MASTER

FINAL REPORT

GEOTHERMAL DIRECT-CONTACT HEAT EXCHANGE

ERDA CONTRACT NO. E(04-3)-1116

10 JUNE 1976

EY-76-C-03-1116

PREPARED FOR THE

UNITED STATES ENERGY RESEARCH
AND DEVELOPMENT ADMINISTRATION

BY

ANKER V. SIMS

(213-684-2541)

THE BEN HOLT CO.
201 SOUTH LAKE AVENUE
PASADENA, CALIFORNIA 91101

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SUMMARY

A glass direct contact heat exchange column was operated in the laboratory. The column was operated at atmospheric pressure using hot water and normal hexane. Column internals tested included an empty column, sieve trays, disk-and-doughnut trays, and two types of packing. Operation was very smooth in all cases and the minimum temperature approaches varied from less than 1°C for packing to 13°C for the empty column. High heat transfer rates were obtained in all cases, however, columns should be sized on the basis of liquid and vapor traffic.

The solubilities of hydrocarbons were determined for normal hexane, pentane and butane in water and sodium chloride and calcium chloride brines at various temperatures. The values seem to be internally consistent and salt content was found to depress hydrocarbon solubility.

Laboratory stripping tests showed that gas stripping can be used to remove hydrocarbon from reject hot water from the direct contact heat exchange column. Although the gas volumes required are small, stripping gas requirements cannot be accurately predicted without testing.

A computer program was used to study the effect of operating variables on thermodynamic cycle efficiencies. Optimum efficiencies for the moderate brine conditions studied were obtained with isopentane as working fluid and relatively low operating pressure.

A preliminary design for a 50 MWe plant was prepared and plant capital cost and operating cost were estimated. These costs were combined with previously developed brine production and power transmission costs to provide an estimate of the cost of delivered power for a geothermal field at Heber, California.

A pilot plant program is described that would be suitable for continuing the investigation of the direct contact process in the field. The program includes a suggested schedule and the estimated cost.

INTRODUCTION

Sufficient work has been done to date to establish geothermal energy as a significant source for meeting our future energy needs. At the present time there are only a few geothermal power plants in operation; in the United States, The Geysers field is the only location of commercial importance. The Geysers field is a vapor dominated reservoir, producing steam that is used directly in steam turbines. Such reservoirs are comparatively rare, and represent only a small portion of the available geothermal energy. Dry rock or magma sources are a potentially much larger source of geothermal energy, but the technology for utilizing these resources has not been developed. Liquid dominated reservoirs are a major source of geothermal energy and technology is available for exploiting them. Thus, liquid dominated reservoirs are the geothermal energy resource of major interest at this time.

The principal power generation processes being considered for use with liquid dominated geothermal reservoirs are the binary, the flashed steam, and the direct contact; although many variations within the general classifications are possible. The flashed steam process is probably the simplest and is suitable for use with heavy scaling brines. The capital cost for the plant is probably lowest of the three, but brine usage for unit of power produced tends to be high. The binary process appears to be more efficient, but depends on the use of shell and tube heat exchangers for transferring heat from the brine to the working fluid. The use of such exchangers is impractical for brines with heavy scaling characteristics, and in any event the cost of the exchangers is a large part of the total plant cost. The direct contact binary process offers the possibility of reduced plant cost, good efficiency, and suitability for use with scaling brines. The process is, however, the least developed of the three and presents areas of technological uncertainty.

TASKS AND METHOD OF APPROACH

The work reported here is a preliminary investigation of some of the technical and economic aspects of the direct contact process, and was undertaken in order to evaluate the commercial potential of the process. The work consisted of laboratory tests, a conceptual design, a 50 MWe plant design, an economic study, and a pilot plant program.

The laboratory tests were made to obtain enough information to develop a realistic preliminary plant design. A glass direct contact heat exchange column was designed, constructed, and operated to get information on general performance characteristics, tray efficiencies, sizing correlations, and temperature approaches. A hot water-normal hexane system was used to permit operation in a glass column at atmospheric pressure. The physical characteristics of this system are close to those proposed for the commercial plant so the answers obtained are applicable to the commercial system. Hydrocarbon stripping tests were made using a glass column and nitrogen as stripping gas in order to establish the feasibility of recovering hydrocarbon from spent hot water. The tests showed that more work is needed in this area but that hydrocarbon stripping is feasible. Solubility tests were conducted to determine the effect of temperature, and brine composition on the solubility of three hydrocarbons. This information was needed to evaluate the impact of hydrocarbon losses on the cost of power production.

A computer program was developed to aid in the evaluation of various possible direct contact thermodynamic cycles. The effect of operating variables on power production was evaluated; including the effect of brine inlet temperature, operating pressure, condensing temperature and choice of working fluid. After a general investigation, an optimum cycle was found for a specific site, the Heber field. The Heber field, located in the Imperial Valley of Southern California, was chosen as representative of a medium temperature low salinity reservoir with excellent commercial potential.

As part of the conceptual design, studies were made on the prevention of working fluid loss in rejected hot water and in noncondensable gases, and the elimination of particulates in expander inlet gas. The studies were used to establish practical solutions to these important design problems, at least for the preliminary design.

The results of the conceptual design were used to develop a preliminary design for a 50 MWe (delivered) plant. The plant was sized for 50 MWe since preliminary studies made as part of another project had established this as an economic size. An estimated capital cost was developed for the 50 MWe plant, for use in the economic study.

The economic study made use of a considerable body of information developed as part of a project done for the Electric Power Research Institute by Holt/Procon, a joint venture of The Ben Holt Co. and Procon Incorporated. This information included plant and operating costs for a closed loop binary and a two-stage flashed steam process as well as costs for brine production and power transmission. The same methods and factors were used for the direct contact process. Thus, it was possible to obtain a comparison among the three processes on a consistent basis. The relative values obtained should be relatively accurate.

Since the process shows promise, a pilot plant program was prepared outlining areas for future work.

CONCLUSIONS AND RECOMMENDATIONS

The laboratory work and design studies show that the use of the direct contact heat exchange process in the production of electric power is technically feasible. Preliminary cost estimates indicate that the direct contact process is competitive with other geothermal processes and with the production of power from petroleum based fuels. Where brine scaling is not a problem, the binary process appears to be the best choice for use with medium or low temperature brines. The direct contact process is competitive with, and may be superior to, the steam flash process for medium and low temperature scaling brines.

Some of the technical points established by the study are:

1. Operation of the laboratory direct contact heat exchanger was smooth and trouble free.

2. Existing correlations used for commercial design of distillation columns are satisfactory for sizing direct contact heat exchange columns.

3. Stage efficiencies for multistage direct contact heat exchange devices can be determined in laboratory or pilot plant equipment.

4. Direct contact heat exchanger size is primarily set by requirements for vapor and liquid traffic, not by volumetric heat transfer needs.

5. Many different contacting devices can be used for the direct contact heat exchanger; the choice is a matter of cost and brine scaling characteristics.

6. Multiple stage devices, or true countercurrent devices, are necessary to achieve good thermal efficiency. Commercial sized spray towers may not be satisfactory.

7. Given a brine inlet temperature and pressure and a condensing temperature: there is an optimum choice of hydrocarbon working fluid and operating pressure. The working fluid may be a single compound or a mixture of close boiling point materials, depending on conditions.

8. If noncondensable gases are present in the brine, they should probably be eliminated before the brine is brought into contact with the working fluid.

9. Satisfactory scrubbers are available for removing particulates from working fluid vapor.

10. Gas stripping can be used to remove hydrocarbons from spent brine, if necessary for economic or environmental reasons.

The technical points still needing resolution are in general those associated with working with actual geothermal brines. These points include:

1. Scale formation. Where does the scale form and where is it deposited, assuming it forms at all.

2. Foaming in the direct contact column. It has been suggested that foaming may be a problem, but no information is available.

3. Hydrocarbon solubility in geothermal brines. The laboratory data show that salt content depresses solubility, but tests on brines under field conditions are needed. Most available hydrocarbon solubility data are for water and for three-phase conditions. The location of interest is the bottom of the contactor, where two-phase conditions prevail. There is also a distinct possibility that solubility rate may be important, and there is very little information on this subject.

4. The fate of noncondensable gases entering the systems. The solubility of noncondensables in working fluid, brine and condensate.

In view of the technical and economic promise of the direct contact process, we strongly recommend that pilot scale tests be conducted in the field on actual geothermal brines.

LABORATORY TESTS

DIRECT CONTACT HEAT EXCHANGE COLUMN

Equipment

The general layout of the direct contact heat exchange equipment is shown in the photographs of Figure 1 and the sketch of Figure 2. An insulated 50-gallon drum, provided with a 5 kw electric immersion heater, is used to supply hot water to the system. Hot water flows by gravity through a metering valve and rotometer to the top of the glass heat exchange column. Except for a very small amount of water vaporized into the overhead vapor stream, the water flows out of the bottom of the column, through plastic tubing and drains into a bucket. The elevation of the tubing outlet is adjusted to maintain the desired liquid level in the column. A pump is used to recycle the drained water to the supply drum.

Normal hexane is provided to the system from a gas tight supply drum. The drum is filled with enough hexane for one or more runs and is pressured with nitrogen to avoid the use of a pump or of an elevated drum. Hexane flows to the bottom of the column through a metering valve and a rotometer and enters the column through a section of 1/4-inch copper tubing. The end of the copper tubing is closed and hexane flows into the column through a single 1/16-inch diameter hole located at the axis of the column. Liquid hexane enters the bottom of the column, flows upward because of density difference, is vaporized and leaves the top of the column as a vapor along with some vaporized water. The vapor is condensed by passing it through a series of double pipe, water-cooled, condensers. The condensed hexane and water drain from the condensers into a 5-gallon drum.

Thermometers are inserted in the lines entering and leaving the glass column to measure the temperatures of the inlet hot water, the outlet cooled water, the inlet cold hexane, and the overhead exit vapor. Weston bimetallic thermometers with 2-inch dials are used at all locations.

The glass heat exchange columns were constructed of 70 mm (2.75-inch) inside diameter glass tubing with Corning No. 6780 "O" ring joints at each end. The end caps, containing the inlet and outlet tubulations, were fitted with matching "O" ring joints. Both 305 mm (one foot) and 1220 mm (4-foot) glass sections were used in testing. The columns were tested empty, equipped with metal trays of various designs, with 6.4 mm (1/4-inch) ceramic Intalox saddles, and with 15.9 mm (5/8-inch) plastic Pall rings. Photographs of several of the column internals used are shown in Figure 3. During most of the runs the column was insulated with 25 mm (one-inch) thick rigid polyurethane pipe insulation.

The various column internals used in the tests were assigned code numbers as a means of identification. The first digit is the length of column in feet, the next digit is number of trays, and the letters are packing or tray modification identifiers. The code numbers and corresponding descriptions are as follows:

1-4. A one-foot column section with 4 sieve trays, spaced 102 mm (4 inches) apart. Spacing is maintained by 3 metal rods passing through the trays. Figure 4 is a sketch of the top tray. As shown, the tray to column seal is accomplished by means of a rubber tubing "O" ring. The holes are flared to provide jet nozzle outlets for the hydrocarbon phase. Downcomers consist of 25 mm (one-inch) sections of 8 mm (5/16-inch) I.D. brass tubing.

Tray No.	Hole Diameter	Number of Holes
1	4.8 mm (3/16 inch)	20
2	4.8 mm (3/16 inch)	10
3	4.8 mm (3/16 inch)	7
4	3.2 mm (1/8 inch)	5

4-8. A 4-foot column section with 8 sieve trays, spaced on 114 mm (4.5-inch) centers. Threaded 3.2 mm (1/8-inch) brass rods with nuts are used for spacing. Tray construction is simplified by eliminating the flaring of the holes and by using 6.4 mm (1/4-inch) diameter neoprene tubing, cemented to the tray, as a tray to column seal.

Tray No.	Hole Diameter	Number of Holes
1	4.8 mm (3/16 inch)	20
2	4.8 mm (3/16 inch)	16
3	4.8 mm (3/16 inch)	10
4	4.8 mm (3/16 inch)	9
5	4.8 mm (3/16 inch)	8
6	4.8 mm (3/16 inch)	5
7	3.2 mm (1/8 inch)	7
8	3.2 mm (1/8 inch)	5

4-3. A 4-foot column section with 3 sieve trays, spaced on 305 mm (12-inch) centers.

Tray No.	Hole Diameter	Number of Holes
1	4.8 mm (3/16 inch)	20
2	4.8 mm (3/16 inch)	12
3	3.2 mm (1/8 inch)	10

4-3A. Same as 4-3 except that number of holes in tray number one is increased to 27.

4-0. A 4-foot column section with no column internals.

4-IS. A 4-foot column section with a 406 mm (16-inch) layer of 6.4 mm (1/4-inch) Intalox ceramic saddles supported on a 3.2 mm (1/8-inch) hardware cloth inverted cone.

4-6. A 4-foot column section with 6 sieve trays, spaced on 152 mm (6-inch) centers. Downcomer area for these trays is increased to .000445 square meters (.00479 square feet), over a tenfold increase. Downcomer length is increased to 140 mm (5.5 inches). Figure 5 is a sketch of the modified trays.

Tray No.	Hole Diameter	Number of Holes
1-5	5.6 mm (7/32 inch)	19
6	5.6 mm (7/32 inch)	9

4-7A. Same as 4-6 with added 19-hole tray.

4-PR. A 4-foot column section with a 940 mm (37-inch) layer of 15.9 mm (5/8-inch) polypropylene Pall rings. The rings are retained in the column by a 3.2 mm (1/8-inch) hardware cloth disk. No bottom support is used, the rings being allowed to fill the space around the hexane inlet tube.

4-7B. Configuration 4-7A modified by extending a section of downcomer to a height of 25 mm (one inch) above the trays.

4-7C. Same as 4-7B except that tray alignment is improved and a column with a more uniform diameter is used.

4-8D/D. A 4-foot column section with 8 disk-and-doughnut trays (4 each), spaced on 127 mm (5-inch) centers. Four 4.8 mm (3/16-inch) diameter threaded rods are used to support the trays. Figure 6 is a sketch of the trays.

Experimental Data

The results from operation of the various column arrangements for the heat exchanger at widely varying conditions are given chronologically in Table 1 and summarized in Table 2. The changes of the various tray/packing system and flow conditions were a consequence of the attempt to find the maximum practical flow conditions and efficiency for the size and complexity of equipment used. For example, the size of downcomers was increased and the hole area was increased in tests with sieve trays to permit determination of maximum practical water and hexane rates allowed by the size of the column employed.

In Table 1 the identification of column configuration, equilibrium flow rates, pertinent equilibrium temperatures and heat balances are given. Tray efficiency where calculable and temperature of approach derived as described below are also given. An examination of both of these latter pieces of data provides one of the means of assessing the potential of a given system of operation.

Graphical representation of equilibrium and operating temperature/enthalpy relationships was used to assess plate efficiency and temperature approach. For illustration, equilibrium and operating temperature/enthalpy relationships are given in Figure 7 wherein the temperature is plotted against enthalpy expressed as Btu per pound of hexane. The equilibrium curve consists of three parts. First, there is a nominal straight line portion for the liquid hexane as it is heated to the bubble point temperature. It will be noted that this temperature is lower than that for pure hexane because of the contribution of the water vapor to the total pressure. The bubble point will vary with the equivalent head of water in the column above the point where the temperature is high enough to cause bubble formation. The second part is a horizontal straight line representing a constant boiling mixture. The third part represents the increasing ratio of water vapor to hexane where liquid hexane is no longer present. A straight line connecting the inlet and outlet water temperatures is the water line. The line should actually be curved to account for water transferred to the vapor phase, but the correction is small and has been neglected. The theoretical tray steps shown are used to estimate the number of equilibrium plates, and tray efficiency is the ratio of theoretical to actual trays. The measure of temperature approach is shown at the point of minimum distance between the water and hydrocarbon curves.

Discussion

Operating Characteristics - The column operated very smoothly. Liquid hexane entered the bottom of the column as a series of drops roughly 3 mm in diameter. After rising a short distance, a lens of vapor formed, attached to the drop. As the drop continued to rise through the column additional vapor was formed, with a rapid increase in volume. Near the top of the column the vapor bubbles were broken up and the vapor and water formed a violently agitated mixture. Clean separation was obtained between vapor and liquid at the top of the column, and no visible drops of hydrocarbon were entrained in the water leaving the bottom of the column. The mixture in the upper section of the column showed no tendency to foam.

The column was very responsive to feed rates. If the column was operating at steady state and with a higher than minimum water rate, reducing the water rate reduced the exit water temperature. When the minimum water rate was reached, further reductions resulted in a buildup of a hydrocarbon layer in the top of the column. The correlation between column behavior and a graphical representation such as Figure 7 was excellent.

Theoretical Stages - In a heat exchanger the theoretical maximum amount of heat is transferred when the two streams involved flow countercurrent.

Simple bulk mixing is often a very effective way of achieving high heat transfer rates, but the maximum theoretical amount of heat transferred is limited by the requirement that the exit streams leave at the same temperature. Performance equivalent to simple countercurrent flow can be achieved with the use of equilibrium stages, or theoretical trays, where the streams flow countercurrent between stages. The minimum number of theoretical stages required will depend on the driving force, that is temperature difference, available for heat transfer. This number of stages can be conveniently determined by a graphical method such as shown in Figure 7. The ratio of theoretical to actual trays can be used as a measure of tray efficiency. For distillation, a mass and heat transfer operation, the number of actual trays required for a given separation in commercial sized equipment has been found to be the same or less than for laboratory equipment.(1) The same should apply to the use of staged equipment for heat transfer.

When dealing with packed columns the concept of packed height equivalent to a theoretical stage can be used. An alternative approach, leading to the same overall result, is to integrate differential driving force as a function of heat transferred to arrive at the "number of transfer units" required. The height equivalent to a transfer unit is then determined from experimental data.

Spray columns, such as the Elgin column, are not amenable to similar treatment because of the major effect of back mixing in such columns. The amount of back mixing, and thus the approach to countercurrent performance, is strongly affected by column diameter and other variables and is difficult to predict.(2)(3)

Sizing Correlations - The maximum throughput for a given column appears to be limited by the vapor and liquid traffic at the top of the column. Flooding rates obtained with the laboratory sieve trays match those that are predicted by the use of Fair's flooding limit correlation.(4) Conventional sizing correlations used by the manufacturers of column internals can therefore be used with confidence.

