16	147072.	136623.
	17	522117.	420510.
	18	396203.	399450.
	19	501261.	451102.
	20	443201.	457480.
	21	532067.	557175.
	22	537328.	680752.
	23	592603.	677222.
	24	635795.	731322.
	25	627175.	760296.
	26	585344.	730195.
	27	695836.	814579.
	28	765886.	752421.
	29	837907.	760150.
	30	707605.	583659.
	31	733262.	702973.
	32	787744.	710624.
	33	634271.	654883.
	34	704720.	750292.

Table I-2

Flowrates Calculated from Energy Balance and the
Other Flowrates using Equations I-6 and I-7 and
Compared with Experimental Values

	Computer Run No.	Brine Flowrate, lbs/hr		Isobutane Flowrate, lbs/hr	
		<u>Experimental</u>	<u>Eq. I-6</u>	<u>Experimental</u>	<u>Eq. I-7</u>
Packed Column	1	965.0	1241.9	1180.0	648.8
	2	1690.0	944.0	1910.0	2568.6
	3	875.0	1014.2	1010.0	624.7
	4	1635.0	1479.3	1910.0	1356.5
	5	930.0	881.7	1050.0	732.0
	6	1210.0	1439.9	680.0	411.9
	7	1840.0	2153.0	970.0	607.0
	8	1840.0	2309.8	970.0	560.9
	9	2345.0	2532.9	1540.0	1048.7
	10	2585.0	2854.9	1520.0	1014.5
	11	1830.0	1863.0	965.0	691.7
	12	1830.0	1902.9	1075.0	763.7
	13	1865.0	1985.4	1075.0	742.2
	14	1825.0	1919.6	1065.0	737.1
Sieve Plate Column	15	2145.0	2190.6	2070.0	1288.3
	16	2325.0	2153.7	2150.0	1429.3
	17	3200.0	2559.1	2165.0	1989.6
	18	2345.0	2357.5	2035.0	1490.6
	19	3210.0	2875.1	2550.0	2099.5
	20	3210.0	3305.6	2520.0	1777.4
	21	3210.0	3355.2	3070.0	2157.0
	22	3620.0	4600.9	2795.0	1583.7
	23	3160.0	3610.8	3640.0	2336.0
	24	3620.0	4163.5	3700.0	2376.0
	25	4195.0	5093.4	3220.0	1916.8
	26	3970.0	4963.7	3090.0	1803.6
	27	3970.0	4648.8	4100.0	2589.8
	28	4095.0	4010.8	4075.0	3041.7
	29	4195.0	3790.2	4305.0	3480.8
	30	3695.0	3037.0	3800.0	3089.7
	31	4085.0	3900.0	3105.0	2448.9
	32	4095.0	3676.0	3190.0	2676.8
	33	4110.0	4229.7	2440.0	1833.0
	34	4720.0	5014.3	3080.0	2076.7

Calculation of Experimental Tray Efficiency

The tray efficiencies are calculated from the experimental temperature and flowrates by a stepwise solution of Equations 8-1 through 8-4 from stage 1 to the Nth stage. For the sieve plate column $N = 12$ for 100% efficient trays. For the packed column, N is taken equal to 15 to approximate $N = \infty$ corresponding to the differential set of Equations 7-8 and 7-9. Figure I-1 shows the relationship of differential sections to the experimental packed column for the stagewise calculations.

Temperature Corrections. All experimental temperatures are corrected by

$$T_{E,J} = T_{I,J} + T_{CAL,J} \quad (I-8)$$

where $T_{CAL,J}$ is obtained from Tables D-6 and D-7.

As described in Appendix E, the temperature $T_{B,N+1}$ of isobutane entering and the temperature of the brine $T_{W,N}$ leaving the bottom of the column intervals must be corrected for heat exchange in the bottom of the column.

These corrections are

$$T_{W,N} = T_{E23} + Q/m_W c_W \quad (I-9)$$

and

$$T_{B,N+1} = T_{E24} + Q/m_B c_B \quad (I-10)$$

where

$$Q = UA(T_{E22} - T_{E24}) \quad (I-11)$$

The quantity UA is given by Equation E-1 for the packed column and Equation E-2 for the sieve plate column.

Figure I-1. Relationship of differential sections to the experimental packed column for stagewise computer calculations.

Sieve Plate Column. Since the sieve plate column has 12 plates, Equation I-9 becomes

$$T_{W12} = T_{E23} + Q/m_W c_W \quad (I-12)$$

and Equation I-10 becomes

$$T_{B13} = T_{E24} + Q/m_B c_B \quad (I-13)$$

where Q is given by Equation I-11 and UA by Equation E-2.

The experimentally determined temperatures $T_{E,J}$ are identified with the appropriate theoretical temperature symbol in Table I-3.

The flowrate of steam leaving the top plate, as in Equation I-2 above is,

$$m_{S1} = [M_W p_{W1} / M_B (P - p_{W1})] m_B \quad (I-14)$$

Where p_{W1} is the vapor pressure of brine at the temperature T_{B1} of isobutane leaving that plate.

Equations 8-1, 8-2, and 8-4 are rearranged in the form

$$m_{W,i} = m_{W,i-1} - m_{S,i} + m_{S,i+1} \quad (I-15)$$

$$h_{B,i+1} = [m_{W,i} h_{W,i} + m_B h_{B,i} + m_{S,i} h_{S,i} - m_{W,i-1} h_{W,i-1} - m_{S,i+1} h_{S,i+1}] / m_B \quad (I-16)$$

$$m_{S,i+1} = [M_W p_{W,i+1} / M_B (P - p_{W,i+1})] m_B \quad (I-17)$$

and solved simultaneously for $m_{W,i}$, $h_{B,i+1}$ and $m_{S,i+1}$ in the downcomer using the experimental temperature $T_{W,i}$ to determine $h_{W,i}$. The enthalpies are related to the temperature by the equation of state for isobutane and brine. As in Section 7, $p_{W,i+1}$ is vapor pressure of water at the temperature $T_{B,i+1}$ of isobutane leaving that plate and $m_{S,i} = 0$ when the isobutane is liquid. The tray efficiency for each plate is then calculated

Table I-3

Identification of TI's and Theoretical Symbols
for Brine and Isobutane Streams Inside

DCHE Sieve Plate Column

<u>Leaving Brine Temperature</u>		<u>Stage Number</u>	<u>Leaving Isobutane Temperature</u>	
<u>TI Number</u>	<u>Symbol</u>		<u>TI Number</u>	<u>J</u>
7	T_W	Brine entering stage 1.		
11	$T_{W,1}$	1	$T_{B,1}$	9
12	$T_{W,2}$	2	$T_{B,2}$	-
13	$T_{W,3}$	3	$T_{B,3}$	-
14	$T_{W,4}$	4	$T_{B,4}$	-
15	$T_{W,5}$	5	$T_{B,5}$	-
16	$T_{W,6}$	6	$T_{B,6}$	-
17	$T_{W,7}$	7	$T_{B,7}$	-
18	$T_{W,8}$	8	$T_{B,8}$	-
19	$T_{W,9}$	9	$T_{B,9}$	-
20	$T_{W,10}$	10	$T_{B,10}$	-
21	$T_{W,11}$	11	$T_{B,11}$	-
23*	$T_{W,12}$	12	$T_{B,12}$	-
Isobutane entering stage 12.			$T_{B,13}$	24**

* corrected by Equation I-12.