The number and size of holes for the sieve trays used in the liquid/liquid section of the laboratory column were determined by applying conventional correlations such as those found in Chemical Engineers' Handbook. The hole area so determined appears to be roughly half of that actually required. The reason for the discrepancy is not known, but may be related to the presence of guide rods and seals in the laboratory equipment.

Volumetric Heat Transfer Coefficient - Volumetric heat transfer coefficients determined from laboratory data varied from 20.4 watts/m^3°C (1,370 Btu/hr ft^3°F) for disk-and-doughnut trays to 189.6 watts/m^3°C (12,700 Btu/hr ft^3°F) for Intalox packing. Although the coefficients are of interest for purposes of comparison, they are not particularly useful for design. The size of equipment will be primarily established by the requirements for handling vapor and liquid traffic. Again, supplying the required heat transfer volume is not sufficient. Stages or actual countercurrent flow must be provided to achieve maximum heat transfer.

Operation with Geothermal Brines - The laboratory operations have been very useful but cannot be expected to yield answers to all the questions that will arise with the use of actual geothermal brines in the field; in particular, the questions of possible foaming of brine-hydrocarbon vapor mixtures and the amount, character and location of possible scale formation.

HYDROCARBON SOLUBILITY

Equipment and Procedure

The solubility of normal hexane, normal pentane, and normal butane in water and brine solutions was determined as a function of temperature. A rocking autoclave (750 ml volume) was used to contain and agitate the water or brine, hydrocarbon liquid, and corresponding vapor phase. A 3 x 3 orthogonal square experimental design was used to explore the effect of temperature, sodium chloride concentration and calcium chloride concentration on the solubility of hexane in water and brines. Sodium chloride and calcium chloride were chosen for the study since they are common constituents of geothermal brines. The temperatures were 27°C (80°F), 88°C (190°F) and 177°C (350°F); salt concentrations were 0, 7 and 14% weight sodium chloride and 0, 4 and 8% weight calcium chloride.

Sampling procedures were modified during the course of the study.

Initially samples were withdrawn from the autoclave into open containers, discarding a volume sufficient to purge sampling lines. Air pressure was used to displace samples from the low temperature runs where the equilibrium pressure was below atmospheric. Later, samples were taken by bubbling them into a volume of Freon TF (Freon 113) solvent. The final method was to take the samples into evacuated glass sample bombs. The solubility values obtained with the evacuated bombs appear to be slightly higher than for the other methods, but the difference is not large.

The analytical procedure is as follows. A measured volume of aqueous sample is acidified with sulfuric acid and extracted with 3 portions of Freon TF solvent. The extracted volume is adjusted to a standard volume by the addition of solvent and the solution is dried using anhydrous sodium sulfate. The infrared absorption of the solution is determined with a Wilks Miran I Fixed Filter Analyzer (3.4 microns) and the hydrocarbon concentration is determined by comparison to calibration curves based on standard solutions.

Discussion

It was necessary to make repeated runs in order to achieve reasonably accurate values for each of the experimental design points. Since 74 runs were made with hexane, the number of runs made with pentane and butane were limited by available time. The test results are summarized in Tables 3, 4 and 5 and in Figure 8. The equation developed to fit the hexane data, and shown in the curves of Figure 8, is:

$$\ln(\text{ppmv}) = 5.345 - 1.488 \times 10^{-2} t + 4.313 \times 10^{-5} t^2 +$$

$$\ln(1 - 9.174 \times 10^{-5}\, St)$$

where: t = °F, and

$$S = \%\,w\ NaCl + \%\,w\ CaCl_2$$

No difference was found between the behavior of sodium chloride and calcium chloride when concentration was expressed as equivalent chloride ion concentration. Since the molecular weight of sodium chloride is 58, that of calcium chloride is 111, and there are two mols of chloride per mol of calcium chloride; the conversion factor for calcium chloride is 1.0 ($58 \times 2/111 = 1.05$). Thus for sodium chloride and calcium chloride the concentration in weight percent can be used directly.

The curves for hexane and propane shown by dashed lines in Figure 8 are from Figure 9A2.1 of the API Technical Data Book.(5) The values from our laboratory tests are considerably lower; and no satisfactory explanation for the difference has been found. The equation fits the data for hexane with an average difference of 15% of the calculated value (standard deviation 16%). As expected, salt depresses the solubility of hydrocarbon in water. The effect seems to be dependent on both salt concentration and temperature. Pentane is more soluble than hexane and butane is more soluble than pentane, which is in agreement with available data from the literature. The single point for the solubility of pentane in brine shows a depressant effect comparable to that for hexane if expressed as a percent of the value for pure water.

HYDROCARBON STRIPPING

Equipment

With the exception of a few runs, the stripping tests employed a 610 mm (2-foot) section of 70 mm I.D. (2.75-inch) glass column similar to that used for the heat exchange column. The column was packed with 610 mm (2 feet) of 16 mm (5/8-inch) polypropylene Pall rings. Nitrogen stripping gas was metered to the bottom of the stripping column through a rotameter. Overhead gas was passed through a dry ice/acetone cold trap to collect stripped hexane and water vapor, which were measured volumetrically. Stripper feed was provided by operating the direct contact heat exchange column and passing the exchange column bottoms to the top of the stripper by gravity flow. Liquid level in the stripper was maintained by adjusting the height of the stripped water product flow outlet. Sample points were provided for both stripper feed and stripped product.

A few runs were made using a 10 plate, 50 mm I.D. (2-inch) glass Oldershaw (sieve tray) column as the stripper. It was difficult to achieve smoother operation with this column because of the low nitrogen to water operating ratio.

Discussion

The results of the stripping tests are summarized in Table 6. Hexane balances given in the table show that on the average the hexane stripped plus hexane in stripped product equal the hexane in the feed. Individual runs, however, show wide variations in loss or gain. This is believed to be the result of the inclusion of more or less entrained hexane in the feed sample; and appears to be a function of sampling rate.

The concentration of hexane in the stripper feed is greater than would be predicted on the basis of solubility alone. Table 7 and Figure 9 show the hexane concentration as a function of water flow rate in the heat exchange column and, although the correlation is only fair, concentration increases with flow rate. The curve in Figure 9 was sent through 62 ppmv at zero flow to be consistent with no liquid carry under at zero flow.

In all of the stripping tests reported in Table 6 it was never possible to achieve a stripped product with less than 19 ppmv. This result was entirely unexpected since preliminary calculations had shown that essentially complete stripping should be easily obtained. The difficulty has been traced to nonhexane contaminants that cannot be stripped and that register as hexane in the analytical method. Passing hot water through a column containing new Pall rings produced a reading equivalent to 10 to

15 ppmv hexane. Freon TF solvent stored in a plastic bottle gave a reading of 109 ppmv. A few preliminary stripping tests using a glass stripping column and feed prepared by stirring hexane and water in a bucket showed that the product could be stripped to 3 ppmv hexane. In brief, the stripped product concentrations given in Table 6 contain a nonhexane blank. A plot of hexane in stripped product as a function of nitrogen/water weight ratio, Figure 10, indicates that the blank is about 20.8 ppmv.

The feed to the stripper contained hexane in excess of that due to solubility. In run 1/29, for instance, the stripper feed contained 1,585 ppmv compared to 62 ppmv for solubility. On the other hand, the quantity of stripping nitrogen used was 74 times the minimum required for removing the nonsoluble hexane. Thus, the nonsoluble hexane has essentially no effect on the stripped product concentration. The soluble hexane is much more difficult to strip, so the stripping curve shown in Figure 10 assumes a stripper feed concentration of 62 ppmv.

The curve shown in Figure 10 is based on the Kremser equation, an equation that is frequently used, in graphical form, in the solution of stripping problems. The equation is

$$E = \frac{S^{n+1} - S}{S^{n+1} - 1}$$

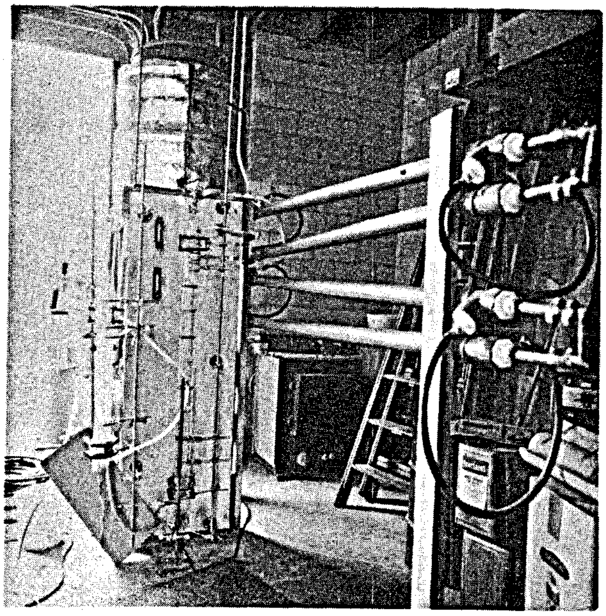
where: n = number of theoretical trays
E = fraction stripped
$S = KV/L$
K = equilibrium constant
V = moles vapor
L = moles liquid

The curve uses $n = 5$ and was made to pass through the "set point." The "set point" was chosen as the point where reducing the nitrogen to water ratio first results in increasing hexane in stripped product. The curve fits the experimental data reasonably well and the assumption of 5 theoretical trays is reasonable for the packing used. Having established the relationship between E and S, and knowing the experimental V/L, it is possible to calculate K. In this case K = 3,214.

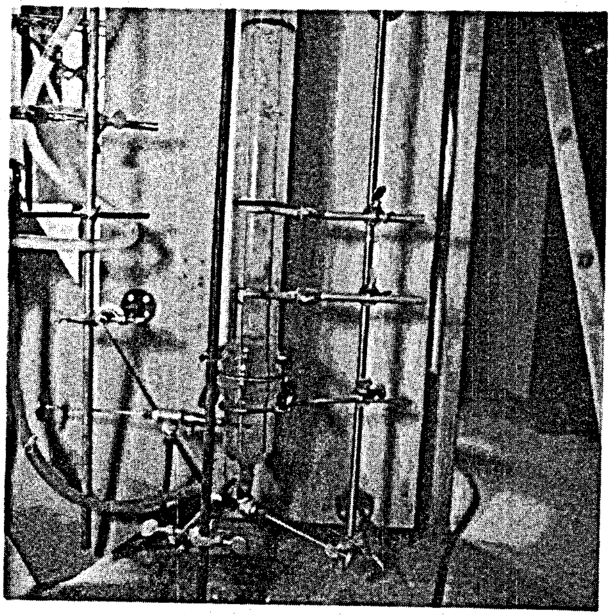
An independent estimate of K can be made for the conditions of the solubility tests. Vapor composition is calculated from hydrocarbon and water vapor pressure data, and hydrocarbon mole fraction is calculated from the solubility of hydrocarbon in water. For 142°F, the calculated K is 93,000. This suggests that the K is a function of liquid composition as well as temperature, and that K = 3,214 is a better value for use in the stripping calculations. Also, it is obvious that additional experimental data would be useful.

FIGURE 1

LABORATORY DIRECT CONTACT HEAT EXCHANGE EQUIPMENT



OVERVIEW



BOTTOM SECTION OF
HEAT EXCHANGE COLUMN

FIGURE 2

LABORATORY EQUIPMENT DIRECT CONTACT HEAT EXCHANGE

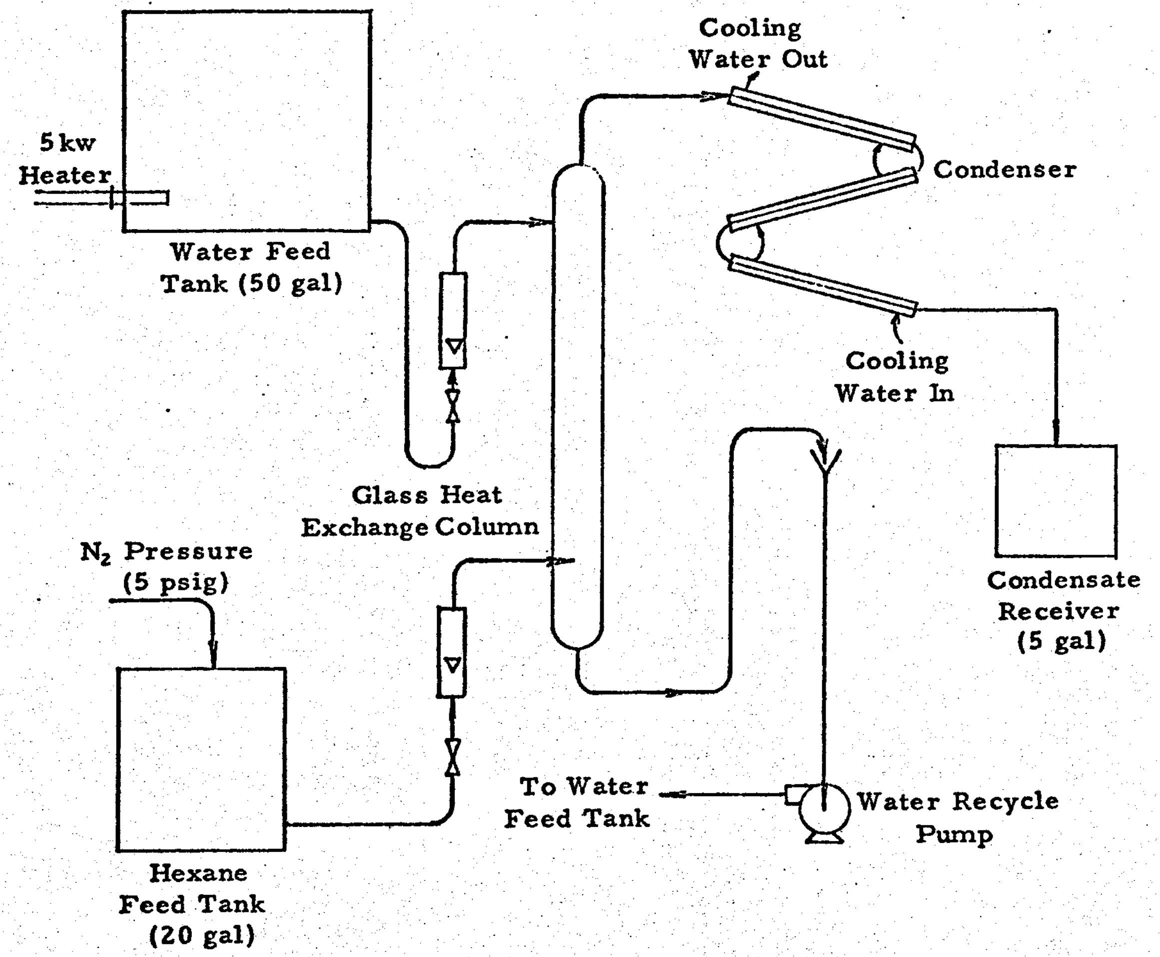
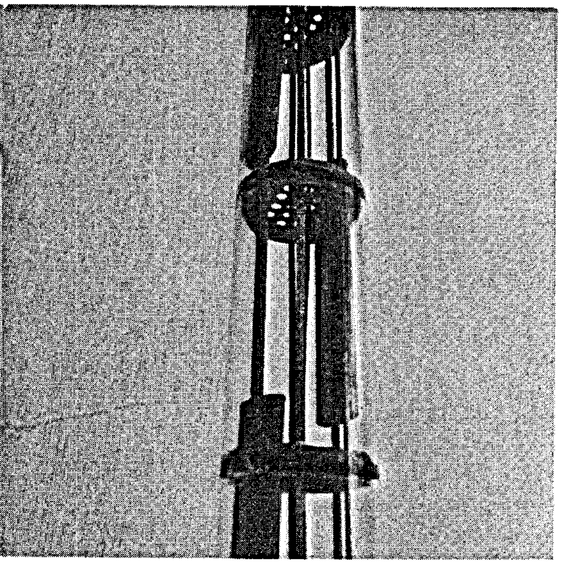


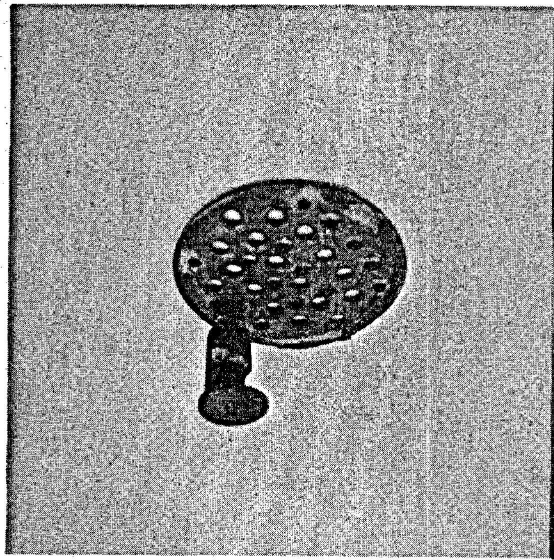
FIGURE 3

HEAT EXCHANGE COLUMN INTERNALS

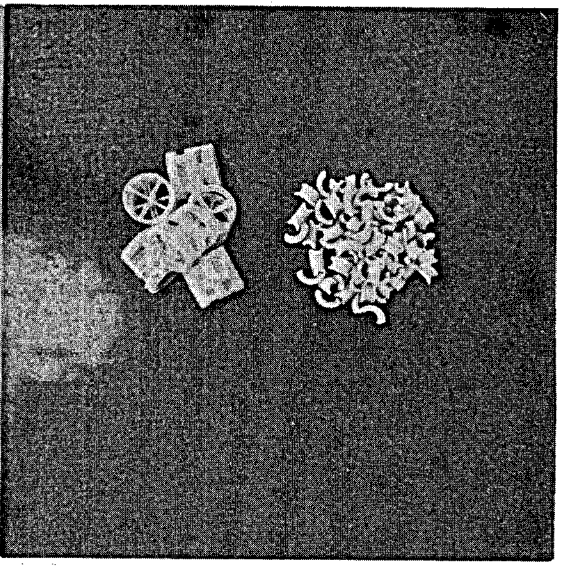
SIEVE TRAYS
(1-4)



SIEVE TRAYS
(4-6)



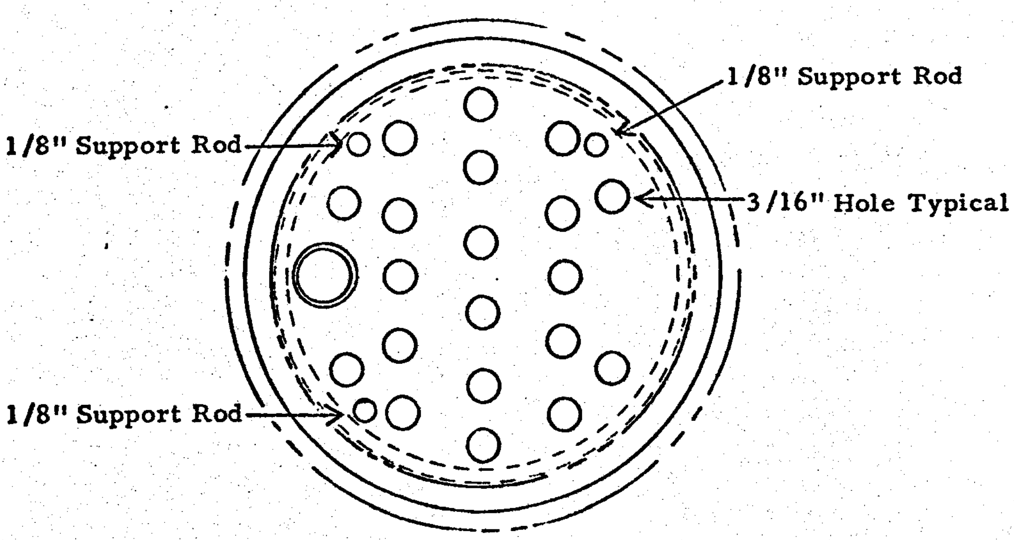
SIEVE TRAY
(1-4 Modified)



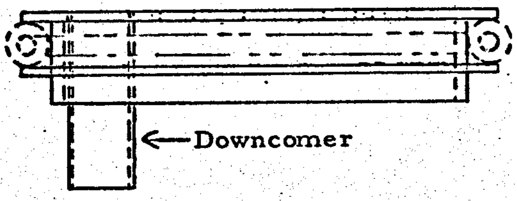
PALL RINGS INTALOX SADDLES
(PR) (IS)

FIGURE 4

SIEVE TRAYS
(1-4)
(Full Scale)



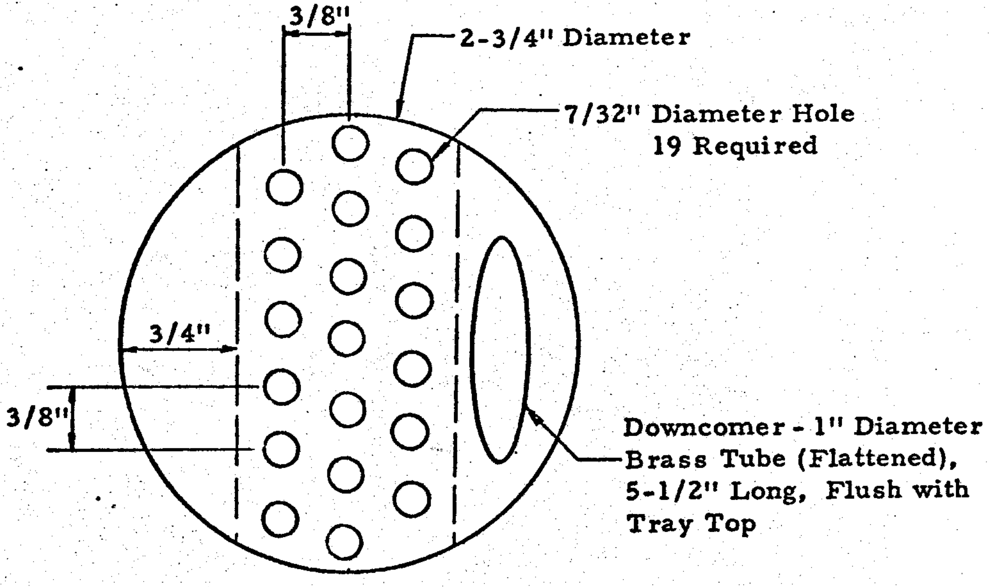
Top Tray
Bottom View



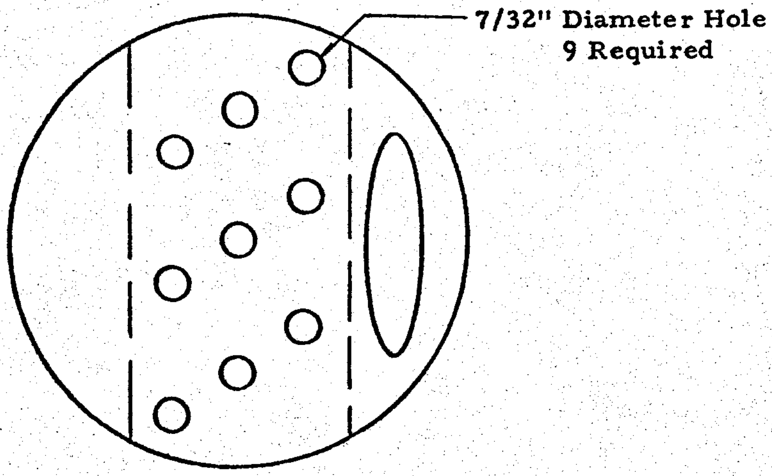
Side View

FIGURE 5

SIEVE TRAYS
(4-6)



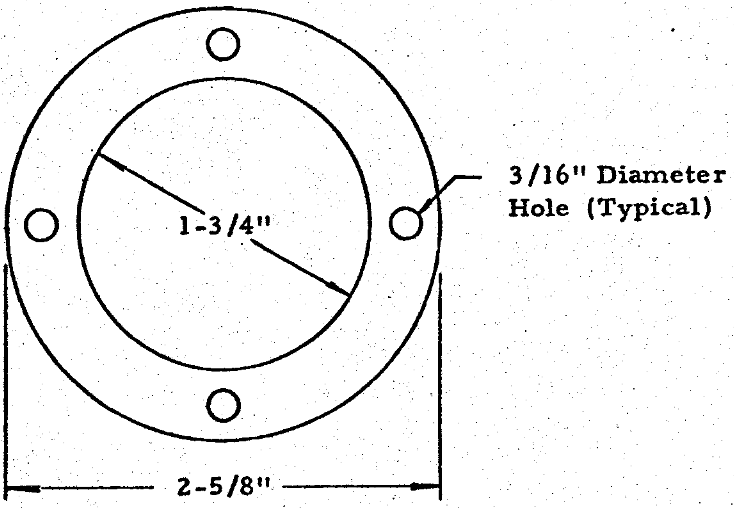
Trays 1 Through 5



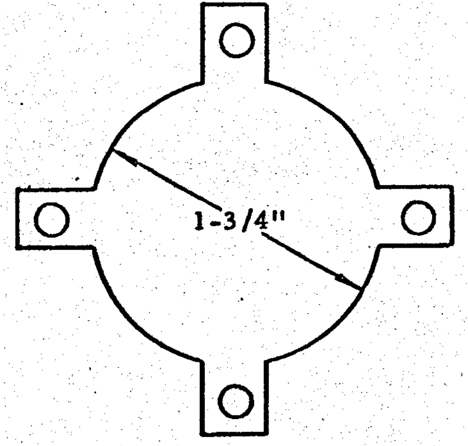
Tray 6
(Bottom Tray)

FIGURE 6

DISK-AND-DOUGHNUT TRAYS
(4-8D/D)



Doughnut



Disk

FIGURE 7

DIRECT CONTACT HEAT EXCHANGE LABORATORY COLUMN

(Run 4-7A, 11/18)

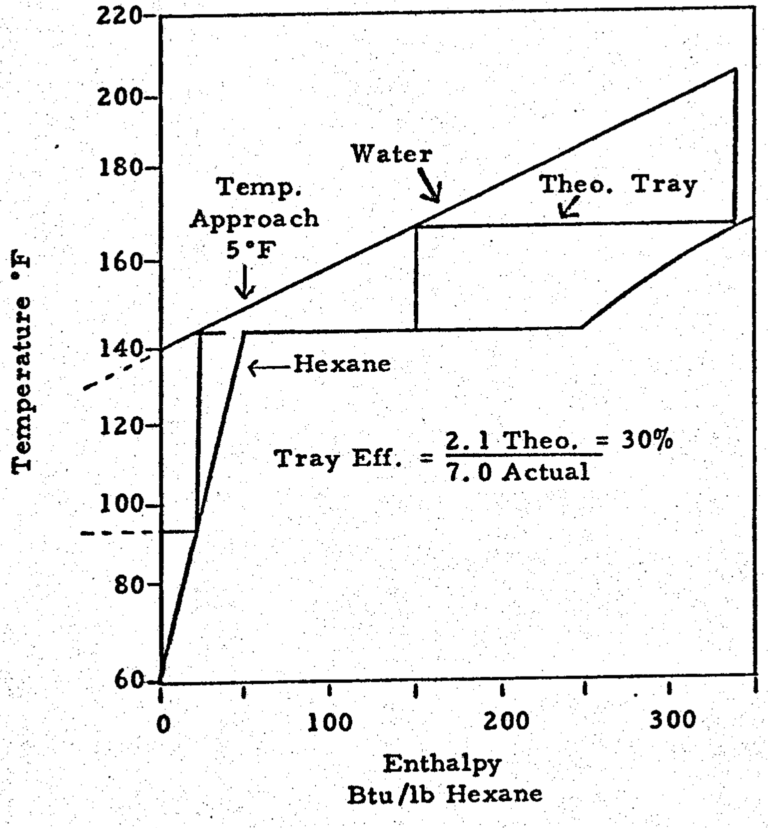


FIGURE 8

SOLUBILITY OF HYDROCARBONS IN WATER AND BRINES

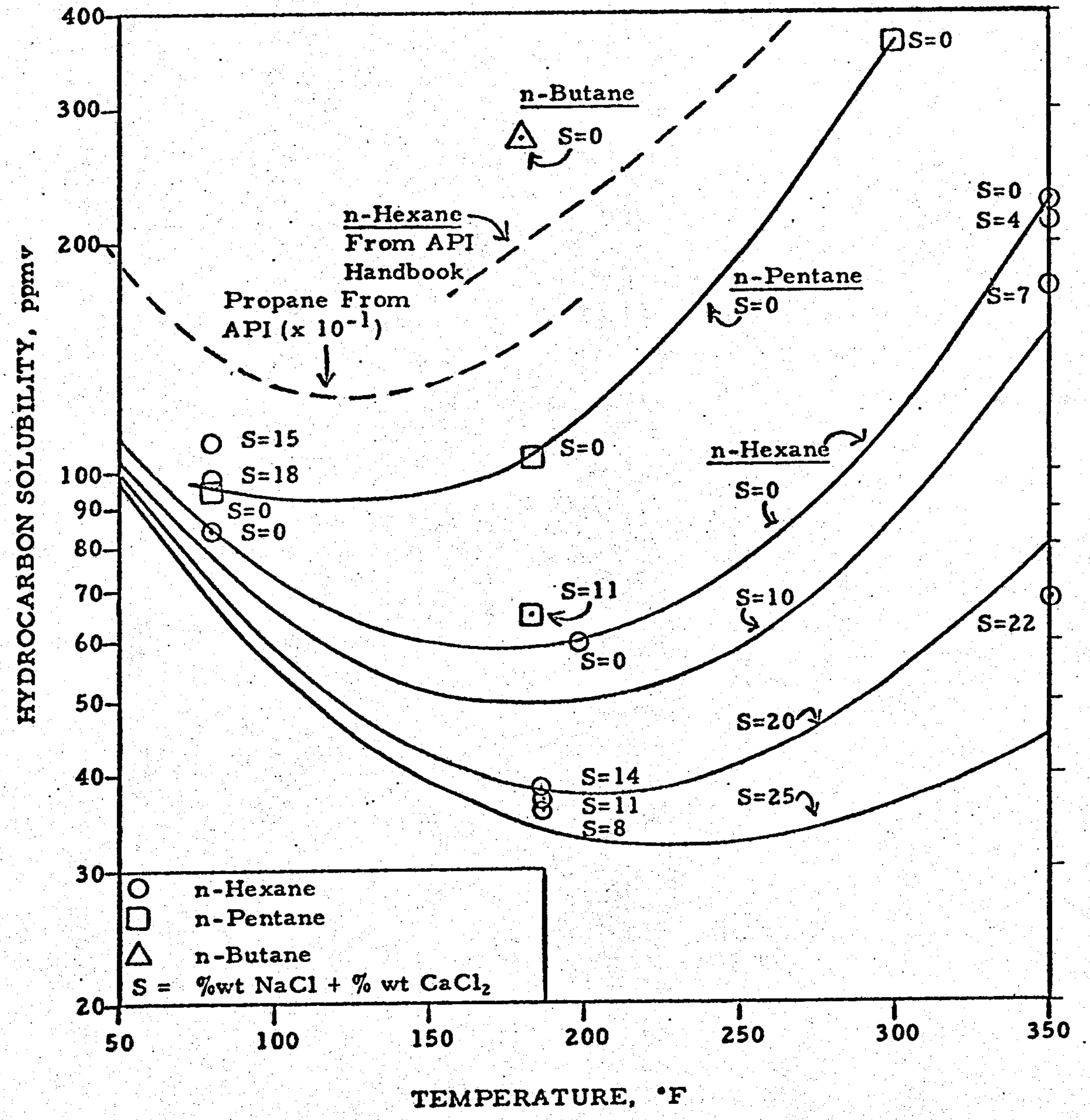


FIGURE 9

HEXANE IN
STRIPPER FEED

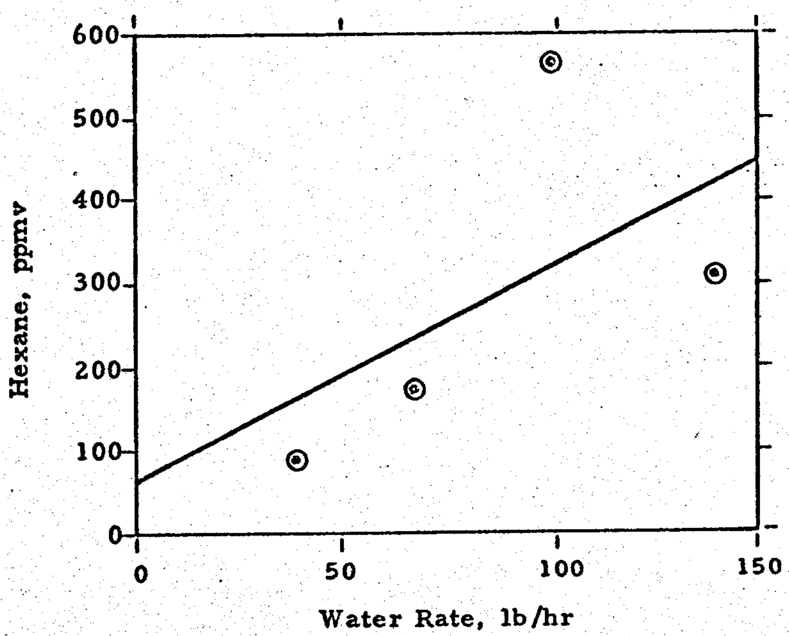


FIGURE 10

STRIPPING TESTS

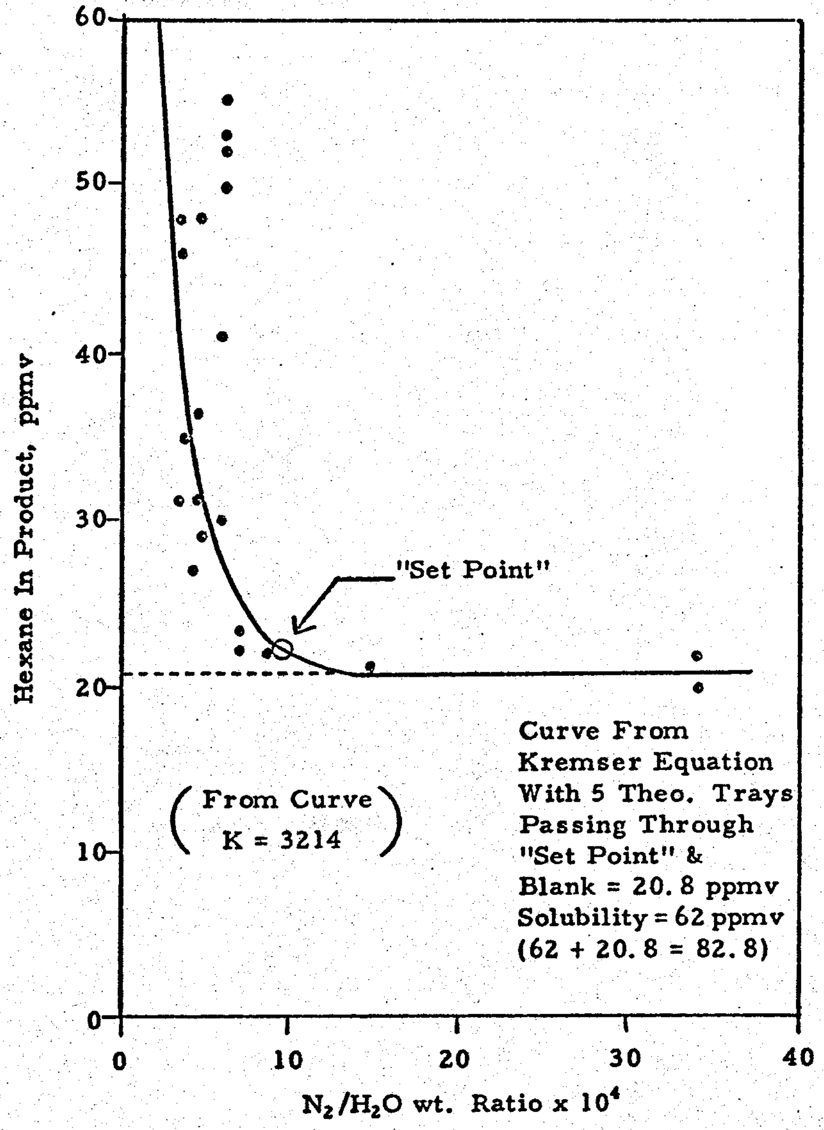


TABLE 1

OPERATION OF HEAT EXCHANGER COLUMN

(METRIC UNITS)