** corrected by Equation I-13.

using Equation 7-3. These computations were executed using the computer program in Appendix J. The results are given in Table I-4.

The computation of the last stage, plate 12, gives the enthalpy $h_{B,13}$ and therefore temperature $T_{B,13}$ of isobutane entering that stage. Since $T_{B,13}$ is also known experimentally using Equation I-13, this provides a check on the calculation. This is equivalent to the overall energy balance used to check consistency of data as described in the previous section.

Packed Column. The packed column is approximated by 13 differential sections as shown in Figure I-1. In addition there are two unpacked sections at either end of the column containing a total of 2 sections. Equation I-9 then becomes

$$T_{W,1} = T_{E,23} + Q/m_W c_W \quad (I-18)$$

and Equation I-10 becomes

$$T_{B,1} = T_{E,24} + Q/m_B c_B \quad (I-19)$$

The sensitive portion of the temperature probes are 0.5 inches long, 1/4 inch in diameter, are in thermal contact with the packing, and are located near the centerline of the packing as shown in Figure I-1. Both the isobutane and brine flow countercurrently over these probes, so that the experimentally measured temperature is the temperature obtained by averaging the temperatures of the two phases in proportion to their volume ratio R . Since at the interface between the i and $i + 1$ section the brine is flowing out of the i^{th} section and the butane is flowing out of the $(i+1)^{th}$ section, the experimental temperature $T_{E,J}$ is related to the temperature of each of the phases at the interface by

$$T_{E,J} = T_{B,i+1} R + T_{W,i} (1-R)/R \quad (I-20)$$

Table I-4

Plate Efficiencies for Sieve Plate Column Based on Measured
Isobutane Flowrate and Calculated Brine Flowrate

Run No.:	15			16		
	<u>Brine</u>	<u>Isobutane</u>		<u>Brine</u>	<u>Isobutane</u>	
Flowrates, lbs/hr:	2241	2070		2197	2150	
<u>Stage No.</u>	<u>Temperatures Leaving Stage°F</u>		<u>Efficiency</u>	<u>Temperatures Leaving Stage°F</u>		<u>Efficiency</u>
Brine Ent'g 1	189.7			233.4		
1	176.4	186.3	1.5928	220.2	226.5	1.3134
2	180.7	167.0	-1.6929	217.8	210.3	.3890
3	179.5	173.4	.2817	215.6	207.2	.3432
4	177.0	171.6	.4812	212.2	204.3	.4632
5	176.0	167.9	.1980	209.2	199.8	.3909
6	174.5	166.4	.2703	204.7	195.8	.5042
7	170.2	164.1	.5866	201.0	189.8	.3970
8	165.3	157.6	.5600	195.7	184.7	.4901
9	161.3	150.0	.4147	191.2	177.2	.3922
10	158.1	143.7	.3078	186.4	170.8	.3816
11	151.1	138.6	.5283	177.9	163.9	.5482
12	146.1	127.2	.3468	169.0	151.3	.5024
Ent'g But. calc'd		118.9			137.7	
Ent'g But. exp'tl		119.1			138.2	

Table I-4 (Continued)

Plate Efficiencies for Sieve Plate Column Based on Measured
Isobutane Flowrate and Calculated Brine Flowrate

Run No.:

17

18

		<u>17</u>			<u>18</u>		
		<u>Brine</u>	<u>Isobutane</u>		<u>Brine</u>	<u>Isobutane</u>	
Flowrates, lbs/hr:		2593	2165		2383	2035	
<u>Stage No.</u>		<u>Temperatures Leaving Stage°F</u>		<u>Efficiency</u>	<u>Temperatures Leaving Stage°F</u>		<u>Efficiency</u>
Brine Ent'g 1		332.3			338.8		
1		277.4	286.7	1.0925	275.4	287.3	1.1036
2		271.3	266.6	.7219	270.1	263.6	.6217
3		266.8	262.6	.6829	260.6	259.5	.9439
4		260.0	258.7	.9152	253.7	249.2	.7529
5		254.9	251.2	.7338	247.7	239.8	.6033
6		245.9	244.4	.9231	240.6	229.8	.5675
7		238.5	230.1	.6377	232.1	219.8	.5812
8		228.9	219.5	.6713	221.8	207.5	.5898
9		219.0	205.1	.5881	210.4	191.8	.5510
10		206.4	189.6	.6006	196.4	173.7	.5525
11		189.5	169.0	.6226	179.4	150.5	.5402
12		169.7	139.8	.5701	171.0	120.7	.2510
Ent'g But. calc'd			103.5			105.4	
Ent'g But. exp'tl			103.6			105.4	

Table I-4 (Continued)

Plate Efficiencies for Sieve Plate Column Based on Measured
Isobutane Flowrate and Calculated Brine Flowrate

Run No.:	19			20		
	<u>Brine</u>	<u>Isobutane</u>		<u>Brine</u>	<u>Isobutane</u>	
Flowrates, lbs/hr:	2909	2550		3347	2520	
<u>Stage No.</u>	<u>Temperatures Leaving Stage °F</u>		<u>Efficiency</u>	<u>Temperatures Leaving Stage °F</u>		<u>Efficiency</u>
Brine Ent'g 1	319.1			322.3		
1	274.0	281.7	1.0933	282.9	287.4	1.0606
2	266.4	262.4	.7933	271.0	270.3	.9726
3	253.8	255.7	1.0797	263.9	263.2	.9556
4	256.7	239.9	-.5292	267.4	256.0	-1.5837
5	251.3	243.9	.5950	263.6	259.9	.6726
6	243.9	236.3	.6610	257.4	255.6	.8732
7	236.0	224.8	.5850	251.4	247.0	.7326
8	225.8	213.7	.6283	242.8	237.3	.7580
9	215.3	198.8	.5607	234.3	221.9	.5788
10	202.7	182.9	.5596	225.3	207.8	.5078
11	184.4	162.9	.6296	207.8	192.3	.6925
12	164.1	132.3	.5604	186.4	160.1	.6188
Ent'g But. calc'd		96.3			117.8	
Ent'g But. exp'tl		96.3			117.9	

Table I-4 (Continued)

Plate Efficiencies for Sieve Plate Column Based on Measured
Isobutane Flowrate and Calculated Brine Flowrate

Run No.:

21

22

		<u>Brine</u>	<u>Isobutane</u>		<u>Brine</u>	<u>Isobutane</u>	
Flowrates, lbs/hr:		3389	3070		4655	2795	
<u>Stage No.</u>		<u>Temperatures Leaving Stage°F</u>		<u>Efficiency</u>	<u>Temperatures Leaving Stage°F</u>		<u>Efficiency</u>
Brine Ent'g 1		327.9			338.0		
1		273.0	284.3	1.1147	294.6	306.9	1.1651
2		263.2	261.1	.9032	281.7	280.3	.9489
3		259.5	251.7	.4868	281.5	272.8	.0439
4		259.8	247.3	-.0505	274.7	272.7	.8718
5		251.6	247.7	.8071	271.6	269.4	.7416
6		243.5	236.8	.7071	267.1	267.4	1.0321
7		236.0	224.7	.5706	262.0	263.4	1.1591
8		225.4	214.6	.6621	254.3	256.9	1.2078
9		215.8	199.7	.5436	247.7	243.3	.7491
10		203.6	185.6	.5759	238.3	227.7	.6401
11		189.0	167.1	.5711	220.4	208.5	.7499
12		163.5	143.9	.7223	192.6	168.5	.6981
Ent'g But. calc'd			100.8			98.8	
Ent'g But. exp'tl			100.8			99.0	