Identification	1-4	4-8				4-3					4-3A	4-0
Date	8/27	9/11	9/16	9/17	9/22	9/26	9/30	10/2	10/3	10/6	10/13	10/10
Flow, kg/hr												
H_2O (l) in	11.75	24.29	18.28	18.16	18.28	18.28	12.38	18.28	18.68	18.04	18.31	35.43
nC_6 (l) in	0.74	1.86	2.01	1.87	1.99	1.87	1.92	2.40	2.73	2.78	2.64	3.76
H_2O (v) out	0.17	0.58	0.56	0.49	0.46	0.35	0.24	0.41	0.41	0.36	0.35	0.49
Temperature, °C												
H_2O in	92	95	95	95	95	95	92	94	95	93	94	96
H_2O (l) out	74	69	67	67	66	69	63	65	64	62	64	73
nC_6 in	28	22	37	37	33	35	29	24	30	21	35	22
nC_6/H_2O out	77	88	84	83	82	80	75	77	77	75	74	75
Heat Balance, watts												
H_2O in	1260	2682	2019	2006	2019	2019	1325	2007	2116	1962	2006	3952
nC_6 in	61	149	179	166	163	164	161	193	231	219	212	301
Total	1321	2831	2198	2172	2182	2183	1486	2200	2347	2181	2218	4253
H_2O out	995	1899	1373	1381	1357	1435	893	1354	1368	1266	1340	2970
H_2O (v) out	126	428	414	360	337	260	179	299	300	265	259	355
nC_6 out	154	397	426	395	397	392	397	498	567	574	544	777
Loss	46	107	-15	36	91	96	17	49	112	76	75	151
Total	1321	2831	2198	2172	2182	2183	1486	2200	2347	2181	2218	4253
Wt. Ratio H_2O/nC_6	15.9	13.1	9.1	9.7	9.7	9.8	6.4	7.6	6.9	6.5	6.9	9.4
Insulation	No	No	Yes	Yes	Yes	Yes	Yes	Yes	Yes	Yes	No	No
Tray Efficiency, %	29	38	25	25	24	53	58	59	61	70	69	1.1*
Temperature Approach, °C	14	7	2	5	4	7	3	5	3	2	5	13

*Theoretical Plates

TABLE 1 (Continued -2)

OPERATION OF HEAT EXCHANGER COLUMN (METRIC UNITS)

Identification	4-IS									4-6		
Date	10/14	(A) 10/15	(B) 10/15	(A) 10/16*	(B) 10/16	10/17	10/21	10/22		(A) 10/28	(B) 10/28	11/3
Flow, kg/hr												
H_2O (l) in	18.1	42.9	18.22	54.07	52.03	55.80	54.09	64.36		51.23	64.37	55.43
nC_6 (l) in	2.41	6.20	2.59	11.52	11.08	8.75	11.63	12.46		8.63	11.16	11.59
H_2O (v) out	0.39	1.55	0.54	1.62	1.20	1.74	1.45	1.88		1.37	1.46	1.15
Temperature, °C												
H_2O in	94	97	95	98	97	95	96	97		97	96	97
H_2O (l) out	58	60	60	57	57	59	56	57		62	60	60
nC_6 in	28	28	32	29	31	26	20	23		26	27	26
nC_6/H_2O out	78	84	81	76	74	81	75	77		78	75	73
Heat Balance, watts												
H_2O in	1969	4858	2014	6253	5883	6164	6034	7233		8937	11137	9646
nC_6 in	202	519	220	952	945	718	887	1002		1096	1429	1461
Total	2171	5377	2234	7205	6828	6882	6921	8235		10033	12566	11107
H_2O out	1185	2901	1241	3486	3379	3733	3441	4152		5584	6787	5872
H_2O (v) out	283	1138	393	1186	876	1276	1059	1380		1555	1650	1304
nC_6 out	502	1310	544	2382	2287	1833	2401	2593		2778	3572	3687
Loss	201	28	56	151	286	40	20	110		116	557**	244
Total	2171	5377	2234	7205	6828	6882	6921	8235		10033	12566	11107
Wt. Ratio H_2O/nC_6	7.3	6.9	9.1	4.7	4.7	6.4	4.6	5.2		5.9	5.8	4.8
Insulation	No	Yes	Yes	Yes	Yes	Yes	Yes	Yes		Yes	Yes	Yes
Equilibrium Plates	Over 2	Over 3	Over 3	Over 2	Over 2	Over 2	Over 2	Over 2	Tray Efficiency, %	33	36	36
Temperature Approach, °C	0	1	1	0	0	0	0	0		3	2	2

*Changed Rotameters. **Relatively high heat loss indicates possible lack of equilibrium.

TABLE 1 (Continued - 3)

OPERATION OF HEAT EXCHANGER COLUMN

(METRIC UNITS)

Identification	4-7A									
	(A)	(B)		(A)	(B)	(C)	(A)	(B)	(A)	(B)
Date	11/13	11/13	11/14	11/17	11/17	11/17	11/18	11/18	11/19	11/19
Flow, kg/hr										
H_2O (l) in	65.86	60.69	64.06	64.70	65.50	62.02	66.86	65.22	62.31	61.96
nC_6 (l) in	9.93	9.93	9.96	9.79	10.53	11.16	11.25	11.78	12.46	4.87
H_2O (v) out	1.64	1.61	1.58	1.74	1.65	1.53	1.57	1.39	1.06	1.32
Temperature, °C										
H_2O in	97	97	96	97	97	97	97	96	94	95
H_2O (l) out	62	62	62	62	61	60	61	60	59	69
nC_6 in	29	31	22	21	21	22	16	17	19	21
nC_6/H_2O out	78	78	78	78	78	76	76	74	69	84
Heat Balance, watts										
H_2O in	7420	6811	7118	7302	7373	6992	7506	7281	6810	6872
nC_6 in	834	847	785	771	830	883	851	898	966	384
Total	8254	7658	7903	8073	8203	7875	8357	8179	7776	7256
H_2O out	4639	4289	4435	4551	4517	4215	4617	4440	4193	4793
H_2O (v) out	1200	1181	1161	1274	1210	1120	1149	1015	774	975
nC_6 out	2065	2066	2072	2038	2190	2311	2326	2427	2535	1031
Loss	350	122	235	210	286	229	265	297	274	457
Total	8254	7658	7903	8073	8203	7875	8357	8179	7776	7256
Wt. Ratio H_2O/nC_6	6.6	6.1	6.4	6.5	6.5	5.6	5.9	5.5	5.0	12.7
Insulation	No	Yes	Yes	Yes	Yes	Yes	Yes	Yes	Yes	No
Tray Efficiency, %	29	29	30	29	30	31	30	30	31	30
Temperature Approach, °C	2	2	2	4	2	2	3	2	2	8

TABLE 1 (Continued - 4)

OPERATION OF HEAT EXCHANGER COLUMN

(METRIC UNITS)

Identification	4-PR								
Date	(A) 11/21	(B) 11/21	(A) 11/24	(B) 11/24	11/25	12/3	(A) 12/4	(B) 12/4	12/5
Flow, kg/hr									
H_2O (l) in	63.12	58.58	65.16	63.42	66.92	58.11	64.93	64.41	64.93
nC_6 (l) in	9.49	11.88	4.20	4.15	13.74	12.50	13.12	14.02	15.08
H_2O (v) out	1.89	1.62	1.99	1.95	1.53	1.49	1.89	1.76	1.41
Temperature, °C									
H_2O in	96	96	98	98	95	98	98	98	96
H_2O (l) out	58	57	70	69	57	56	56	56	56
nC_6 in	19	23	25	27	22	18	15	15	19
nC_6/H_2O out	80	76	89	89	73	74	76	74	70
Heat Balance, watts									
H_2O in	7075	6570	7410	7228	7392	6615	7404	7345	7257
nC_6 in	732	951	342	343	1042	924	987	1059	1173
Total	7807	7521	7752	7571	8434	7539	8391	8404	8430
H_2O out	4090	3764	5133	4955	4304	3683	4122	4075	4099
H_2O (v) out	1386	1186	1466	1439	1120	1088	1326	1291	1029
nC_6 out	1986	2455	906	894	2821	2575	2716	2893	3078
Loss	345	116	247	283	189	193	227	145	224
Total	7807	7521	7752	7571	8434	7539	8391	8404	8430
Wt. Ratio H_2O/nC_6	6.6	4.9	15.5	15.3	4.87	4.65	4.95	4.60	4.31
Insulation	Yes	Yes	Yes	Yes	Yes	Yes	Yes	Yes	Yes
Equilibrium Plates	Over 2	Over 2	3	Over 3	Over 2	Over 2	Over 2	Over 2	Over 2
Temperature Approach, °C	0	0	11	9	0	0	Less than 1	Less than 1	0

TABLE 1 (Continued - 5)

OPERATION OF HEAT EXCHANGER COLUMN

(METRIC UNITS)

Identification	4-7B*							4-8D/D			
Date	12/9	12/12	12/18	12/19	12/22	12/24	12/29	3/15	3/16	3/17	3/19
Flow, kg/hr											
H_2O (l) in	62.66	64.93	63.83	64.93	64.58	65.16	64.99	46.53	46.53	18.79	18.67
nC_6 (l) in	12.25	11.16	11.16	11.16	11.16	5.28	4.96	7.68	7.68	2.52	2.84
H_2O (v) out	1.65	1.68	1.69	1.61	1.67	1.76	1.80	1.09	1.12	0.35	0.39
Temperature, °C											
H_2O in	97	97	97	97	97	97	98	97	97	96	95
H_2O (l) out	58	60	60	60	60	70	71	62	64	61	60
nC_6 in	20	13	14	13	12	15	20	26	27	26	24
nC_6/H_2O out	74	77	77	77	171	86	86	76	77	76	76
Heat Balance, watts											
H_2O in	7101	7342	7183	7342	7302	7368	7403	5245	5245	2089	2068
nC_6 in	938	822	833	826	825	396	389	630	635	207	229
Total	8039	8164	8016	8168	8127	7764	7792	5875	5880	2296	2297
H_2O out	4145	4400	4362	4413	4376	5189	5217	3292	3373	1297	1268
H_2O (v) out	1210	1230	1240	1183	1227	1294	1327	801	817	259	282
nC_6 out	2532	2318	2322	2312	2322	1113	1044	1591	1596	521	586
Loss	152	216	92	260	202	168	204	191	94	219	161
Total	8039	8164	8016	8168	8127	7764	7792	5875	5880	2296	2297
Wt. Ratio H_2O/nC_6	5.1	5.8	5.7	5.8	5.8	12.3	13.1	6.0	6.1	7.5	6.6
Insulation	Yes	Yes	Yes	Yes	Yes	Yes	Yes	No	No	No	No
Tray Efficiency, %	39	Over 29	30	31	30	29	29	25	25	28	28
Temperature Approach, °C	1	3	3	3	3	11	11	3	4	2	2

*Abbreviated data for operation of 4-7C modification included in section on stripping.

TABLE 1

OPERATION OF HEAT EXCHANGER COLUMN

(ENGLISH UNITS)

Identification	1-4	4-8				4-3					4-3A	4-0
Date	8/27	9/11	9/16	9/17	9/22	9/26	9/30	10/2	10/3	10/6	10/13	10/10
Flow, lb/hr												
H_2O (l) in	25.91	53.54	40.29	40.03	40.29	40.29	27.29	40.29	41.19	39.77	40.37	78.10
nC_6 (l) in	1.63	4.09	4.43	4.12	4.16	4.12	4.24	5.28	6.01	6.12	5.82	8.29
H_2O (v) out	0.38	1.28	1.24	1.08	1.01	0.78	0.54	0.90	0.91	0.80	0.78	1.07
Temperature, °F												
H_2O in	198	203	203	203	203	203	197.7	202.0	202.5	200.2	201.6	204.7
H_2O (l) out	165	156	152	153	150	156	146	149.4	148.0	143.0	147.6	163.7
nC_6 in	82	72	98	98	92	95	84	75	86	70	74	72.5
nC_6/H_2O out	170	190	184	182	180	176	166.5	171.0	170.0	166.4	165.4	167.3
Heat Balance, Btu/hr												
H_2O in	4301	9155	6890	6845	6890	6890	4522	6849	7223	6697	6847	13488
nC_6 in	209	507	611	568	557	560	551	660	787	747	722	1028
Total	4510	9662	7501	7413	7447	7450	5073	7509	8010	7444	7569	14516
H_2O out	3395	6480	4686	4713	4631	4899	3047	4620	4668	4321	4573	10137
H_2O (v) out	431	1462	1414	1230	1149	887	612	1021	1024	906	883	1212
nC_6 out	525	1354	1453	1347	1356	1339	1356	1700	1935	1958	1856	2653
Loss	159	366	-52	123	311	325	58	168	383	259	257	514
Total	4510	9662	7501	7413	7447	7450	5073	7509	8010	7444	7569	14516
Wt. Ratio H_2O/nC_6	15.9	13.1	9.1	9.7	9.7	9.8	6.4	7.6	6.9	6.5	6.9	9.4
Insulation	No	No	Yes	Yes	Yes	Yes	Yes	Yes	Yes	Yes	No	No
Tray Efficiency, %	29	38	25	25	24	53	58	59	61	70	69	1.1*
Temperature Approach, °F	26	13	3	9	7	13	5	9	6	3	9	24

*Theoretical Plates

TABLE 1 (Continued -2)

OPERATION OF HEAT EXCHANGER COLUMN

(ENGLISH UNITS)

Identification	4-IS									4-6		
		(A)	(B)	(A)	(B)					(A)	(B)	
Date	10/14	10/15	10/15	10/16*	10/16	10/17	10/21	10/22		10/28	10/28	11/3
Flow, lb/hr												
H_2O (l) in	39.8	94.6	40.16	119.2	114.7	123.02	119.25	141.88		112.95	141.90	122.20
nC_6 (l) in	5.32	13.67	5.70	25.39	24.43	19.28	25.65	27.48		19.02	24.61	25.56
H_2O (v) out	0.85	3.41	1.18	3.57	2.64	3.83	3.19	4.15		3.02	3.21	2.54
Temperature, °F												
H_2O in	201	207.3	203.2	209	207	203	204.7	206		206.4	205.0	206.0
H_2O (l) out	136	140.7	140.8	135	135	139	133.3	135		144.1	140.0	140.3
nC_6 in	83	83.3	89.5	84	87	79	68.7	74		79.0	80.0	78.0
nC_6/H_2O out	173	183.0	178	169	165.5	177	166.3	171		172.2	167.6	163.3
Heat Balance, Btu/hr												
H_2O in	6721	16580	6875	21342	20077	21038	20594	24687		19702	24552	21266
nC_6 in	689	1770	752	3250	3225	2449	3027	3421		2416	3150	3221
Total	7410	18350	7627	24592	23302	23487	23621	28108		22118	27702	24487
H_2O out	4044	9901	4237	11898	11531	12740	11745	14172		12311	14963	12946
H_2O (v) out	965	3885	1342	4048	2989	4355	3613	4709		3428	3638	2874
nC_6 out	1713	3370	1855	8130	7805	6256	8195	8849		6124	7875	8128
Loss	688	94	193	516	977	136	68	378		255	1226**	539
Total	7410	18350	7627	24592	23302	23487	23621	28108		22118	27702	24487
Wt. Ratio H_2O/nC_6	7.3	6.9	9.1	4.7	4.7	6.4	4.6	5.2		5.9	5.8	4.8
Insulation	No	Yes	Yes	Yes	Yes	Yes	Yes	Yes		Yes	Yes	Yes
Equilibrium Plates	Over 2	Over 3	Over 3	Over 2	Over 2	Over 2	Over 2	Over 2	Tray Efficiency, %	33	36	36
Temperature Approach, °F	0	2	2	0	0	0	0	0		5	3	3

*Changed Rotameters.