Table I-4 (Continued)

Plate Efficiencies for Sieve Plate Column Based on Measured
Isobutane Flowrate and Calculated Brine Flowrate

Run No.:	23			24		
	<u>Brine</u>	<u>Isobutane</u>		<u>Brine</u>	<u>Isobutane</u>	
Flowrates, lbs/hr:	3642	3640		4203	3700	
<u>Stage No.</u>	<u>Temperatures Leaving Stage°F</u>		<u>Efficiency</u>	<u>Temperatures Leaving Stage°F</u>		<u>Efficiency</u>
Brine Ent'g 1	339.5			335.4		
1	271.9	285.8	1.1146	275.7	288.0	1.1148
2	267.3	259.5	.5399	270.9	264.9	.6170
3	261.2	255.6	.6854	268.9	261.6	.3534
4	252.2	249.5	.8717	261.5	259.9	.9035
5	244.9	239.0	.7132	255.7	252.4	.7796
6	236.8	229.1	.6784	248.2	245.2	.8329
7	226.7	219.4	.7348	240.8	233.5	.6700
8	213.9	206.8	.7834	229.8	223.6	.7799
9	203.0	190.2	.6295	218.1	208.1	.7012
10	192.0	175.4	.5702	204.6	190.8	.6626
11	172.9	160.0	.7480	185.4	169.9	.7130
12	153.0	132.1	.6557	161.1	138.5	.6827
Ent'g But. calc'd		101.5			95.9	
Ent'g But. exp'tl		101.6			96.0	

Table I-4 (Continued)

Plate Efficiencies for Sieve Plate Column Based on Measured
Isobutane Flowrate and Calculated Brine Flowrate

Run No.:

25

26

		<u>Brine</u>	<u>Isobutane</u>		<u>Brine</u>	<u>Isobutane</u>	
Flowrates, lbs/hr:		5155	3220		5022	3090	
<u>Stage No.</u>		<u>Temperatures Leaving Stage °F</u>		<u>Efficiency</u>	<u>Temperatures Leaving Stage °F</u>		<u>Efficiency</u>
Brine Ent'g 1		335.8			333.5		
1		284.9	304.1	1.2324	284.0	302.1	1.2237
2		279.3	274.2	.6892	279.5	273.7	.6085
3		278.9	272.3	.1087	275.3	272.0	.7210
4		274.6	272.2	.7804	275.4	270.1	-.0389
5		271.5	270.2	.8278	271.9	270.1	.7964
6		267.1	268.4	1.1749	267.6	268.0	1.0475
7		261.0	265.1	1.5006	262.1	264.5	1.2850
8		253.7	258.2	1.4499	254.3	258.2	1.3299
9		245.3	246.2	1.0566	248.1	245.0	.8016
10		235.0	228.3	.7549	238.1	230.8	.7321
11		219.2	208.0	.7390	221.9	211.0	.7476
12		189.0	174.4	.8057	188.7	176.1	.8399
Ent'g But. calc'd			102.3			95.3	
Ent'g But. exp'tl			102.4			95.5	

Table I-4 (Continued)

Plate Efficiencies for Sieve Plate Column Based on Measured
Isobutane Flowrate and Calculated Brine Flowrate

Run No.:

27

28

	<u>Brine</u>	<u>Isobutane</u>		<u>Brine</u>	<u>Isobutane</u>	
Flowrates, lbs/hr:	4693	4100		4050	4075	
<u>Stage No.</u>	<u>Temperatures Leaving Stage°F</u>		<u>Efficiency</u>	<u>Temperatures Leaving Stage°F</u>		<u>Efficiency</u>
Brine Ent'g 1	335.2			340.6		
1	276.2	288.0	1.1111	271.1	285.9	1.1192
2	270.7	265.1	.6627	265.7	258.6	.6034
3	268.9	261.2	.3189	261.7	253.9	.5054
4	260.8	259.7	.9353	253.4	249.8	.8238
5	254.9	251.2	.7618	245.4	240.3	.7597
6	249.2	243.6	.6714	236.3	229.6	.7306
7	241.0	234.2	.7075	226.8	218.7	.7024
8	231.1	223.1	.7135	214.9	207.0	.7503
9	220.2	209.1	.6634	202.8	191.6	.6844
10	206.9	193.0	.6566	190.9	175.4	.6058
11	185.7	172.3	.7604	171.9	158.9	.7445
12	161.3	137.4	.6710	154.1	131.3	.6085
Ent'g But. calc'd		94.3			104.1	
Ent'g But. exp'tl		94.3			104.2	

116

Table I-4 (Continued)

Plate Efficiencies for Sieve Plate Column Based on Measured
Isobutane Flowrate and Calculated Brine Flowrate

Run No.:

29

30

Flowrates, lbs/hr:

Stage No.

Brine Ent'g 1

1

2

3

4

5

6

7

8

9

10

11

12

Ent'g But. calc'd

Ent'g But. exp'tl

Brine

Isobutane

3826

4305

Temperatures Leaving Stage°F

Efficiency

Brine

Isobutane

3067

3800

Temperatures Leaving Stage°F

Efficiency

340.1

265.1

257.1

252.1

241.2

229.9

219.2

207.7

194.1

181.9

168.7

152.5

140.1

282.9

252.6

244.9

239.5

227.1

214.9

203.0

189.6

173.3

158.1

141.1

119.6

102.6

1.1346

.7797

.5821

.9296

.8891

.8337

.8291

.8589

.7389

.7135

.7404

.5458

333.3

257.9

251.1

249.7

241.5

234.5

225.6

215.0

201.8

190.5

178.7

160.6

140.0

268.4

219.9

212.5

209.9

201.5

194.2

184.6

172.8

157.8

144.5

130.2

107.5

80.7

1.0748

.3038

.0700

.3417

.2979

.3615

.4107

.4770

.4085

.4081

.5435

.5591

102.6

87.3

Table I-4 (Continued)

Plate Efficiencies for Sieve Plate Column Based on Measured
Isobutane Flowrate and Calculated Brine Flowrate

Run No.:	31			32		
	<u>Brine</u>	<u>Isobutane</u>		<u>Brine</u>	<u>Isobutane</u>	
Flowrates, lbs/hr:	3946	3105		3719	3190	
<u>Stage No.</u>	<u>Temperatures Leaving Stage °F</u>	<u>Efficiency</u>		<u>Temperatures Leaving Stage °F</u>	<u>Efficiency</u>	
Brine Ent'g 1	336.7			339.0		1.1308
1	264.7	280.5	1.1232	253.4	273.2	2.8424
2	256.0	262.9	1.6524	247.9	255.0	2.6794
3	256.3	255.3	-1.3953	242.3	249.3	2.5182
4	250.8	255.6	1.7713	235.4	243.7	2.1604
5	245.0	249.0	1.5203	228.8	235.9	1.6949
6	240.0	240.3	1.0320	222.8	227.7	1.4815
7	232.2	233.5	1.0932	214.4	219.9	1.2553
8	223.4	222.3	.9422	204.1	208.3	.9772
9	213.1	208.9	.8306	193.8	193.3	.7294
10	202.3	192.3	.6838	184.5	177.6	.8698
11	183.6	174.0	.7956	167.8	162.8	.7875
12	157.6	140.3	.7507	146.4	134.9	
Ent'g But. calc'd		89.3			96.8	
Ent'g But. exp'tl		89.4			96.9	

Table I-4 (Continued)

Plate Efficiencies for Sieve Plate Column Based on Measured
Isobutane Flowrate and Calculated Brine Flowrate

Run No.:

33

34

Flowrates, lbs/hr:

Brine

Isobutane

Brine

Isobutane

4289

2440

5081

3080

Stage No.

Temperatures Leaving Stage °F

Efficiency

Temperatures Leaving Stage °F

Efficiency

Brine Ent'g 1

338.8

339.2

1

272.3

303.3

1.3039

295.0

307.8

1.1693

2

259.5

265.5

1.3055

286.2

280.7

.7626

3

260.3

259.0

4.7059

284.9

274.4

.1983

4

256.5

259.7

1.7234

278.0

274.0

.7748

5

254.4

256.1

1.7143

273.6

271.3

.7964

6

253.1

253.7

1.2745

268.7

269.0

1.0305

7

248.1

251.9

1.6155

259.7

265.3

1.4575

8

243.4

242.5

.9109

250.2

254.1

1.2583

9

237.5

234.3

.7867

238.3

235.8

.9063

10

230.3

223.0

.6648

227.7

210.1

.5458

11

215.6

208.2

.7993

212.3

187.2

.5513

12

187.1

175.1

.8251

192.1

151.6

.4987

Ent'g But. calc'd

102.0

101.0

Ent'g But. exp'tl

102.2

101.2

The relationship between the experimentally determined temperatures and the differential sections temperatures is shown in Table I-5.

Since four temperatures inside the column and the outlet brine temperature were measured, the five efficiencies defined in Table I-6 can be determined using Equations 7-1, 7-2, and 7-4.

The calculational procedure is similar to that of the sieve plate column.

The flow rate of steam leaving the top section is given by Equation I-14. Equations 7-1, 7-2, and 7-4 are rearranged the same as for the sieve plate column and become Equations I-15, I-16, and I-17. The brine temperature leaving each section is obtained by rearranging Equation 7-3 for the tray efficiency in the form

$$T_{W,i} = [T_{W,i-1} - E_L (T_{W,i-1} + T_{B,i}/2)] / (1 - E_L/2) \quad (I-21)$$

where L is defined in Table I-6. At sections 1, 5, 9, 13, and 15 Equation I-20 relates the experimental temperature to the interface temperatures $T_{B,i+1}$ and $T_{W,i}$. Thus Equations I-15, I-16, and I-17 for the appropriate sections together with Equation I-20 can be solved simultaneously for E_L .

The computer program in Appendix J, solves these equations by iteration on E_L as follows. The section efficiency E_L is initially taken as 0.9. Then for the next differential section Equation I-21 is solved for $T_{W,i}$. Next Equations I-15, I-16, and I-17 are solved simultaneously. Then $T_{B,i+1}$ is calculated from $h_{B,i+1}$ and the equation of state. This is repeated for each differential section in the segment 1. Then $T_{W,i}$ and $T_{B,i+1}$ from the last differential section is compared with $T_{E,J}$ using Equation I-20. A new value of E_L is estimated and a new temperature comparison made. This is repeated until Equation I-20 is satisfied. Then the whole procedure is repeated for the next segment L .

Table I-5

Temperature Symbols for Brine and Isobutane Phases Leaving a Given Differential Section or Crossing a Given Interface Between Sections of the Packed Column.

The Interface Temperatures Related by Equation I-20 to Given Experimental Temperatures are Identified.

Differential Section Number I	Temperature Symbol Leaving or Crossing		Interface Between Sections	Experimental Temperature TI Number J
	Brine	Isobutane		
-	$T_{W,0}$	Brine Entering Section 1		7
1	-	$T_{B,1}$		9
1	$T_{W,1}$	-		-
-	$T_{W,1}$	$T_{B,2}$	1 & 2	18
2	$T_{W,2}$	$T_{B,2}$		-
3	$T_{W,3}$	$T_{B,3}$		-
4	$T_{W,4}$	$T_{B,4}$		-
5	$T_{W,5}$	$T_{B,5}$		-
-	$T_{W,5}$	$T_{B,6}$	5 & 6	19
6	$T_{W,6}$	$T_{B,6}$		-
7	$T_{W,7}$	$T_{B,7}$		-
8	$T_{W,8}$	$T_{B,8}$		-
9	$T_{W,9}$	$T_{B,9}$		-
-	$T_{W,9}$	$T_{B,10}$	9 & 10	20
10	$T_{W,10}$	$T_{B,10}$		-
11	$T_{W,11}$	$T_{B,11}$		-
12	$T_{W,12}$	$T_{B,12}$		-
13	$T_{W,13}$	$T_{B,13}$		-
-	$T_{W,13}$	$T_{B,14}$	13 & 14	21
14	$T_{W,14}$	$T_{B,14}$		-
15	$T_{W,15}$	-		23*
15	-	$T_{B,15}$		24**

* corrected by Equation I-18.

** corrected by Equation I-19.

Table I-6

Definitions of Packed Column Segment Efficiencies

<u>Segment L</u>	<u>Efficiency Symbol</u>	<u>Sections i</u>
1	E_1	1
2	E_2	2 to 5
3	E_3	6 to 9
4	E_4	10 to 13
5	E_5	14 to 15

Finally, as with the sieve plate column, the calculated value of isobutane temperature $T_{B,16}$ entering the bottom section is compared with the experimental temperature $T_{E,24}$ using Equation I-19.

Appendix J

Data Reduction Computer Programs

The subroutines used in these programs are given in Appendix C.

Consistency (Subroutine Geobal)

```

00160C      READ EXPERIMENTAL TEMPS AND USE CORRECTION
00170C      FACTORS TO CONVERT TO CORRECTED EXPERIMENTAL TEMPS
00180C              ICOLTP = 1 PACKED
00190C              = 2 SIEVE
00200C
00210      PROGRAM GEOBAL(INPUT,OUTPUT,TAPE3,TAPE5,TAPE6)
00220      COMMON /CSOAVE/NOC,JC,CID,TY,PC,COMEGA,API,MOLW,KWAT,
00230+      AC,B,MM,A,FUG,Z,DH,DS
00240      REAL TY(5),PC(5),COMEGA(5),API(5),MOLW(5),KWAT(5),
00250+      AC(5),B(5),MM(5),A(5),FUG(5),Z,DH,DS
00260      REAL MW,MS,MB,MD
00270      INTEGER NOC,CID(5),IC
00280      DATA NOC/1/,PC(1)/529.1/,TY(1)/734.75/,COMEGA(1)/.1848/,
00290+      CID(1)/5/,MOLW(1)/58.124/,JC/1/
00300C
00310      CALL SRKIN
00320C
00330      REWIND 5
00340      REWIND 3
00350      READ(3,)NORUNS
00360      DO 102 I=1,NORUNS
00370      READ(3,)MW,MB,P,WP,ICOLTP
00380      IF(ICOLTP.EQ.2) GO TO 10
00390      READ(3,)TI7,TI8,TI9, TI18,TI19,TI20,TI21,TI22,TI23,TI24
00400      CTI7=TI7-.7
00410      CTI8=TI8+1.5
00420      CTI9=TI9-.1
00430      CTI18=TI18+0.0
00440      CTI19=TI19-0.6
00450      CTI20=TI20+0.3
00460      CTI21=TI21+0.1
00470      CTI22=TI22+0.9
00480      CTI23=TI23+0.2
00490      CTI24=TI24+0.0
00500      GO TO 20
00510      10 READ(3,)TI7,TI8,TI9,TI10,TI11,TI12,TI13,TI14,
00520+      TI15,TI16,TI17,TI18,TI19,TI20,TI21,TI22,TI23,TI24
00530      CTI7=TI7-0.7
00540      CTI8=TI8+1.5
00550      CTI9=TI9-0.1
00560      CTI10=TI10-0.5
00570      CTI11=TI11-0.1
00580      CTI12=TI12-2.1
00590      CTI13=TI13-1.2