**Relatively high heat loss indicates possible lack of equilibrium.

TABLE 1 (Continued -3)

OPERATION OF HEAT EXCHANGER COLUMN **(ENGLISH UNITS)**

Identification	4-7A									
	(A)	(B)		(A)	(B)	(C)	(A)	(B)	(A)	(B)
Date	11/13	11/13	11/14	11/17	11/17	11/17	11/18	11/18	11/19	11/19
Flow, lb/hr										
H_2O (l) in	145.20	133.80	141.22	142.63	144.41	136.72	147.39	143.79	137.36	136.59
nC_6 (l) in	21.90	21.90	21.96	21.58	23.21	24.61	24.81	25.97	27.47	10.73
H_2O (v) out	3.61	3.55	3.49	3.83	3.64	3.37	3.46	3.06	2.34	2.92
Temperature, °F										
H_2O in	206.5	205.7	204.0	206.8	206.2	206.2	205.8	204.8	201.2	203.7
H_2O (l) out	144.0	144.5	143.0	144.1	141.6	140.0	141.6	139.8	138.1	155.5
nC_6 in	84.3	88.0	70.8	69.1	70.0	71.0	60.0	62.0	65.5	70.0
nC_6/H_2O out	173.2	172.7	172.0	173.1	172.0	168.8	168.8	165.0	156.2	184.1
Heat Balance, Btu/hr										
H_2O in	25342	23245	24293	24936	25164	23862	25619	24850	23241	23455
nC_6 in	2847	2891	2679	2633	2832	3015	2903	3064	3296	1309
Total	28189	26136	26972	27569	27996	26877	28522	27914	26537	24764
H_2O out	15843	14639	15135	15544	15415	14387	15759	15155	14311	16360
H_2O (v) out	4099	4030	3961	4349	4131	3821	3923	3465	2641	3328
nC_6 out	7052	7052	7071	6960	7474	7888	7939	8284	8653	3519
Loss	1195	415	805	716	976	781	901	1010	932	1557
Total	28189	26136	26972	27569	27996	26877	28522	27914	26537	24764
Wt. Ratio H_2O/nC_6	6.6	6.1	6.4	6.5	6.5	5.6	5.9	5.5	5.0	12.7
Insulation	No	Yes	Yes	Yes	Yes	Yes	Yes	Yes	Yes	No
Tray Efficiency, %	29	29	30	29	30	31	30	30	31	30
Temperature Approach, °F	4.5	4.5	4	7	4	3	5	4	3	15

TABLE 1 (Continued -4)

OPERATION OF HEAT EXCHANGER COLUMN

(ENGLISH UNITS)

Identification	4-PR								
	(A)	(B)	(A)	(B)			(A)	(B)	
Date	11/21	11/21	11/24	11/24	11/25	12/3	12/4	12/4	12/5
Flow, lb/hr									
H_2O (l) in	139.16	129.14	143.66	139.82	147.52	128.11	143.15	141.99	143.15
nC_6 (l) in	20.92	26.18	9.26	9.14	30.28	27.55	28.92	30.90	33.24
H_2O (v) out	4.16	3.57	4.38	4.30	3.38	3.28	3.99	3.89	3.11
Temperature, °F									
H_2O in	205.5	205.6	208.0	208.4	203.7	208.2	208.5	208.5	205
H_2O (l) out	135.5	134.4	157.9	156.9	134.0	132.8	133.2	132.8	132
nC_6 in	65.5	73.8	77.1	80.8	71.0	64.8	59.5	59.8	67
nC_6/H_2O out	176.7	168.7	192.0	192.2	163.3	165.0	169.0	165.7	158
Heat Balance, Btu/hr									
H_2O in	24148	22423	25290	24687	25229	22578	25271	25067	24768
nC_6 in	2500	3246	1167	1170	3558	3154	3369	3615	4005
Total	26648	25669	26457	25857	28787	25732	28640	28682	28773
H_2O out	13959	12846	17520	16924	14688	12570	14069	13907	13990
H_2O (v) out	4730	4047	5005	4914	3824	3714	4524	4405	3512
nC_6 out	6778	8378	3093	3053	9629	8788	9269	9873	10504
Loss	1181	398	839	966	646	660	778	497	767
Total	26648	25669	26457	25857	28787	25732	28640	28682	28773
Wt. Ratio H_2O/nC_6	6.6	4.9	15.5	15.3	4.87	4.65	4.95	4.60	4.31
Insulation	Yes	Yes	Yes	Yes	Yes	Yes	Yes	Yes	Yes
Equilibrium Plates	Over 2	Over 2	3	Over 3	Over 2	Over 2	Over 2	Over 2	Over 2
Temperature Approach, °F	0	0	19	17	0	0	Less than 1	Less than 2	0

TABLE 1 (Continued -5)

OPERATION OF HEAT EXCHANGER COLUMN

(ENGLISH UNITS)

Identification	4-7B							4-8D/D			
Date	12/9	12/12	12/18	12/19	12/22	12/24	12/29	3/15	3/16	3/17	3/19
Flow, lb/hr											
H_2O (l) in	138.14	143.15	140.71	143.15	142.38	143.66	143.28	102.57	102.57	41.43	41.15
nC_6 (l) in	27.00	24.61	24.61	24.61	24.61	11.65	10.94	16.94	16.94	5.55	6.25
H_2O (v) out	3.64	3.70	3.73	3.56	3.69	3.87	3.97	2.41	2.46	0.78	0.85
Temperature, °F											
H_2O in	207.4	207.0	206.2	207.0	207.0	207.4	208.4	206.5	206.5	204.1	203.5
H_2O (l) out	137.3	139.8	140.8	140.0	139.8	158.8	160.0	144.3	147.1	141.0	139.5
nC_6 in	67.3	55.0	56.5	55.5	54.5	59.5	68.3	79.2	81.0	79.0	74.5
nC_6/H_2O out	165.9	171.0	171.2	170.0	170.9	186.5	187.5	169.6	170.1	169.2	168.8
Heat Balance, Btu/hr											
H_2O in	24235	25057	24516	25057	24922	25146	25280	17902	17902	7131	7058
nC_6 in	3200	2806	2842	2818	2815	1351	1329	2151	2168	705	781
Total	27435	27863	27358	27875	27737	26497	26609	20053	20070	7836	7839
H_2O out	14148	15017	14888	15060	14936	17710	17816	11237	11512	4426	4328
H_2O (v) out	4128	4198	4232	4038	4187	4415	4530	2733	2790	884	964
nC_6 out	8641	7912	7924	7890	7924	3798	3566	5429	5446	1779	2000
Loss	518	736	314	887	690	574	697	654	322	747	547
Total	27435	27863	27358	27875	27737	26497	26609	20053	20070	7836	7839
Wt. Ratio H_2O/nC_6	5.1	5.8	5.7	5.8	5.8	12.3	13.1	6.0	6.1	7.5	6.6
Insulation	Yes	Yes	Yes	Yes	Yes	Yes	Yes	No	No	No	No
Tray Efficiency, %	39	Over 29	30	31	30	29	29	25	25	28	28
Temperature Approach, °F	2	5	5	5	5	19	19	6	8	3	3

TABLE 2

SUMMARY OF OPERATION OF HEAT EXCHANGER COLUMN

(METRIC UNITS)

Identification	1-4	4-8	4-3	4-0	4-IS	4-6	4-7A, B	4-PR	4-8D/D*
Number of Tests	1	4	6	1	8	3	17	9	4
Flow, kg/hr									
H_2O (l) in	11.7	18.1-24.3	12.4-18.7	25.4	18.1-64.4	51.3-64.4	60.7-66.9	58.1-66.9	18.7-46.5
nC_6 (l) in	0.7	1.9- 2.0	1.9- 2.8	3.8	2.4-12.5	8.6-11.6	4.9-12.5	4.1-15.1	2.5- 7.7
Temperature, °C									
H_2O (l) out	74	66-69	62-69	73	56-61	60-62	58-71	56-70	60-64
nC_6/H_2O out	77	82-88	74-80	75	74-84	73-78	74-87	70-89	76-77
Tray Efficiency, %	29	24-38	53-70	1.1**	Over 2 or 3**	33-36	29-39	Over 2 or 3**	25-28
Temperature Approach, °C	14	2-7	2-7	13	0-1	2-3	1-11	0-11	2-4

*Operation carried out in conjunction with a stripper.

**Equilibrium plates.

TABLE 2

SUMMARY OF OPERATION OF HEAT EXCHANGER COLUMN

(ENGLISH UNITS)

Identification	1-4	4-8	4-3	4-0	4-IS	4-6	4-7A, B	4-PR	4-8D/D*
Number of Tests	1	4	6	1	8	3	17	9	4
Flow, lb/hr									
H_2O (l) in	25.9	40.0-53.5	27.3-41.2	78.1	39.8-141.9	113.0-141.9	133.8-147.4	128.1-147.5	41.2-102.6
nC_6 (l) in	1.6	4.1- 4.4	4.1- 6.1	8.3	5.3- 27.5	19.0- 25.6	10.7- 27.5	9.1- 33.2	5.6- 17.0
Temperature, °F									
H_2O (l) out	165	150-156	143-156	164	133-141	140-144	137-160	132-158	140-147
nC_6/H_2O out	170	180-190	166-176	167	166-183	163-172	166-188	158-192	169-170
Tray Efficiency, %	29	24-38	53-70	1.1**	Over 2 or 3**	33-36	29-39	Over 2 or 3**	25-28
Temperature Approach, °F	26	3-13	3-13	24	0-2	3-5	2-19	0-19	3-8
Average Minimum	-	6	5	-	0	4	4	0	5

*Operation carried out in conjunction with a stripper.

**Equilibrium plates.

TABLE 3

HEXANE SOLUBILITY
(ppmv)

(Metric/English Units)

Temperature, °C (°F)	177 (350)			87 (188)			27 (80)		
$CaCl_2$, % w	0	4	8	0	4	8	0	4	8
NaCl - 0% w	-	212.9	-	-	-	35.8	83.9	-	-
NaCl - 7% w	170.3	-	-	-	37.2	-	-	-	109.8
NaCl - 14% w	-	-	68.3	38.4	-	-	-	97.8	-
Number of Values in Average	10	8	6	7	5	4	15	5	5
Standard Deviation	33.9	30.0	11.0	9.9	8.4	8.8	25.3	30.4	33.3

TABLE 4

NORMAL PENTANE SOLUBILITY
(ppmv)

(Metric/English Units)

Temperature, °C (°F)	149 (300)	84 (183)		27 (80)
$CaCl_2$, % w	0	0	4	0
NaCl - 0% w	360	104	-	95
NaCl - 7% w	-	-	65	-
Number of Values in Average	4	6	4	4

TABLE 5

COMPARISON OF SOLUBILITY IN WATER OF NORMAL HEXANE, NORMAL PENTANE AND NORMAL BUTANE

(Metric/English Units)

Hydrocarbon	nC_6			nC_5			nC_4
Temperature, °C (°F)	27 (80)	93 (200)	177 (350)	27 (80)	84 (183)	149 (300)	82 (180)
Concentration, ppmv	84	60	226	95	104	360	275
Mole Fraction ($x\ 10^5$)	1.2	.83	3.1	1.5	1.6	5.7	5.0
Number of Values in Average	15	8	2	4	6	4	1

TABLE 6

OPERATION OF STRIPPER

(METRIC UNITS)

DATE	1/22	1/23	1/26	1/28	1/29	2/2	2/3	2/4	2/5	2/6	2/10
HEAT EXCHANGER											
Identification	4-7C										
Flow, kg/hr											
H_2O in	46.36	45.81	25.97	62.01	46.31	46.99	47.13	46.54	64.82	63.87	64.23
nC_6 in	9.58	9.30	6.13	12.66	9.26	9.26	9.26	9.26	4.94	4.83	4.94
Temperature, °C											
C_6/H_2O (v) out	74	73	76	72	73	74	74	72	84	86	85
H_2O (l) out	61	60	64	60	61	61	61	60	71	72	71
STRIPPER											
N_2 Rate, kg/hr	.0266	.0266	.0266	.0266	.0266	.0199	.0139	.0202	.0269	.0266	.0268
H_2O Outlet Temp. °C	-	57	60	57	58	58	57	57	68	68	68
Weight Ratio of N_2/H_2O	.00057	.00058	.00089	.00043	.00057	.00042	.00030	.00043	.00041	.00042	.00042
Concentration of C_6, ppmv											
Feed	-	-	-	-	1,676	-	-	1,086	674	284	230
Stripped (Basis Feed)	844	1,275	580	1,513	1,514	667	531	967	270	283	248
Product	-	-	-	-	71	120	-	48	35	30	31
Loss	-	-	-	-	91	-	-	51	369	-29	-49
Stripping Efficiency, %	-	-	-	-	95.5	84.8	-	95.3	88.5	90.4	88.9

TABLE 6 (Continued -2)

OPERATION OF STRIPPER (METRIC UNITS)

DATE	2/11	2/12	2/13	2/17	2/18	2/19	2/23	2/24	2/25	2/26	2/27
HEAT EXCHANGER											
Identification	4-7C										
Flow, kg/hr											
H_2O in	63.82	60.10	64.96	64.96	49.71	46.00	64.86	47.54	64.46	46.31	46.31
nC_6 in	4.87	11.31	8.66	8.90	9.26	4.94	7.63	9.30	7.76	7.63	7.67
Temperature, °C											
C_6/H_2O (v) out	86	74	81	80	76	82	83	76	85	78	79
H_2O (l) out	71	59	64	63	59	68	66	59	70	62	63
STRIPPER											
N_2 Rate, kg/hr	.0194	.0266	.0202	.0209	.0205	.0265	.0266	.0266	.0274	.0266	.0266
H_2O Outlet Temp. °C	69	57	61	61	57	64	63	56	68	60	61
Weight Ratio of N_2/H_2O	.00030	.00044	.00031	.00032	.00041	.00056	.00041	.00056	.00042	.00057	.00057
Concentration of C_6, ppmv											
Feed	235	569	724	315	143	310	267	211	253	435	-
Stripped (Basis Feed)	294	428	869	452	176	265	256	169	219	628	761
Product	31	29	48	46	27	30	(est) 35	41	36	52	53
Loss	-90	112	-193	-183	-60	15	-24	1	-2	-245	-
Stripping Efficiency, %	90.5	93.6	94.8	90.8	86.7	89.8	88.0	80.5	85.8	92.4	93.5

TABLE 6 (Continued -3)

OPERATION OF STRIPPER

(METRIC UNITS)

DATE	3/2	3/4	3/9	3/11	3/12	3/15	3/16	3/17	3/19
HEAT EXCHANGER									
Identification	4-7C					4-8D/D			
Flow, kg/hr									
H_2O in	31.11	31.18	31.11	18.42	31.49	46.53	46.53	18.79	18.67
nC_6 in	5.82	3.41	3.74	2.53	5.99	7.68	7.68	2.52	2.84
Temperature, °C									
C_6/H_2O (v) out	77	81	77	78	77	77	77	76	76
H_2O (l) out	58	66	67	61	60	62	64	61	60
STRIPPER									*
N_2 Rate, kg/hr	.0266	.0267	.0211	.0269	.0206	.0267	.0267	.0637	.0635
H_2O Outlet Temp. °C	55	61	62	54	57	60	62	54	-
Weight Ratio of N_2/H_2O	.00086	.00086	.00068	.00146	.00066	.00057	.00057	.00339	.00340
Concentration of C_6, ppmv									
Feed	182	249	196	97	135	-	991	191	63
Stripped (Basis Feed)	132	237	152	79	165	-	938	142	45
Product	19	22	23	21	22	50	55	22	20
Loss	31	-10	21	-3	-52	-	-2	27	-2
Stripping Efficiency, %	87.4	91.5	86.9	79.0	88.2	-	94.5	86.6	69.2

*Ten-plate sieve tray stripper.

TABLE 6

OPERATION OF STRIPPER

(ENGLISH UNITS)

DATE	1/22	1/23	1/26	1/28	1/29	2/2	2/3	2/4	2/5	2/6	2/10
HEAT EXCHANGER											
Identification						4-7C					
Flow, lb/hr											
H_2O in	102.20	101.00	65.86	136.70	102.10	103.60	103.90	102.60	142.90	140.80	141.60
nC_6 in	21.11	20.50	13.52	27.92	20.41	20.41	20.41	20.41	10.90	10.64	10.90
Temperature, °F											
C_6/H_2O (v) out	165	164	169	162	164	165	165	162	184	186	185
H_2O (l) out	142	140	148	140	142	142	141	140	159	161	160
STRIPPER											
N_2 Rate, lb/hr	.0586	.0586	.0586	.0586	.0586	.0439	.0307	.0445	.0592	.0586	.0590
H_2O Outlet Temp. °F	-	135	140	135	136	137	135	134	154	154	154
Weight Ratio of N_2/H_2O	.00057	.00058	.00089	.00043	.00057	.00042	.00030	.00043	.00041	.00042	.00042
Concentration of C_6, ppmv											
Feed	-	-	-	-	1,676	-	-	1,086	674	284	230
Stripped (Basis Feed)	844	1,275	580	1,513	1,514	667	531	967	270	283	248
Product	-	-	-	-	71	120	-	48	35	30	31
Loss	-	-	-	-	91	-	-	51	369	-29	-49
Stripping Efficiency, %	-	-	-	-	95.5	84.8	-	95.3	88.5	90.4	88.9

TABLE 6 (Continued -2)

OPERATION OF STRIPPER

(ENGLISH UNITS)

DATE	2/11	2/12	2/13	2/17	2/18	2/19	2/23	2/24	2/25	2/26	2/27
HEAT EXCHANGER											
Identification	4-7C										
Flow, lb/hr											
H_2O in	140.70	132.50	143.20	143.20	109.60	101.40	143.00	104.80	142.10	102.10	102.10
nC_6 in	10.73	24.94	19.10	19.19	20.41	10.90	16.83	20.50	17.10	16.83	16.92
Temperature, °F											
C_6/H_2O (v) out	186	166	178	176	169	180	181	168	185	173	174
H_2O (l) out	160	139	147	146	139	154	150	139	158	144	145
STRIPPER											
N_2 Rate, lb/hr	.0427	.0587	.0445	.0461	.0452	.0584	.0587	.0587	.0603	.0586	.0587
H_2O Outlet Temp. °F	156	134	142	142	134	148	146	133	154	140	141
Weight Ratio of N_2/H_2O	.00030	.00044	.00031	.00032	.00041	.00056	.00041	.00056	.00042	.00057	.00057
Concentration of C_6, ppmv											
Feed	235	569	724	315	143	310	267	211	253	435	-
Stripped (Basis Feed)	294	428	869	452	176	265	256	169	219	628	761
Product	31	29	48	46	27	30	(est) 35	41	36	52	53
Loss	-90	112	-193	-183	-60	15	-24	1	-2	-245	-
Stripping Efficiency, %	90.5	93.6	94.8	90.8	86.7	89.8	88.0	80.5	85.8	92.4	93.5

TABLE 6 (Continued -3)

OPERATION OF STRIPPER

(ENGLISH UNITS)

DATE	3/2	3/4	3/9	3/11	3/12	3/15	3/16	3/17	3/19
HEAT EXCHANGER									
Identification	4-7C					4-8D/D			
Flow, lb/hr									
H_2O in	68.59	68.73	68.59	40.61	69.43	102.57	102.57	41.43	41.15
nC_6 in	12.84	7.52	8.24	5.58	13.21	16.94	16.94	5.55	6.25
Temperature, °F									
C_6/H_2O (v) out	170	177	177	173	170	170	170	169	169
H_2O (l) out	137	150	152	142	140	144	147	141	140
STRIPPER									*
N_2 Rate, lb/hr	.05871	.05882	.04658	.05941	.04552	.05888	.05882	.1404	.1401
H_2O Outlet Temp. °F	131	142	144	130	134	140	143	130	-
Weight Ratio of N_2/H_2O	.00086	.00086	.00068	.00146	.00066	.00057	.00057	.00339	.00340
Concentration of C_6, ppmv									
Feed	182	249	196	97	135	-	991	191	63
Stripped (Basis Feed)	132	237	152	79	165	-	938	142	45
Product	19	22	23	21	22	50	55	22	20
Loss	31	-10	21	-3	-52	-	-2	27	-2
Stripping Efficiency, %	87.4	91.5	86.9	79.0	88.2	-	94.5	86.6	69.2

*Ten-plate sieve tray stripper.