```

Consistency (cont'd)

```
00600      CTI14=TI14-0.3
00610      CTI15=TI15-0.2
00620      CTI16=TI16+0.0
00630      CTI17=TI17+0.2
00640      CTI18=TI18+0.1
00650      CTI19=TI19+0.1
00660      CTI20=TI20+0.2
00670      CTI21=TI21+0.3
00680      CTI22=TI22+0.9
00690      CTI23=TI23+0.2
00700      CTI24=TI24+0.0
00710C
00720C      CALCULATE HEAT LOSS FROM WATER
00730      20 CONTINUE
00740C
00750C      CORRECTS FOR HEADER HEAT TRANSFER
00760C
00770      U=50.4*MW/1500.-11.*(MW/1500.）**2+1.7*(MW/1500.）**3
00780      CFB=CFIB(CTI24)
00790      Q=U*(CTI22-CTI24)
00800      CTI24=CTI24+Q/(MB*CFB)
00810      CALL WPROP(1,CTI23,P,0,2,DUM,CF)
00820      CF=CF/18.02
00830      CTI23=CTI23+Q/(MW*CF)
00840      CALL WPROP(0,CTI9,P,0,4,FW9,DUM)
00850      CALL PHASEB(CTI9,P,FW9)
00860      MS=FW9/(P-FW9)*MB*18.02/58.124
00870      MD=.0041*MW
00880      CALL WPROP(1,CTI7,P,P,2,HW7,DUM)
00890      HW7=(HW7*(1.-WP/100.)+.002*WP*CTI7)/18.02
00900      CALL WPROP(1,CTI23,P,P,2,HW23,DUM)
00910      HW23=(HW23*(1.-WP/100.)+.002*WP*CTI23)/18.02
00920      QW=MW*HW7-(MW-MS)*HW23
00930C
00940C      CALCULATE HEAT GAINED BY BUTANE
00950C
00960      CALL WPROP(0,CTI9,P,0,4,FW9,DUM)
00970      PB9=P-FW9
00980      CALL WPROP(0,CTI9,P,FW9,2,HW9,DUM)
00990      HW9=HW9/18.02
01000      CALL ETHB(CTI9,PB9,HB9)
01010      CALL ETHB(CTI23,P,HB23)
01020      CALL ETHB(CTI24,P,HB24)
01030      QB=(MB-MD)*HB9-MB*HB24+MS*HW9
01040C
01050      CMWNUM=(MB-MD)*HB9-MB*HB24+MS*HW9-MS*HW23
01060      CMWDEN=HW7-HW23+.0041*(HB9-HB23)
01070      CMW=CMWNUM/CMWDEN
01080C
```

Consistency (cont'd)

```
01090      CMBNUM=MW*HW7-MW*HW23+MD*HB9-MD*HB23
01100      CMBDEN=HB9-HB24+(PW9/(P-PW9))*(HW9-HW23)
01110      CMB=CMBNUM/CMBDEN
01120      PRINT101,QW,QB,MW,CMW,MB,CMB,CTI7,CTI9,CTI23,CTI24,
01130+      MS,MD,HB24,HB23,HW7,HW9,HW23
01140      WRITE(5,)CMW
01150      102 CONTINUE
01160      101 FORMAT(2F8.0,4F8.1,4F6.1,F8.1,F8.4,5F8.1)
01170      STOP
01180      END
01190      SUBROUTINE PHASEB(T,P,PS)
01200      IPHASE=1
01210      IF(P.EQ.500) GO TO 10
01220      IF(T.LT.274.75) PS=PS*(T/274.74)**40
01230      GO TO 20
01240      10 IF(T.LT.270) PS=PS*(T/270.)**40
01250      20 RETURN
01260      END
01270C****
```

Sieve Plate

05/26/77. 16.35.28.
LIST OF SIVEBAL

```
00100C   THIS PROGRAM IS USED TO ANALYZE DATA FROM THE
00110C   GEOTHERMAL SIEVE PLATE COLUMN HEAT EXCHANGER.
00120C   FRIMODIG 4,77
00130C
00140   PROGRAM SIVEBAL(INPUT,OUTPUT,TAPE3,TAPE5,TAPE6)
00150   COMMON /CSOAVE/NOC,JC,CID,TY,PC,COMEGA,API,MOLW,KWAT,
00160+   AC,B,MM,A,FUG,Z,DH,DS
00170   DIMENSION T(30),FS(30),CPS(30),AMS(30),AMW(30),
00180+   PR(30),HB(30),HW(30),HS(30),CORFAC(16),CPW(35),AT(35)
00190   REAL TY(5),PC(5),COMEGA(5),API(5),MOLW(5),KWAT(5),
00200+   AC(5),B(5),MM(5),A(5),FUG(5),Z,DH,DS
00210   REAL MW,MS,MR
00220   INTEGER NOC,CID(5),IC
00230   DATA NOC/1/,PC(1)/529.1/,TY(1)/734.75/,COMEGA(1)/.1848/,
00240+   CID(1)/5/,MOLW(1)/58.124/,JC/1/
00250C
00260   CALL SRKIN
00270C
00280   REWIND 5
00290   REWIND 3
00300C
00310C   READ EXPERIMENTAL TEMPS AND USE CORRECTION
00320C   FACTORS TO CONVERT TO CORRECTED TEMPS
00330C
00340   READ(3,)NORUNS
00350   DO 50 NEXP=1,NORUNS
00360   READ(3,)AMW(1),MB,P,WP,ICOLTF
00370   READ(3,)AT(1),ATI8,AT(2),ATI10,(AT(I),I=3,23,2),ATI22,AT(25),
00380+   AT(26)
00390   T(1)=AT(1)-0.7
00400   TI8=ATI8+1.5
00410   T(2)=AT(2)-0.1
00420   TI10=ATI10-0.5
00430   T(3)=AT(3)-0.1
00440   T(5)=AT(5)-2.1
00450   T(7)=AT(7)-1.2
00460   T(9)=AT(9)-0.3
00470   T(11)=AT(11)-0.2
00480   T(13)=AT(13)+0.0
00490   T(15)=AT(15)+0.2
00500   T(17)=AT(17)+0.1
00510   T(19)=AT(19)+0.1
00520   T(21)=AT(21)+0.2
00530   T(23)=AT(23)+0.3
00540   TI22=ATI22+0.9
00550   T(25)=AT(25)+0.2
00560   T(26)=AT(26)+0.0
```