TABLE 7

HEXANE AS A FUNCTION OF WATER RATE

Water Rate, kg/hr (lb/hr)	18 (40)		31 (69)		45 (100)		64 (140)	
	Hexane in Feed ppmv	Water Temp. °F	Hexane in Feed ppmv	Water Temp. °F	Hexane in Feed ppmv	Water Temp. °F	Hexane in Feed ppmv	Water Temp. °F
	79	142	132	137	1,514**	142**	270	159
	142*	141*	237	150	667	142	283	161
	45*	140*	152	152	967	140	248	160
	-	-	165	140	176	139	294	160
	-	-	-	-	265	154	428	139
	-	-	-	-	169	139	869**	147**
	-	-	-	-	628	144	452	146
	-	-	-	-	761	145	256	150
	-	-	-	-	938	147	219	158
Average	89	141	172	145	571	144	306	154
		61°C		63°C		62°C		68°C

*Disc-and-Doughnut Heat Exchanger
**Omitted from average

CONCEPTUAL DESIGN

THERMODYNAMIC CYCLE STUDIES

Computer Program

DIRCON is an acronym for "DIRect CONtact Binary Cycle Process." This computer program calculates the important process parameters and state points of the direct contact binary cycle process. The program accepts as input data the thermal and chemical properties of the geothermal fluid, the design parameters of the plant equipment, and the properties and conditions of the working fluid. The program outputs the various state points, power output and auxiliary requirements and the geothermal fluid requirements. The results of the runs made as part of this study are summarized in Tables 8 through 14.

Thermodynamics

The physical and thermodynamic properties for hydrocarbon fluids required by the program are provided by the Modified Benedict-Webb-Rubin (MBWR) equation of state as formulated by Dr. Kenneth Starling, (6). The P-V-T data are derived directly from the equation of state. The enthalpy, entropy and fugacity data are derived from thermodynamic expressions based on the equation of state. The desired properties are calculated for either a pure hydrocarbon or any mixture of up to seven hydrocarbons. The program component library contains twenty-two compounds. The ideal gas enthalpy and entropy data which provide the basis for the enthalpy and entropy calculations are from the API Data Book.(5)

When water is considered alone, the P-V-T, enthalpy, entropy and vapor pressure data are calculated from equation provided in Keenan and Keyes "Thermodynamic Properties of Steam."(7) Mixtures of water and hydrocarbon are treated as ideal immiscible systems and the thermodynamic properties are determined accordingly.

The isentropic expansion in the turbine for a gaseous mixture of water and hydrocarbon is calculated from the MBWR for which equation coefficients exist for water in the gas phase only.

Mathematical Techniques

The MBWR equation is a sixth order polynomial of density with an exponential term. It, therefore, has roots at several densities. The highest and lowest roots correspond with the liquid and gas states respectively. The intermediate roots are not meaningful. The technique for determining the state point parameters involves solving the P-V-T equation for density and then using this value in the enthalpy, entropy, and fugacity equations. The density is found by a trial and error method which finds the vicinity of the desired root and a "false position" method for converging to this root. Vapor-liquid equilibria are determined by equating the fugacity of the vapor phase with that of the liquid phase. This is done by trial and error using the Newton method. There are occasionally convergence problems with these techniques but most of these have been eliminated by various "damping" methods.

Process Model

The basic process flow sheet is shown in Figure 16. The computer process model calculates the state points at each significant point in the process. It does an adiabatic flash of the incoming geothermal fluid at the contactor pressure. The model then determines the heat and material balance for the process which provides the geothermal fluid exit conditions and the amount of steam added to the hydrocarbon working fluid. The model also determines the internal energy consumption and the power production.

Program Organization

The program code is written in Fortran IV as implemented on the Burroughs 5700 computer. The code is organized as a driver program and twenty-two subroutines. There are approximately 1,050 Fortran statements.

Cycle Optimization

The optimum referred to in cycle optimization is not easy to define. It is not simply maximum thermodynamic efficiency based on heat available from the brine or on heat extracted from the brine. It is more nearly expressed as the minimum cost per unit of delivered power; but determining accurate costs for each proposed cycle would be difficult, and the optimum might not be consistent with the maximum utilization of the geothermal resource.

Economic studies of the production of electric power from geothermal brines indicate that brine production costs are a very important factor in the delivered cost of power. Furthermore, maximizing the power produced per unit of brine is also consistent with maximum resource utilization. Therefore, power produced per unit of brine was used to measure cycle performance.

Power production is affected by a large number of variables. Those used as input to the computer program, and the range covered for each are:

Hydrocarbon - isobutane
normal butane
isopentane
normal pentane
normal hexane
isopentane/normal hexane 60/40 (mol %)
butane/pentane/hexane 25/50/25
isobutane/hexane 50/50
n-butane/i-pentane 50/50

Condensing Temperature - 35 to 54°C (95 to 130°F)

Contactor Pressure Drop - 21 kPa (3 psi)

Contactor Temperature Approach - 2.8 to 5.6°C (5 and 10°F)

Operating Pressure - 480 to 4,928 kPa (55 to 700 psig)

Expander Inlet Temperature - 65 to 163°C (150 to 325°F)

Expander Efficiency - 0.85

Pump Efficiency - 0.75 and 0.80

Brine Flash Pressure - 463 to 1,515 kPa (52.4 to 205 psig)

Brine Specific Heat - 1.0

Brine Inlet Temperature - 110 to 182°C (230 to 360°F)

Cooling Water Temperature Approach - 5.6°C (10°F) (not part of program)

Since brine inlet temperature and condensing temperature are normally set by geothermal field conditions, it is convenient to set these variables for a series of cycle study runs. The major remaining variables are hydrocarbon working fluid, operating pressure, and expander inlet temperature.

The results for a series of runs with isopentane at 825 kPa (105 psig) operating pressure are shown in Figure 11. Gross power is the power produced by the expander-generator set and includes an 0.85 efficiency for the expander and 0.98 efficiency for the generator. The gross power per unit of hydrocarbon increases with expander inlet temperature from 15,870 kwh/M kg (7,200 kwh/10^6 lb) at 102°C (115°F) to 33,070 kwh/M kg (15,000 kwh/10^6 lb) at 149°C (300°F). If, however, the gross power is expressed in terms of units of brine the gross power decreases with increasing expander inlet temperature. Gross power per unit of brine decreases from 12,350 kwh/M kg (5,600 kwh/10^6 lb) at 102°C (115°F) to 8,380 kwh/M kg (3,800 kwh/10^6 lb) at 149°C (300°F). This behavior can be explained by plotting the series of cycles on a temperature/enthalpy diagram such as Figure 12. Point 1 represents the liquid hydrocarbon entering the direct contact heat exchanger; point 2 the boiling point in contact with brine; point 3 the dew point, that is the point at which all of the hydrocarbon has been vaporized; point 4 the hydrocarbon and water vapor entering the expander inlet for this run, or the direct contact heat exchanger outlet; point 5 the expander outlet, or the condenser inlet; point 6 the place within the condenser where all water vapor has been condensed. The cycle is completed by condensing the remaining hydrocarbon vapor and returning to point 1. Point 7 represents the hot brine entering the top of the direct contact heat exchanger column. The enthalpy of the brine is taken as equal to the enthalpy of the working fluid at the expander inlet so that heat transferred to the working fluid equals heat transferred from the brine. Point 8 is the minimum temperature approach, or pinch point. Point 9 is the cold brine leaving the column. The slope of the brine line is a measure of the brine to hydrocarbon ratio. A steep slope corresponds to a low brine to hydrocarbon ratio, and a low brine exit temperature. The cycle with the highest expander inlet temperature shows the highest expander work per unit weight of hydrocarbon, as measured by the enthalpy difference between points 4 and 5. This is achieved, however, at the cost of a high brine exit temperature, that is a high brine to hydrocarbon ratio. The cycle with the lowest expander inlet temperature, on the other hand, shows a lower value for expander work per unit weight of hydrocarbon but also a lower brine to hydrocarbon ratio. For this example, the maximum work per unit weight of brine occurs at the lowest expander inlet temperature consistent with complete vaporization of the hydrocarbon.

Maximum power per unit weight brine does not always coincide with minimum expander inlet temperature, for instance for the case of isobutane at 4,238 kPa (600 psig) operating pressure, Figure 13. In this case the gross power per unit weight hydrocarbon reaches a maximum at 147°C (296°F) and the gross power per unit weight brine reaches a maximum at 144°C (292°F). Again, the behavior can be explained by reference to temperature/enthalpy diagrams for the cycles, Figure 14. In this case the expander work for the low expander inlet temperature is very much less than that for the higher temperatures - enough less so that the low brine to hydrocarbon ratio cannot compensate. At the highest expander inlet temperature shown, the increased expander work is offset by the higher brine to hydrocarbon ratio, as compared to a somewhat lower temperature. The result is that maximum power is achieved at an intermediate expander inlet temperature.

The procedure outlined above was repeated for various operating pressures for both isopentane and isobutane. Figure 15 is a plot of maximum power per unit weight brine as a function of operating pressure for the two hydrocarbons, all for the same brine inlet and condenser temperatures. For these conditions maximum net power, 10,400 kwh/10^6 kg brine (4,700 kwh/10^6 lb brine), is reached with isopentane at 1,067 kPa (140 psig) operating pressure. The maximum net power with isobutane, 9,260 kwh/10^6 kg brine (4,200 kwh/10^6 lb brine), occurs at an operating pressure of 4,238 kPa (600 psig). Assuming that the wellhead pressure and reinjection pressure for the brine is 1,067 kPa, the net power would be further reduced by the power required for pumping the brine from wellhead to operating pressure; shown by the dotted line in Figure 15. Isopentane would be the working fluid of choice where the brine is available at 172°C and 825 kPa, and the condensing temperature is 38°C.

Similar studies for other operating conditions will lead to the choice of other hydrocarbons as working fluids. The choice may in some cases, be modified by other considerations such as the pressure required for brine reinjection.

In an attempt to get a better fit between the brine cooling curve and the hydrocarbon heating curve, wide boiling range hydrocarbon mixtures were tried as working fluids (see Table 14). Although it was possible to achieve a low brine outlet temperature, the work done by the expander was very low and as a consequence the net power production per unit of brine was low. This behavior is probably characteristic of wide boiling range working fluids since the amount of enthalpy change in the expander is limited by the condensation of the heavier components of the mixture.

Single component working fluids give optimum results where a compound with the desired properties is available. Where the desired properties lie between those of two compounds of a series, a mixture of the two can be expected to give better results.

NONCONDENSABLE GASES

Geothermal brines contain noncondensable gases. The quantity and composition of these gases varies with the location, but they generally consist of carbon dioxide with minor amounts of hydrogen sulfide, nitrogen, methane and ammonia. The gases produced with the brine are apt to be completely dissolved in the brine at bottom hole conditions but separate readily with pressure reduction or flashing.

Since complete and accurate data are lacking for composition, equilibrium K's, and solubilities, it is difficult to predict the fate of the noncondensable gases within the system. It is reasonable to assume that the gases will be essentially completely stripped from the brine in the direct contact heat exchanger. Calculations indicate, however, that the noncondensables are relatively soluble in hydrocarbons at the conditions occurring in the condenser. The noncondensable concentration will tend to build up in the working fluid loop. The level of noncondensables can be allowed to increase until it is possible to remove them by venting gas from the condenser, the amount to be vented depending on the concentration.

Hydrocarbon can be recovered from the vented gas by any of a number of well known methods; compression and cooling to condense the hydrocarbon as liquid, selective absorption of the carbon dioxide and hydrogen sulfide with a solvent such as diethanolamine, selective adsorption of carbon dioxide with molecular sieves, or condensation of the vented stream by compression and refrigeration followed by fractional distillation. The use of any of these methods would add significantly to the capital and operating costs of the unit. These costs could very probably be borne if the power generation process is otherwise financially attractive. It would appear, however, to be better practice to remove the noncondensables from the brine ahead of the direct contact heat exchanger.

In the process used for the cost estimate the noncondensable gases are removed from the incoming brine by flashing. The small amount of steam produced by flashing, containing essentially all of the noncondensables, is used to preheat and vaporize working fluid in shell and tube heat exchangers. Thus by the use of a small amount of relatively inexpensive equipment, it is possible to avoid the problem of noncondensable gases in the working fluid loop.

PARTICULATES IN EXPANDER INLET VAPOR

Scrubbers have been field tested on steam from flashed geothermal brines. Although the tests were for steam rather than hydrocarbon vapor containing steam, the results indicate that satisfactory scrubbers are available as state of the art.

CONTACTOR INTERNALS

The laboratory tests have shown that a wide variety of internals would be satisfactory for use in the direct contact heat exchange column. The choice will depend on field testing and may well be site specific. If field tests indicate heavy scaling the choice will probably be an open design such as disk-and-doughnut trays or shed trays. Sieve trays would probably be preferable for low scaling conditions. Shed trays have been used for purposes of cost estimating, although the choice of internals will not have a major impact on total plant cost.

WORKING FLUID LOSS IN REJECTED BRINE

The brine leaving the direct contact heat exchanger will contain hydrocarbon working fluid. Although normally considered insoluble, a small amount of hydrocarbon will be dissolved in the brine. In addition, there can be entrained hydrocarbon. Factors such as the large difference in density between the hydrocarbon and the brine and the low viscosities of both fluids make it possible to design for essentially no entrainment of hydrocarbon in the brine leaving the column, as confirmed by laboratory test results. The problem is, therefore, the dissolved hydrocarbon.

The solubility of the hydrocarbons decreases with increasing molecular weight and is a function of temperature. Also, the solubility is substantially reduced by the presence of dissolved salts. For the process design used in the cost estimate, the solubility of the pentane working fluid based on our laboratory data is about 100 ppmv or 62 ppmw. If none of this hydrocarbon is recovered, the loss for the 50 MWe plant would be 260 kg/hr (570 lb/hr) or 0.0052 kg/kwh (0.0114 lb/kwh). Literature values for the solubility of pentane in water are higher than our laboratory values and would give estimated losses of 680 kg/hr (1,500 lb/hr) or 0.014 kg/kwh (0.030 lb/kwh). To be conservative, the higher figures were used in the cost estimate.

Although the hydrocarbon loss could be tolerated from the standpoint of cost, the hydrocarbon can be recovered. The industrial processes usually used for this purpose are steam stripping, gas stripping, and adsorption on activated carbon. Stripping is efficient for the low molecular weight hydrocarbons used as working fluids and is less expensive than adsorption with carbon. Since reheating the brine is out of the question, steam stripping would require operation under vacuum. The exit brine temperature of 78°C (172°F) would correspond to an operating pressure of about 41 kPa (6 psia). A pump to restore the brine pressure to the column operating pressure would require a 1,640 kw driver, that is about 3.3% of the net power output of the plant. The power required for gas stripping is very much less, so gas stripping is the method of choice.

The laboratory work on stripping hexane from water shows that vapor-liquid equilibrium K values should be experimentally determined. Equilibrium K values based on solubility and vapor composition data are unreliably high for the hexane case; 93,000 compared to 3,214 from the laboratory stripping tests. Applying the same ratio of solubility K to laboratory K would give a K of about 2,000 isopentane and the conditions shown in the flow sheet of Figure 16.

Assuming five equilibrium stages and 95% recovery, the corresponding stripping gas rate for the 50 MWe plant would be 4,250 m^3/hr (16°C, 1 Atm.) (150,000 scfh). This relatively small volume of gas could be recycled through a stripper-condenser loop with a 25 KWe compressor. The hydrocarbon concentrations are so low that cooling the stripper offgas to 38°C (100°F) would not condense the hydrocarbon. The hydrocarbon is removed from the stripping gas and recovered by means of an activated carbon adsorption column. A flow sheet for the system is shown in Figure 17.

A preliminary estimate gives a plant cost of $400,000 for the stripper unit. Since accurate design data are not available, this cost should be considered approximate. A summary of the estimated plant cost is as follows:

Major Equipment	
Stripper	$ 18,800
Carbon Columns (2), Including Carbon	101,400
Accumulators (2)	8,000
Condensers (2)	5,000
Compressor	8,000
	$141,200
Plant Cost (x 2.8)	$400,000

Assuming an overall hydrocarbon recovery of 85% and a value of $0.06/lb for hydrocarbon, the annual savings would be $570,000. Estimated utilities and maintenance costs are $40,000/yr; no additional operating labor would be required. The estimated gross savings are, therefore, $530,000/yr. Since the simple before-tax-payout is less than one year, hydrocarbon recovery would appear to be attractive. In the absence of better data, however, this conclusion is tentative.