Sieve Plate (cont'd)

```

00570C
00580C      CORRECTIONS FOR HEADER HEAT TRANSFER
00590C
00600      U=50.4*AMW(1)/1500.-11.0*(AMW(1)/1500.)*2+1.7*(AMW(1)
00610+      /1500.)*3
00620      CFB=CFIB(T(26))
00630      Q=U*(TI22-T(26))
00640      T(26)=T(26)+Q/(MB*CFB)
00650      CALL WPROP(1,T(25),P,0,2,DUM,CP)
00660      CP=CP/18.02
00670      T(25)=T(25)+Q/(AMW(1)*CP)
00680C
00690      CALL GEOBAL(T(1),T(2),T(25),T(26),AMW(1),MB,P,WF)
00700C
00710C
00720C      PRINT INPUT DATA
00730C
00740      PRINT 181
00750  181  FORMAT(5/)
00760      PRINT,*                      RUN*,NEXP,/
00770      PRINT 107
00780  107  FORMAT(15X,*TI*,9X,*RAW TEMP*,9X,*CORR TEMP*)
00790      PRINT,*                      7*,AT(1),T(1)
00800      PRINT,*                      8*,ATI8,TI8
00810      PRINT,*                      9*,AT(2),T(2)
00820      PRINT,*                      10*,ATI10,TI10
00830      IZ=11
00840      DO 72 IF=3,23,2
00850      PRINT,IZ,AT(IF),T(IF)
00860  72  IZ=IZ+1
00870      PRINT,*                      22*,ATI22,TI22
00880      PRINT,*                      23*,AT(25),T(25)
00890      PRINT,*                      24*,AT(26),T(26)
00900      PRINT,/,*FLOWRATES MB*,MB
00910      PRINT,*                      MW*,EMW
00920      PRINT,*CALC BRINE RATE*,AMW(1)
00930      PRINT,*PRESSURE*,P
00940      PRINT,*UA*,U
00950      PRINT,*Q*,Q
00960      PRINT,*CF BRINE*,CF
00970      PRINT,*CF IBUTANE*,CPB
00980      PRINT 139
00990  139  FORMAT(/,6X,*STAGE*,5X,*TH20*,6X,*TCH4*,6X,*E*,
01000+      10X,*MW*,8X,*HW*,7X,*HB*,7X,*MS*,9X,*HS*,
01010+      9X,*PS*,7X,*PB*,7X,*E2*)
01020  140  FORMAT(I10,2F10.1,F10.4,7F10.1,F10.4)
01030  141  FORMAT(10X,2F10.1,10X,7F10.1)
01040C

```

Sieve Plate (cont'd)

```
01050C      THIS SECTION PERFORMS COMPLETE M/H BALANCE
01060C      AROUND EACH PLATE OF GEOTHERMAL HEAT EX,(FRINODIG 4,77)
01070C
01080C
01090C      CHECK IF BRINE OUT OF TOP PLATE IS LIQUID
01100C
01110      Istage=1
01120      I=1
01130      PS(I+1)=0.
01140      F=2.0
01150      AMS(I+1)=0.
01160C
01170C      CALCULATE STEAM OUT OF TOP PLATE
01180C
01190      CALL WPROP(0,T(I+1),P,0,4,PS(I+1),CPS(I+1))
01200      CALL PHASEB(T(I+1),P,PS(I+1))
01210      AMS(I+1)=PS(I+1)/(P-PS(I+1))*MB*18.02/58.124
01220C
01230C      GUESS TEMP FOR BUTANE ENTERING PLATE
01240C
01250      10 T(I+3)=T(I+1)
01260C
01270C      PERFORM BALANCE AROUND PLATE
01280C
01290      30 CONTINUE
01300      CALL PICEBAL(AMS,PS,MB,F,AMW,PB,HB,T,WP,HW,HS,I)
01310C
01320C      CALCULATE ENTHALPY OF BUTANE IN, BY EQ. 2
01330C
01340      EQUAT2=(AMW(I+2)*HW(I+2)+MB*HB(I+1)+AMS(I+1)*HS(I+1)
01350+      -AMW(I)*HW(I)-AMS(I+3)*HS(I+3))/MB
01360C
01370C      CHECK DIFF. BETEEN ENTHALPY OF BUTANE IN CALC BY
01380C      EQ. 2 AND BY SUBROUTINE ETHB
01390C
01400      33 DIFF=ABS(EQUAT2-HB(I+3))/EQUAT2*100.
01410      IF(DIFF.LE..01) GO TO 45
01420C
01430C      SET NEW VALUE IF ENTHALPY HASN'T CONVERGED.
01440C
01450      IF(AMS(I+3).GT.0.) GO TO 40
01460      TNG=((EQUAT2-HB(I+3))*F)+T(I+3)
01470      CALL ETHB(TNG,PB(I+3),HNC,1)
01480      F=(HNC-HB(I+3))/(TNG-T(I+3))
01490      T(I+3)=TNG
01500      GO TO 30
```

Sieve Plate (cont'd)

```

01510C
01520C      IF STEAM IS PRESENT USE ITER TECHNIQUE TO
01530C      FIND INTERSECTION OF EQUAT2 AND HB
01540C
01550      40 CONTINUE
01560          CALL PICEBAL(AMS,PS,MB,P,AMW,PB,HB,T,WP,HWS,HS,I)
01570          EQUAT2=(AMW(I+2)*HW(I+2)+MB*HB(I+1)+AMS(I+1)*HS(I+1)
01580+          -AMW(I)*HW(I)-AMS(I+3)*HS(I+3))/MB
01590          IF(EQUAT2.GT.HB(I+3)) GO TO 71
01600          T(I+3)=T(I+3)-1.
01610          GO TO 40
01620      71 CONTINUE
01630          CALL PICEBAL(AMS,PS,MB,P,AMW,PB,HB,T,WP,HWS,HS,I)
01640          EQUAT2=(AMW(I+2)*HW(I+2)+MB*HB(I+1)+AMS(I+1)*HS(I+1)
01650+          -AMW(I)*HW(I)-AMS(I+3)*HS(I+3))/MB
01660          IF(EQUAT2.LT.HB(I+3)) GO TO 74
01670          T(I+3)=T(I+3)+.1
01680          GO TO 71
01690      74 CONTINUE
01700          CALL PICEBAL(AMS,PS,MB,P,AMW,PB,HB,T,WP,HWS,HS,I)
01710          EQUAT2=(AMW(I+2)*HW(I+2)+MB*HB(I+1)+AMS(I+1)*HS(I+1)
01720+          -AMW(I)*HW(I)-AMS(I+3)*HS(I+3))/MB
01730          DH2=ABS(EQUAT2-HB(I+3))
01740          IF(EQUAT2.GT.HB(I+3)) GO TO 45
01750          T(I+3)=T(I+3)-.01
01760          GO TO 74
01770C
01780C      SET UP VALUES FOR NEXT PLATE
01790C
01800      45 E=(T(I)-T(I+2))/(T(I)-T(I+1))
01810          E2=(T(I)-T(I+2))/(T(I)-.5*(T(I+1)+T(I+2)))
01820          PRINT 140,ISTAGE,T(I),T(I+1),E,AMW(I),HW(I),HB(I+1),
01830+          AMS(I+1),HS(I+1),PS(I+1),PB(I+1),E2
01840          ISTAGE=ISTAGE+1
01850          IF(I.EQ.23) GO TO 49
01860          I=I+2
01870          GO TO 10
01880      49 PRINT 141,T(I+2),T(I+3),AMW(I+2),HW(I+2),HB(I+3),AMS(I+3),
01890+          HS(I+3),PS(I+3),PB(I+3)
01900          PRINT 185
01910      185 FORMAT(14/)
01920      50 CONTINUE
01930          END

```

Appendix K

Example Sieve Tray Calculation

Typical operating conditions are shown in Figure K-1 for a sieve tray DCHE column using supercritical isobutane working fluid and a mass flow rate ratio r of 1.0.

Under these conditions, the limiting throughput occurs at the bottom tray which is at 175°F and 650 psia. The physical properties of the fluids at the bottom tray conditions are:

$$\begin{aligned}\rho_W &= 60.6 \text{ lb/ft}^3 \\ \rho_B &= 30.4 \text{ lb/ft}^3 \\ \sigma &= 0.0041 \text{ lb. force/ft} \\ \mu_W &= 2.42 \times 10^{-4} \text{ lb/(ft)(sec)}\end{aligned}$$

For this example, a 12 foot diameter column is taken as a basis because it represents the largest readily transportable column diameter.

Column Throughput

Since the area A_T of a 12-foot diameter tray is 113.1 ft², Equation 7-18 gives $m_W = 1,582,000 \text{ lb/hr}$. This corresponds to superficial brine mass velocity at flooding based on total cross sectional area of 14,000 lb/hr-ft².

Sieve Plate Configuration

Downcomer Area. The terminal velocity u_t of a 1/32-inch diameter isobutane drop in brine is estimated from the Hu-Kinter correlation (Hu-Kinter, 1955) using the dimensionless groups.

C = drag coefficient

$$N_{We} = (u_t^2 D_p \rho_W / \sigma g_c)$$

$$N_{Re} = (D_p u_t \rho_W / \mu_W)$$

$$P = \frac{g_c^3 \rho_W^2 \sigma^3}{g \mu_W^4 (\rho_W - \rho_B)}$$

Figure K-1. Operating conditions for the supercritical isobutane-brine system with equal mass flowrates.

Since

$$CN_{We} = 4D_p^2 (\rho_W - \rho_B)/3 \quad (K-1)$$

and using bottom tray fluid properties to obtain $P = 2.55 \times 10^{12}$, then

$$CN_{We}(P)^{0.15} = 4.80 \quad (K-2)$$

From the Hu-Kinter correlation and Equation K-2

$$N_{Re}/P^{0.15} = 1.9 \quad (K-3)$$

Using $P = 2.55 \times 10^{12}$ and the definition of Reynolds number in Equation K-3, and solving

$$u_t = 0.213 \text{ ft/sec}$$

Inserting the value of u_t and the column throughput of 1,582,000 lbs/hr in Equation 7-12 gives the downcomer area

$$A_D = 34.1 \text{ ft}^2$$

Bubbling area. For this design, 1/4-inch holes are placed on 1/2-inch equilateral triangle centers which results in a hole area equal to 22% of the bubbling area. An isobutane velocity through the holes of 1.5 ft/sec is used, which is in agreement with the criteria discussed in Section 7.

From Equation 7-13, $Q_B = 14.45 \text{ ft}^3/\text{sec}$. Solving Equations 7-14 and 7-15 simultaneously gives

$$A_B = 43.8 \text{ ft}^2.$$

Configuration. The tray layout using these areas is shown in Figure 7-3, page 51.

Tray Spacing

Depth of the coalesced isobutane layer below the plate is given by

$$z_B = \frac{u_H^2 \rho_B}{30.1 (\rho_W - \rho_B)} + \frac{4.5 u_W^2 \rho_W}{2g_C (\rho_W - \rho_B)} \quad (7-21)$$

The first term, pressure drop through the holes, with $u_H = 1.5$ ft/sec is

$$\frac{u_H^2 \rho_B}{30.1 (\rho_W - \rho_B)} = 0.08 \text{ ft}$$

The second term, pressure drop due to brine flow through the downcomers gives

$$\frac{4.5 u_W^2 \rho_W}{2g_C (\rho_W - \rho_B)} = .006 \text{ ft}$$

and

$$z_B = 0.086 \text{ ft}$$

The height for brine flow is given by

$$z_W = 0.5 A_D / W_L \quad (7-22)$$

Obtaining $W_L = 9.9$ ft from Figure 7-3

$$z_W = 1.72 \text{ ft}$$

The minimum tray spacing, Equation 7-19, is

$$z = z_B + z_W = 1.81 \text{ ft}$$

Therefore a tray spacing of 24 inches with downcomers extending 4 inches below the plate is convenient.

Appendix L

Derivation of Cost Equations

In the computer model description Appendix A, the section on Cost makes use of several different equations for the various vessel and internals costs. These equations are derived below. All costs are escalated to mid-1975 using the Jacobs Engineering Index (Savay, 1975).

Cost Equation for the Vessel Shell

$$C_v = 16.21 [136D(H+0.8D) \left(\frac{6PD}{SE} + c + x \right)]^{0.66} \quad (L-1)$$

This is the basic vessel cost obtained by combining the following three equations. The vessel cost is given (Perry, 1973) as a function of the total weight.

$$C_v = 9.05W^{-0.34} \quad (L-2)$$

The vessel weight is given (Perry, 1950) as a function of vessel height, diameter, and shell thickness.

$$W = 136D (H+0.8D) (t+x) \quad (L-3)$$

Shell thickness is calculated from the standard stress relationship (Rase and Barrow, 1957).

$$t = \frac{6PD}{SE} + c \quad (L-4)$$

Applying the Jacobs Engineering Index (Savay, 1975) for escalation from 1968 results in the 16.21 factor used in Equation L-1. In these equations

C_v = F.O.B. cost of one vessel shell, including nozzles, manholes, etc.

c = Corrosion allowance, inches

D = Vessel diameter, feet

E = Joint efficiency

H = Vessel height or length, feet
 P = Design pressure (operating pressure plus 10%), psig
 S = Allowable working stress, psi
 t = Vessel wall thickness, inches
 W = Total vessel weight, pounds
 x = Equivalent thickness to be added to account for
 nozzles, manholes, clips, internal support members,
 etc., inches
 9.05 = Cost factor (1968 basis) for carbon steel vessels
 (Perry, 1973)

The following values were used in the program, based on a carbon steel vessel.

S = 13,750 psi
 E = 0.85
 c = 0.125 inch
 x = 0.08 inch
 D = 12 feet, maximum

The Field costs factor for vessels and columns is 1.3 (Guthrie, 1969). This factor accounts for such items as freight, insurance, construction equipment, small tools, temporary facilities and other miscellaneous field costs.

Incorporating the above values in Equation L-1 and applying the 1.3 factor to obtain total direct materials costs for the columns and vessels results in the total direct materials costs of one vessel shell.

$$C_V = 21.08 [136 D (H+0.8D) (PD/1948.+0.205)]^{0.66} \quad (L-5)$$

Cost of Column Internals

Sieve tray column. The shop installed cost of sieve trays is \$8.66 per ft² (Perry, 1973). The cost of internals for one column is

$$C_T = \left(\frac{\pi D^2}{4}\right) (8.66) (N)$$

$$C_T = 7.83 ND^2 \quad (L-6)$$

Thus the direct materials cost of one sieve tray column, 12 feet or less in diameter, is the sum of Equations L-5 and L-6:

$$C_S = C_V + 7.83ND^2 \quad (L-7)$$

Packed column. Using a value of \$70 per ft³ for Goodloe packing, and a height equivalent to one stage of 0.7 feet, the cost of internals for one packed column, 12 feet or less in diameter is:

$$C_K = 70 \left(\frac{\pi D^2}{4} \right) (0.7N)$$

$$C_K = 38.5ND^2 \quad (L-8)$$

Thus the direct materials cost of one packed column 12 feet or less in diameter is the sum of Equations L-5 and L-8.