FIGURE 11

POWER - ISOPENTANE 105 PSIG

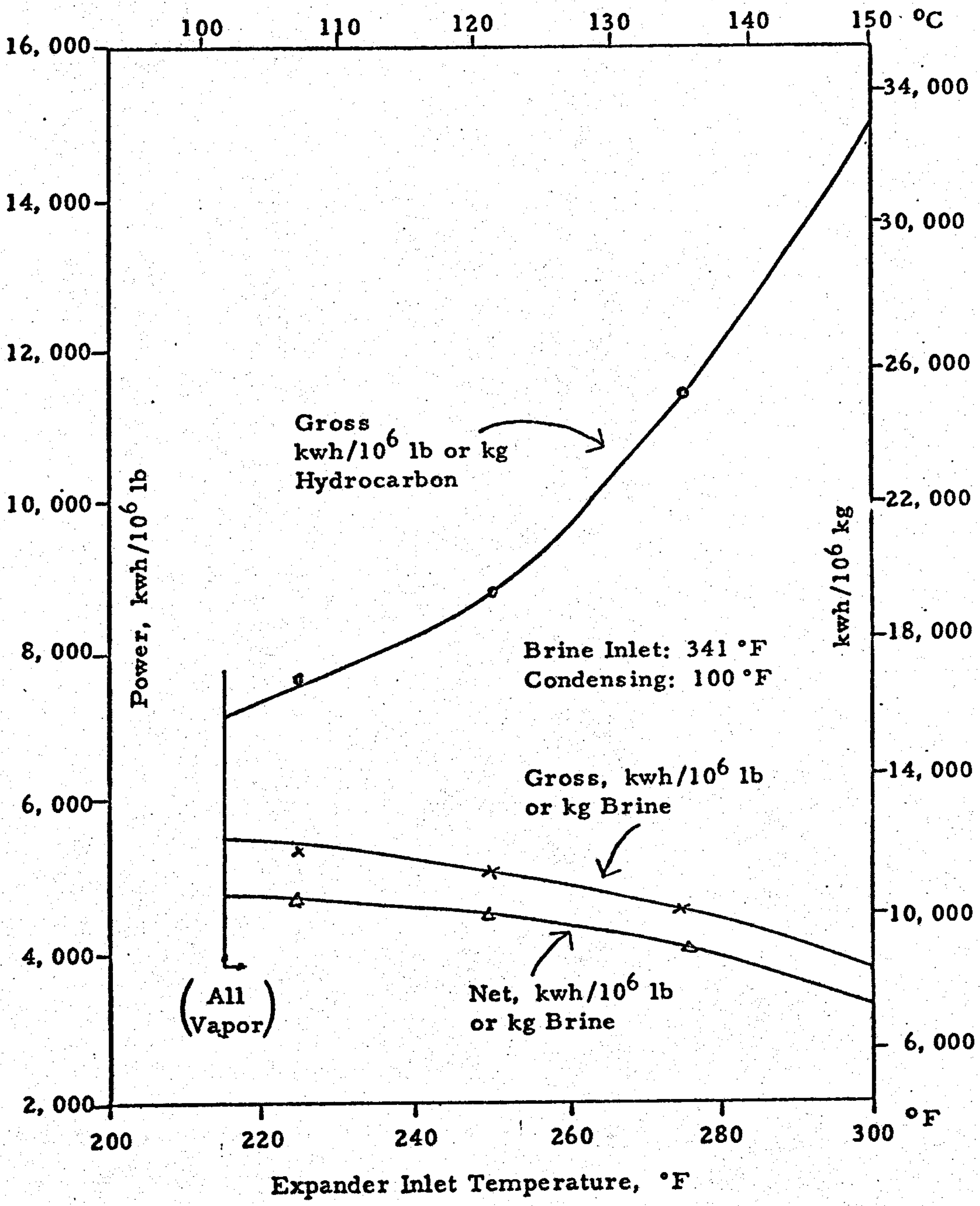


FIGURE 12

THERMODYNAMIC CYCLES - ISOPENTANE 105 PSIG

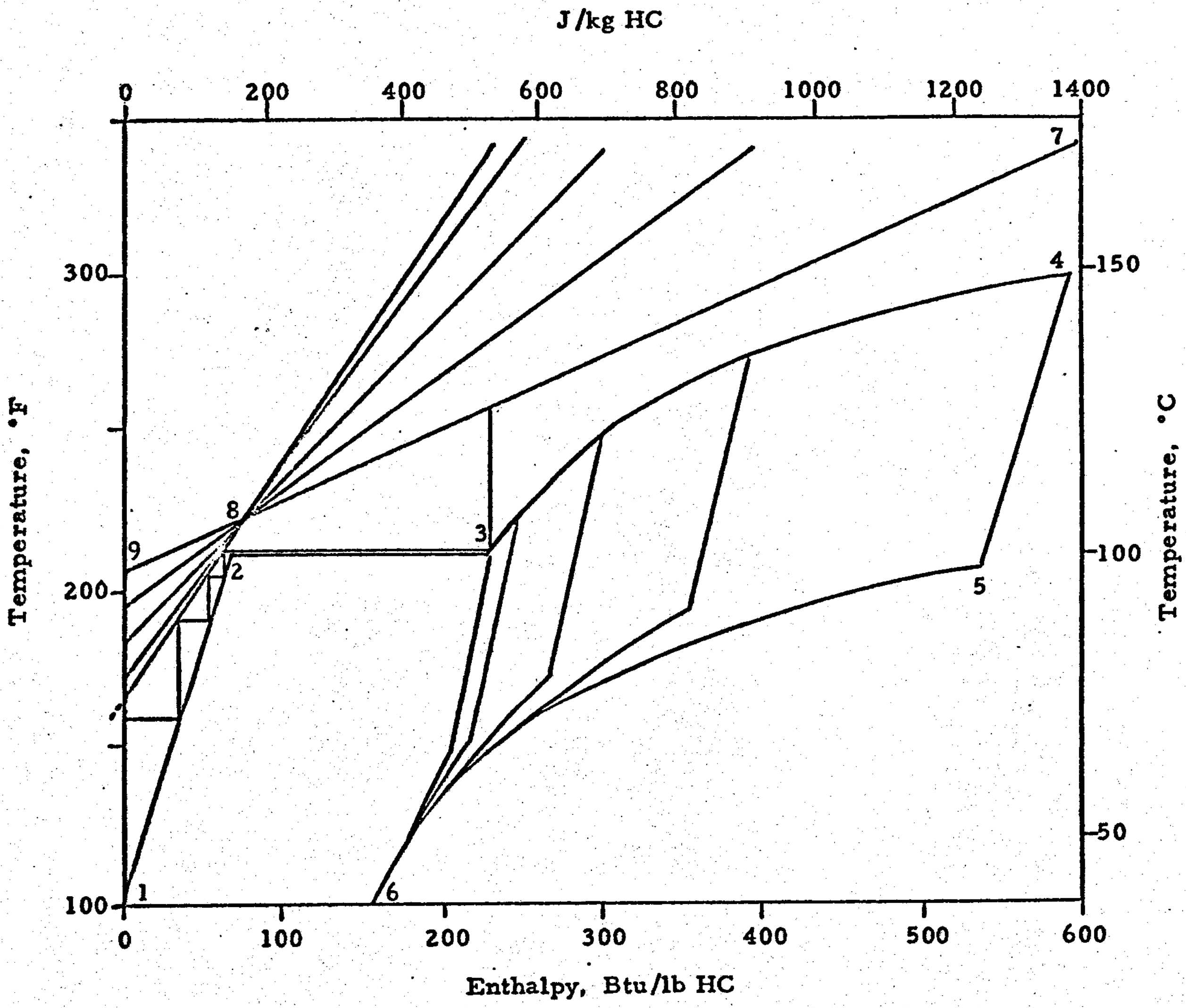


FIGURE 13

POWER – ISOBUTANE 600 PSIG

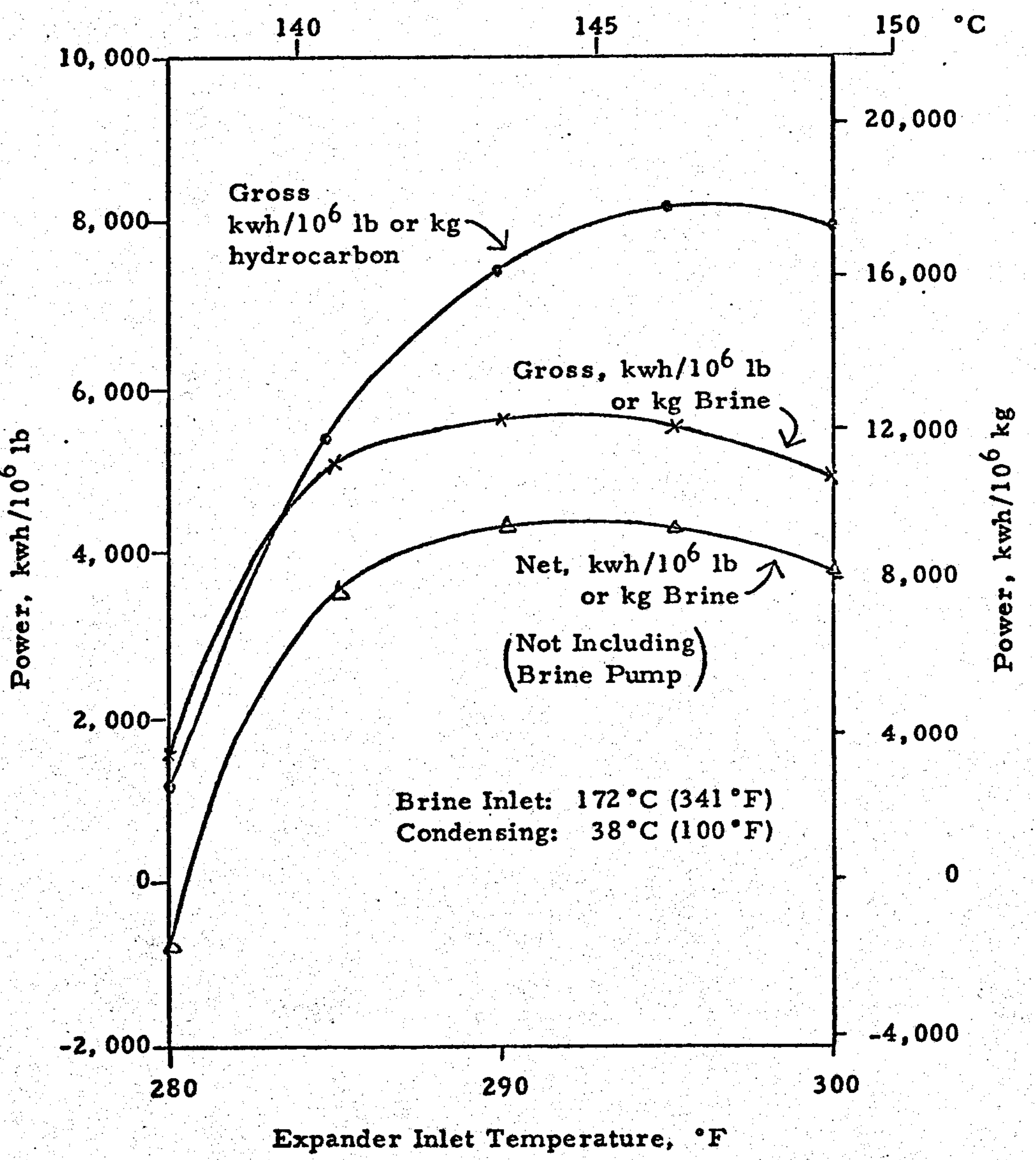


FIGURE 14

THERMODYNAMIC CYCLES - ISOBUTANE 600 PSIG

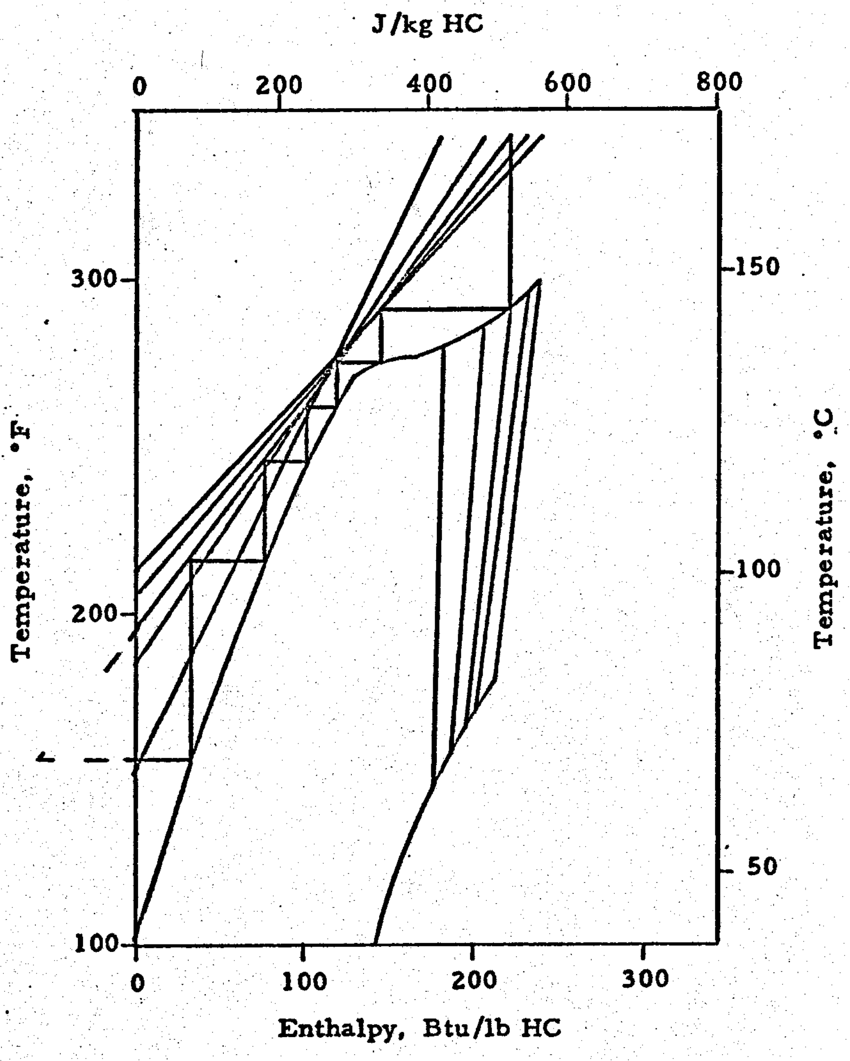
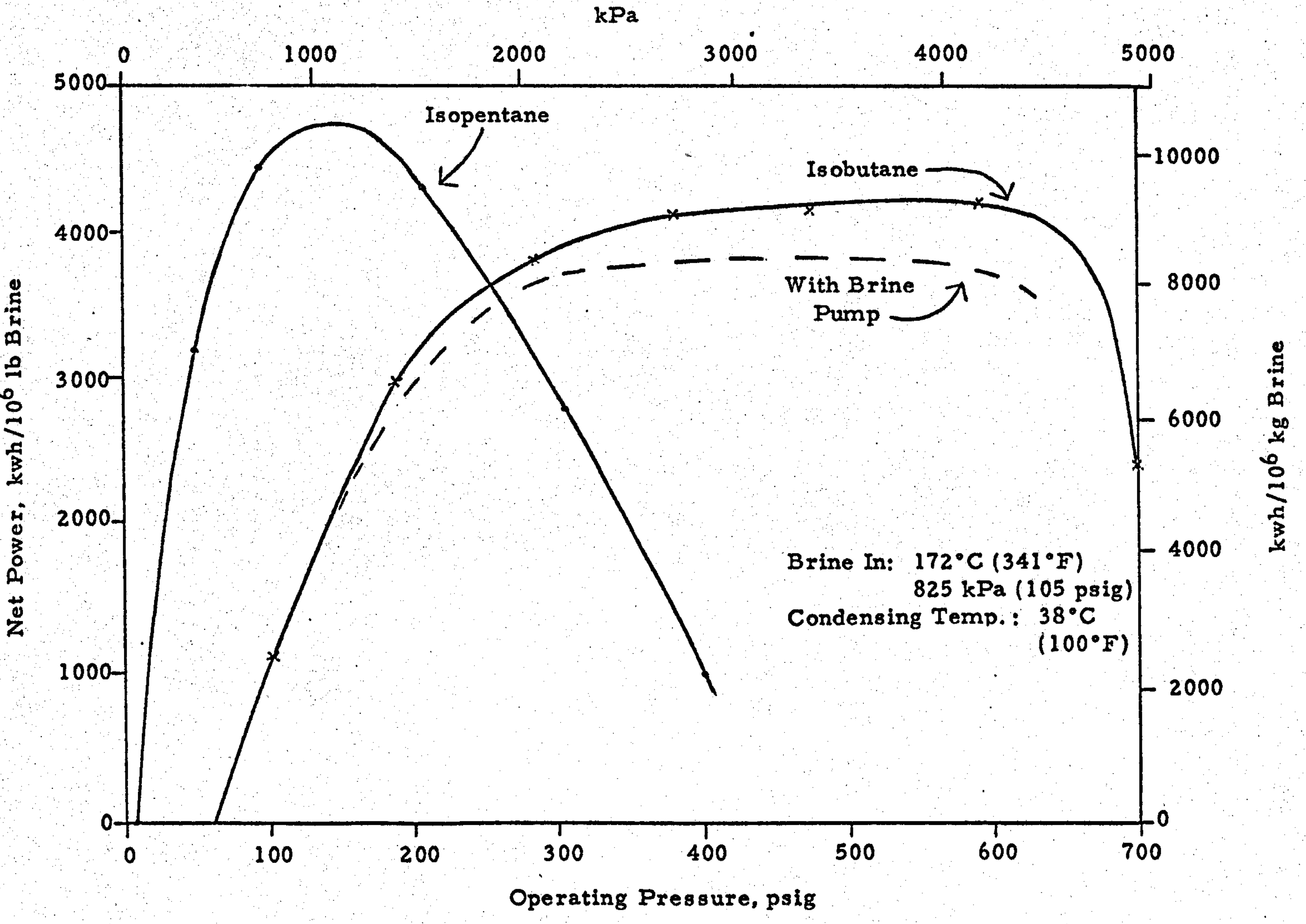
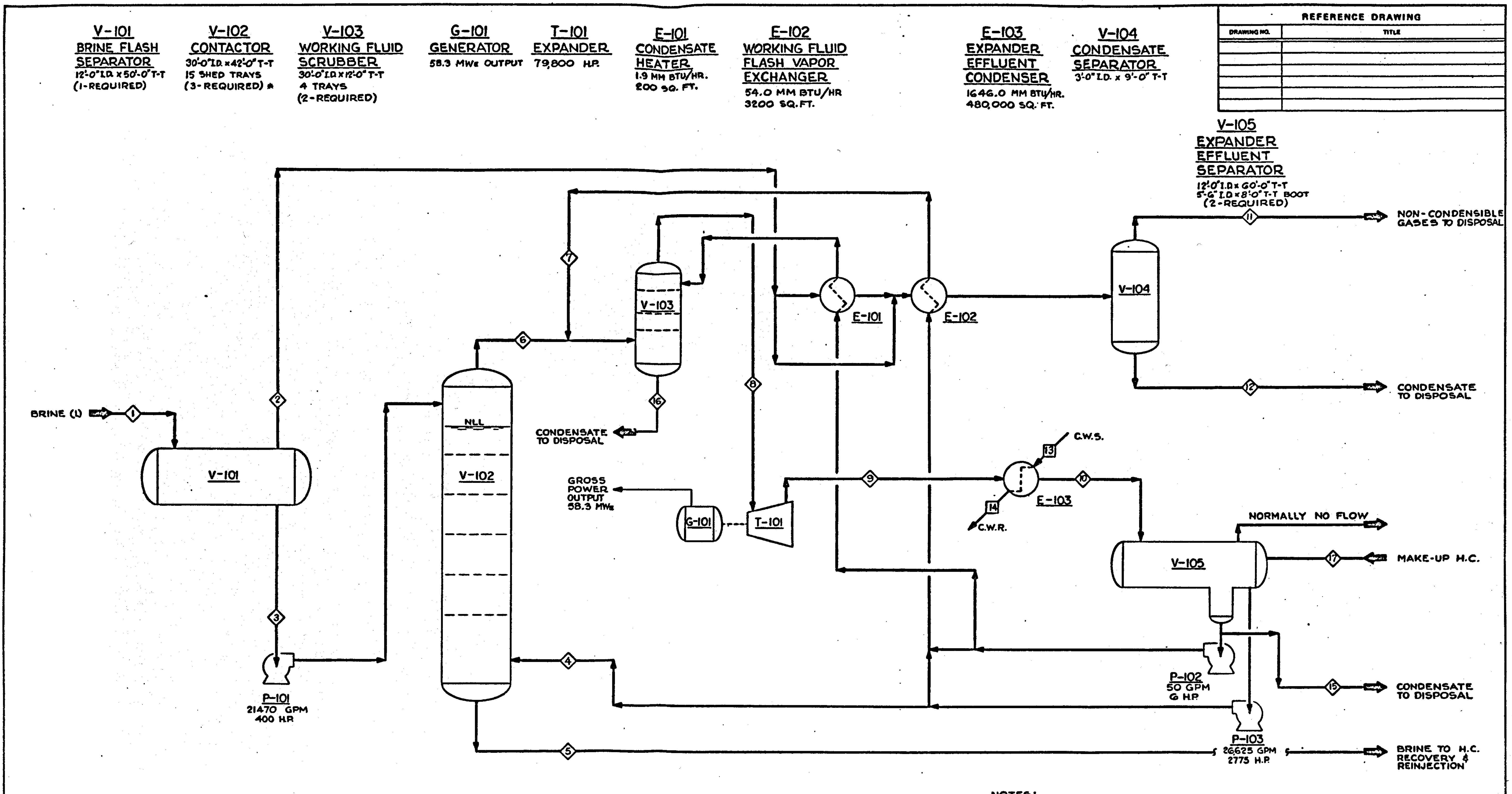