$$C_P = C_V + 38.5ND^2 \quad (L-9)$$

Mixer-Settler. Although the mixer-settler vessels have no additional internals costs, there are other costs associated with the system. Each stage requires two pumps and one mixer. FOB costs for the pumps are estimated to be \$7000 each, and for the mixers, \$10,000 each. As discussed in Appendix A, each stage of the mixer-settler system is sized independently. Capacity relationships are given in Appendix M. The total cost is the sum of the cost of the vessels, pumps and mixers for each stage:

$$C_M = \sum_{i=1}^N C_{V,i} + N(10,000) + 2N(7000) \quad (L-10)$$

Field cost factors for the pumps and mixers is 1.0. Thus the total direct materials cost for the mixer-settler system is

$$C_M = \sum_{i=1}^N C_{V,i} + 24,000N \quad (L-11)$$

Appendix M

Derivation of Mixer-Settler Capacity Equations

The mixer-settler vessels are sized according to three separate criteria for the three zones illustrated in Figure M-1. The final separation stage, I-Cold Brine Separator, is designed for good separation of the cold brine and isobutane to minimize loss of the isobutane working fluid. II-Intermediate Stages are designed for the same degree of isobutane-brine separation as the sieve tray downcomers discussed in Section 7. III-Hot Isobutane Separator capacity is based on vapor-liquid separation criteria to prevent brine carryover in the hot isobutane vapor to the turbine.

I. Cold Brine Separator

The vessel sizing basis for good separation of brine and isobutane is that the terminal velocity u_t under Stoke's Law conditions corresponds to an isobutane drop diameter D_p of 0.005 inches or 127 microns (Perry, 1973), (Oliver, 1966). The terminal velocity under Stoke's Law conditions (i.e. $N_{Re} < 0.3$) is

$$u_t = \frac{g D_p^2 (\rho_W - \rho_B)}{18\mu_W} \quad (M-1)$$

Substituting numerical values for D_p and g

$$u_t = 3.1 \times 10^{-7} \left(\frac{\rho_W - \rho_B}{\mu_W} \right) \quad (M-2)$$

To relate the drop velocity to the settler dimensions described below refer to Figure M-2. The time t for separation of a drop of isobutane from the brine is the residence time of brine in the vessel. This time is the volume of the vessel occupied by the brine divided by the brine flow rate

$$t = \frac{V_W}{Q_W} \quad (M-3)$$

HOT
ISOBUTANE
OUTLET

HOT
BRINE
INLET

III HOT ISOBUTANE
SEPARATOR

II INTERMEDIATE
STAGES

COLD
ISOBUTANE
INLET

I COLD BRINE
SEPARATOR

COLD
BRINE
OUTLET

Figure M-1. Schematic of the countercurrent flow mixer-settler system.

ISOBUTANE - BRINE
MIXTURE INLET

ISOBUTANE OUTLET

ISOBUTANE PHASE

D

z

BRINE PHASE

H

BRINE OUTLET

- M3 -

Figure M-2. Nomenclature for a settling vessel.

For steady state, this time is also equal to the ratio of the isobutane volume and flow rate and to the total mixture volume and sum of the flow rates

$$t = \frac{V_B}{Q_B} = \frac{V_T}{Q_T} \quad (M-4)$$

The total volume excluding heads is

$$V_T = \frac{\pi D^2 H}{4} \quad (M-5)$$

The total volumetric flow rate can be written

$$Q_T = \left(\frac{m_W r}{\rho_B} + \frac{m_W}{\rho_W} \right) \frac{1}{3600} \quad (M-6)$$

Combining Equations M-4, M-5, and M-6

$$t = \frac{\pi D^2 H (900)}{m_W \left(\frac{r}{\rho_B} + \frac{1}{\rho_W} \right)} \quad (M-7)$$

Figure M-2 shows the distance Z through which the isobutane drop separates from the brine. The velocity of the drop can therefore be expressed as

$$u \approx \frac{Z}{t} \quad (M-8)$$

The distance Z is given by the geometry of the vessel and the flow rate ratio r. However, for r between 0.5 and 1.0 the ratio of the distance Z to the diameter D is between 0.37 and 0.5. As a simplification this ratio is assigned a value of 0.4, or

$$Z = 0.4D \quad (M-9)$$

Combining Equations M-7, M-8 and M-9 and simplifying

$$D H = \frac{450 m_W \left(\frac{r}{\rho_B} + \frac{1}{\rho_W} \right) \mu_W}{(\rho_W - \rho_B)} \quad (M-10)$$

II. Intermediate Stages

The vessel sizing basis for the intermediate stages is that the terminal velocity corresponds to an isobutane drop size of 1/32-inch which is equivalent to the downcomer velocity criteria in the sieve tray DCHE column discussed in Section 7.

Using the values developed in Section 7 and Appendix K, the expression for the terminal velocity for 1/32-inch drops can be written

$$u_t = 0.30 \sqrt{\frac{\rho_W - \rho_B}{\rho_W}} \quad (M-11)$$

From Equation M-8 the velocity in the vessel can be expressed as

$$u = \frac{Z}{t} \quad (M-8)$$

where Z is approximated by Equation M-9 and t is obtained from the vessel dimensions and the flow rates of isobutane and brine given by Equation M-7. Combining these equations with Equation M-11 and simplifying

$$D H = \frac{m_W \left(r \frac{\rho_W}{\rho_B} + 1 \right)}{2100 \sqrt{\rho_W (\rho_W - \rho_B)}} \quad (M-12)$$

III. Hot Isobutane Separator

From conventional criteria for allowable mass velocity in a vapor-liquid separator (Perry, 1973)

$$G = 1000 [\rho_B (\rho_W - \rho_B)]^{0.512} \quad (M-13)$$

The minimum cross sectional area is

$$A = \frac{m_W r}{G} = \frac{\pi D^2}{4} \quad (M-14)$$

Combining Equations M-13 and M-14

$$D = \frac{0.036 \sqrt{r m_W}}{[\rho_B (\rho_W - \rho_B)]^{0.256}} \quad (M-15)$$

Combining Equations M-13 and M-14 and solving for the cross sectional area

$$A = \frac{m_W r}{1000 [\rho_B (\rho_W - \rho_B)]^{0.512}} \quad (M-16)$$

The height of liquid brine z_L in the vessel is

$$z_L = \frac{V_W}{A} \quad (M-17)$$

Allowing 1 minute of residence time for the brine hold up in the vessel the brine volume is

$$V_W = \frac{m_W}{60 \rho_W} \quad (M-18)$$

Combining this expression with Equations M-16 and M-17

$$z_L = \frac{16.7}{\rho_W r} [\rho_B (\rho_W - \rho_B)]^{0.512} \quad (M-19)$$

To obtain an approximate value for z_L the densities are evaluated for conditions in the vessel of 290°F and 650 psia. Then

$$z_L \approx \frac{6.0}{r} \quad (M-20)$$

Allowing 6 feet of height for inlet piping and vapor disengaging, the total vessel height is

$$H = Z_L + 6 \quad (M-21)$$

For $r=1.0$, $Z_L=6.0$ by Equation M-20 and

$$H = 12 \text{ ft} \quad (M-22)$$

Equation M-22 is a valid approximation when the conditions in stage III are 280-320°F and 600-650 psia, and when the isobutane to brine mass ratios are in the range of 0.5 to 1.0 combined with liquid residence times of 0.5 to 1.0 minutes, respectively.