FIGURE 15

MAXIMUM NET POWER VS. PRESSURE



Brine In: 172°C (341°F)
825 kPa (105 psig)
Condensing Temp.: 38°C
(100°F)



* 2 REQUIRED FOR NORMAL OPERATION - 1 SPARE

	①	②	③	④	⑤	⑥	⑦	⑧	⑨	⑩	⑪	⑫	⑬	⑭	⑮	⑯	⑰
TOTAL M LB/HR	9584.2	55.3	9528.9	7677.7	9215.6	7991.0	233.3	8224.3	8224.3	8224.3	1.8	53.5	123854	123854	298.8	16	1.5
H_2O M LB/HR	9582.1	53.2	9528.9	8.8	9214.1	323.6	9.4	333.0	333.0	333.0	<0.1	53.2	123854	123854	298.8	16	-
HC M LB/HR	-	-	-	7668.9	1.5 (2)	7667.4	223.9	7891.3	7891.3	7891.3	-	-	-	-	-	-	1.5
$CO_2 + H_2S$ M LB/HR	2.1	2.1	-	-	-	-	-	-	-	-	1.8	0.3	-	-	-	-	-
TEMP. °F/PR. PSIG	360/129+	355/129	355/129	110/-	172/140	230/129	230/129	230/129	157/12	110/10	175/125	175/125	90/-	1033/-	110/10	230/129	110/10
PHASE	L	V	L	L	L	V	V	V	V	L	V	L	L	L	L	L	L

NOTES:
(1.) BRINE T.D.S. 15000 PPM
(2) EQUILIBRIUM SOLUBILITY OF ICS IN ICS-WATER SYSTEM AT ITS TOTAL PRESSURE.

FIGURE 16

MARK	REVISION	DR.	CH	DATE
A	ISSUED FOR APPROVAL	J.H.P.		

THE BEN HOLT CO.

TITLE:
DIRECT CONTACT POWER PLANT
50 MWe EXPORT POWER
HEBER FIELD

CUSTOMER: ERDA

DRAWN BY: JOE POPE
SCALE:
APPROVED:
DATE:
DRAWING NO. 7520-D-3201
REV. A

FIGURE 17

STRIPPING UNIT FLOW SHEET
(Preliminary)

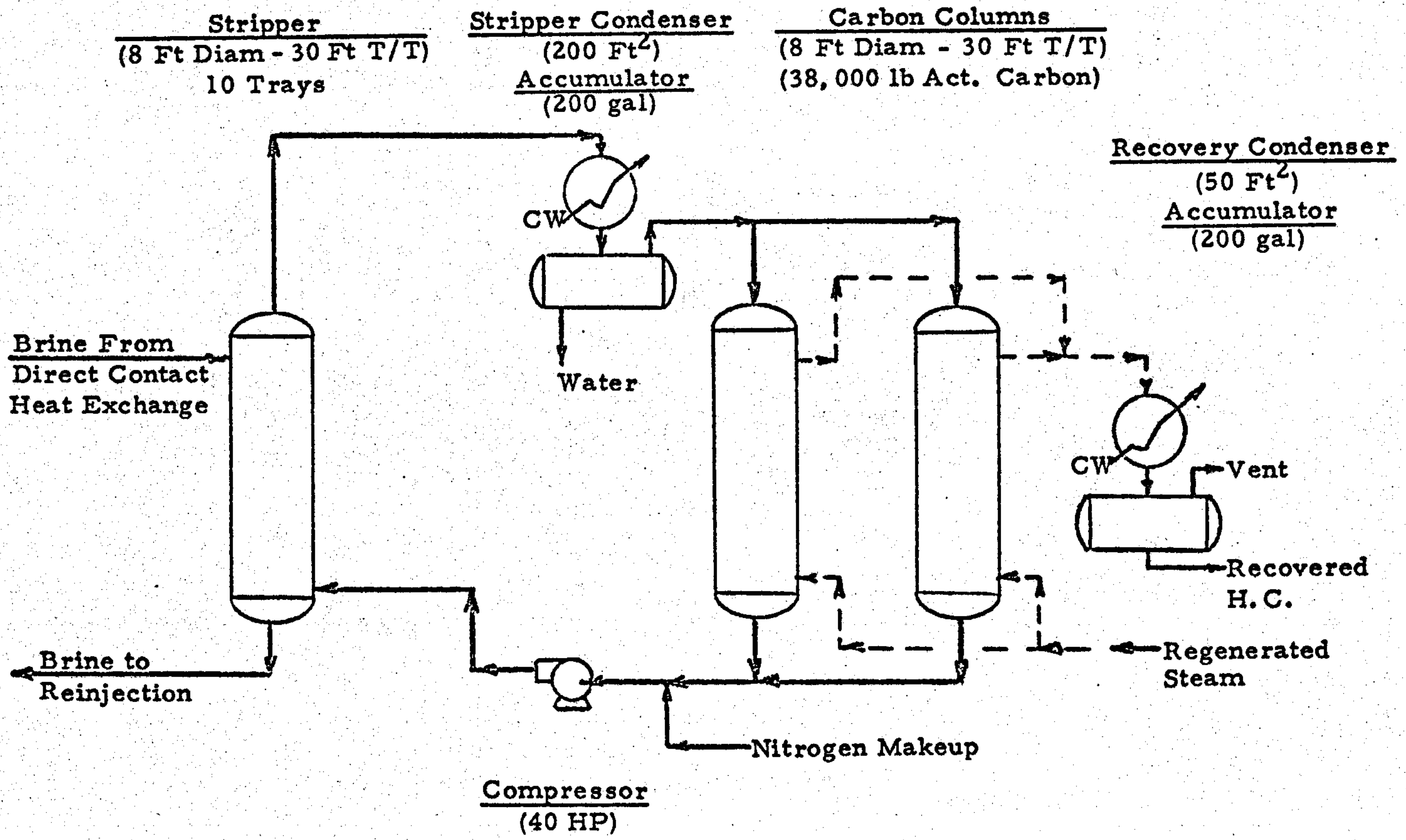


TABLE 8

CYCLE STUDIES - ISOBUTANE

BRINE INLET TEMPERATURE, °F	341
CONTACTOR TEMPERATURE APPROACH, °F	10
CONTACTOR PRESSURE DROP, PSI	3
FLASHING PRESSURE, PSIG	105
BRINE SPECIFIC HEAT	1.0
EXPANDER EFFICIENCY	0.85
CONDENSING TEMPERATURE, °F	100
PUMP EFFICIENCY	0.75
CRITICAL TEMPERATURE, °R	734.7
CRITICAL PRESSURE, PSIA	529.1
MOLECULAR WEIGHT	58.12

TABLE 8 (Continued -2)

CYCLE STUDIES - ISOBUTANE

Operating Pressure, psig	105						205
Expander Inlet Temperature, °F	150	200	225	250	285	300	300
Heating Curve							
Boiling Point, °F/H, Btu/lb HC	133.6/ 21.2	133.6/ 21.2	133.6/ 21.2	133.6/ 21.2	133.6/ 21.2	133.6/ 21.2	182.2/ 54.8
Dew Point, °F/H, Btu/lb HC	133.6/154.4	133.6/154.4	133.6/154.4	133.6/154.4	133.6/154.4	133.6/154.4	182.1/174.8
Temperature, °F/H, Btu/lb HC	139.1/158.3	155.7/171.1	164.1/178.2	172.3/185.9	184.1/198.0	-	221.4/215.0
Temperature, °F/H, Btu/lb HC	144.5/162.3	177.9/191.4	194.5/210.3	211.2/233.9	234.5/279.1	-	260.7/275.2
Expander Inlet, °F/H	150.0/166.5	200.9/217.5	225.0/258.5	250.0/321.7	285.0/503.7	300.0/675.5	300.0/383.7
Cooling Curve							
Temperature, °F/H, Btu/lb HC	130.4/161.0	178.5/209.6	202.2/249.1	225.6/310.4	257.1/488.2	-	239.2/354.3
Temperature, °F/H, Btu/lb HC	120.2/153.7	152.3/180.6	168.1/199.1	183.7/223.3	204.7/275.9	-	192.8/241.1
Temperature, °F/H, Btu/lb HC	110.1/146.9	126.1/158.5	134.0/165.4	141.8/173.9	152.3/191.7	-	146.4/179.8
Temperature, °F/H, But/lb HC	99.9/140.7	99.9/140.7	99.9/140.7	99.9/140.7	99.9/140.6	99.9/140.6	99.9/140.6
Water Vaporized, lb/lb HC	0.0099	0.0330	0.0581	0.1028	0.2482	0.3939	0.1359
Outlet Condensate, lb/lb HC	0.0	0.0	0.0	0.0	0.0		0.0
Brine/HC, lb/lb	0.724	0.975	1.174	1.481	2.360	3.188	2.087
Brine Outlet Temperature, °F	114.4	121.5	124.9	128.5	133.7	136.1	164.6
Power, kwh/10^6 lb Brine							
Gross	2103	2256	2241	2143	1891	1651	3939
HC Pump	144	107	89	71	44	33	153
Cooling Water	958	736	626	520	339		379
Net	1001	1413	1526	1552	1508		3407
Brine Pump	-	-	-	-	-	-	114

H = Enthalpy

TABLE 8 (Continued -3)

CYCLE STUDIES - ISOBUTANE

Operating Pressure, psig	305	405				440		
Expander Inlet Temperature, °F	**300**	**250**	**275**	**300**	**325**	**275**	**300**	**325**
Heating Curve								
Boiling Point, °F/H, Btu/lb HC	215.8/ 81.5	242.2/106.4	242.2/106.4	242.2/106.4	242.2/106.4	250.3/115.3	250.3/115.3	250.3/115.3
Dew Point, °F/H, Btu/lb HC	215.8/188.5	242.2/197.8	242.2/197.8	242.2/197.8	242.2/197.8	250.4/200.0	250.4/200.0	250.4/200.0
Temperature, °F/H, Btu/lb HC	243.8/221.2	244.8/201.4	253.1/212.6	261.5/224.0	269.8/235.7	258.6/212.3	266.9/224.2	275.2/236.4
Temperature, °F/H, Btu/lb HC	271.9/261.9	247.4/204.9	264.1/227.6	280.7/252.1	297.4/280.3	266.8/224.1	283.5/248.9	300.1/276.7
Expander Inlet, °F/H	300.0/318.0	250.0/208.4	275.0/243.4	300.0/285.2	325.0/340.1	275.0/236.0	300.0/276.5	325.0/327.8
Cooling Curve								
Temperature, °F/H, Btu/lb HC	220.0/284.8	160.5/188.3	182.5/216.2	204.5/251.1	225.6/298.7	177.9/209.6	199.3/242.6	200.9/286.3
Temperature, °F/H, Btu/lb HC	180.0/216.2	140.3/169.4	155.0/183.5	169.6/200.9	183.7/222.4	151.9/180.3	166.2/196.5	180.6/217.1
Temperature, °F/H, Btu/lb HC	140.0/171.4	120.1/153.9	127.5/159.6	134.8/166.1	141.8/173.5	125.9/158.4	133.1/164.5	140.2/171.7
Temperature, °F/H, Btu/lb HC	99.9/140.7	99.9/140.7	99.9/140.7	99.9/140.7	99.9/140.7	99.9/140.7	99.9/140.7	99.9/140.7
Water Vaporized, lb/lb HC	0.0821	0.0237	0.0376	0.0588	0.0921	0.0344	0.0535	0.0831
Outlet Condensate, lb/lb HC	0.0	0.002	0.0	0.0	0.0	0.0009	0.0	0.0
Brine/HC, lb/lb	1.920	1.079	1.440	1.862	2.410	1.389	1.839	2.402
Brine Outlet Temperature, °F	182.4	153.8	178.1	194.6	207.4	177.3	197.3	211.8
Power, kwh/10^6 lb Brine								
Gross	4844	5208	5284	5124	4942	5325	5156	4969
HC Pump	279	695	521	403	311	595	449	344
Cooling Water	393	661	502	396	316	520	399	268
Net	4172	3852	4261	4325	4315	4210	4308	4357
Brine Pump	227	341	341	341	341	381	381	381

H = Enthalpy

TABLE 8 (Continued - 4)

CYCLE STUDIES - ISOBUTANE

Operating Pressure, psig	600					700
Expander Inlet Temperature, °F	280	285	290	295	300	300
Heating Curve						
Boiling Point, °F/H, Btu/lb HC	275.0/139.5	275.0/139.5	275.0/139.5	275.0/139.5	275.0/139.5	275.0/131.7
Dew Point, °F/H, Btu/lb HC	275.0/166.8	275.0/166.8	275.0/166.8	275.0/166.8	275.0/166.8	275.0/154.8
Temperature, °F/H, Btu/lb HC	276.6/170.8	278.3/175.3	280.0/180.5	281.6/187.5	283.3/198.6	283.3/169.2
Temperature, °F/H, Btu/lb HC	278.3/175.3	281.7/187.5	285.0/207.6	288.3/217.7	291.7/225.2	291.7/186.3
Expander Inlet, °F/H	280.0/180.6	285.0/207.7	290.0/221.6	295.0/231.9	300.0/241.2	300.0/208.4
Cooling Curve						
Temperature, °F/H, Btu/lb HC	147.0/176.1	156.9/184.9	163.7/192.3	170.8/200.6	178.5/210.7	162.5/190.9
Temperature, °F/H, Btu/lb HC	131.3/162.2	137.9/167.6	142.4/171.6	147.2/175.9	152.3/180.9	141.6/170.9
Temperature, °F/H, Btu/lb HC	115.6/150.8	118.9/153.2	121.2/154.8	123.6/156.6	126.1/158.6	120.8/154.5
Temperature, °F/H, Btu/lb HC	99.9/140.7	99.9/140.7	99.9/140.7	99.9/140.7	99.9/140.7	99.9/140.7
Water Vaporized, lb/lb HC	0.0269	0.0294	0.0320	0.0348	0.0379	0.0320
Outlet Condensate, lb/lb HC	0.0114	0.0094	0.0087	0.0069	0.0035	0.0095
Brine/HC, lb/lb	0.623	1.084	1.316	1.484	1.635	1.224
Brine Outlet Temperature, °F	57.7	156.4	179.4	191.4	200.1	178.5
Power, kwh/10^6 lb Brine						
Gross	1387	5090	5624	5521	4902	4106
HC Pump	1267	911	802	721	655	1132
Cooling Water	1117	653	541	483	448	582
Net	-997	3526	4281	4317	3799	2392
Brine Pump	563	563	563	563	563	677

H = Enthalpy

TABLE 9

CYCLE STUDIES - ISOPENTANE

BRINE INLET TEMPERATURE, °F	341
CONTACTOR TEMPERATURE APPROACH, °F	10
CONTACTOR PRESSURE DROP, PSI	3
FLASHING PRESSURE, PSIG	105
BRINE SPECIFIC HEAT	1.0
EXPANDER EFFICIENCY	0.85
CONDENSING TEMPERATURE, °F	100
PUMP EFFICIENCY	0.75
CRITICAL TEMPERATURE, °R	828.1
CRITICAL PRESSURE, PSIA	490.4
MOLECULAR WEIGHT	72.15

TABLE 9 (Continued -2)

CYCLE STUDIES - ISOPENTANE

Operating Pressure, psig	55	105				
Expander Inlet Temperature, °F	175	215	225	250	275	300
Heating Curve						
Boiling Point, °F/H, Btu/lb HC	172.6/ 43.3	212.2/ 68.6	212.2/ 68.6	212.2/ 68.6	212.2/ 68.6	212.2/ 68.6
Dew Point, °F/H, Btu/lb HC	172.7/199.2	212.2/225.5	212.2/225.5	212.2/225.5	212.2/225.5	212.2/225.5
Temperature, °F/H, Btu/lb HC	173.4/200.1	213.1/226.9	216.5/231.9	224.8/245.4	233.1/260.8	-
Temperature, °F/H, Btu/lb HC	174.2/201.1	214.1/228.3	220.7/238.6	237.4/269.6	254.1/311.5	-
Expander Inlet, °F/H	175.0/202.0	215.0/229.6	225.0/245.7	250.0/300.0	275.0/392.3	300.0/591.6
Cooling Curve						
Temperature, °F/H, Btu/lb HC	131.1/186.3	145.3/204.3	153.8/218.4	174.1/266.9	192.8/351.6	-
Temperature, °F/H, Btu/lb HC	120.7/176.7	130.1/188.7	135.8/197.1	149.4/222.8	161.8/257.2	-
Temperature, °F/H, Btu/lb HC	110.2/165.8	115.0/170.6	117.8/173.8	124.6/182.6	130.8/193.1	-
Temperature, °F/H, Btu/lb HC	99.8/156.4	99.8/156.4	99.8/156.4	99.8/156.4	99.8/156.4	99.8/156.3
Water Vaporized, lb/lb HC	0.0074	0.0374	0.0468	0.0828	0.1525	0.3174
Outlet Condensate, lb/lb HC	0.0	0.0	0.0	0.0	0.0	-
Brine/HC, lb/lb	0.971	1.290	1.413	1.823	2.514	3.991
Brine Outlet Temperature, °F	137.1	168.0	172.5	183.2	193.5	203.8
Power, kwh/10^6 lb Brine						
Gross	4636	5633	5413	5077	4533	3758
HC Pump	101	151	137	106	77	49
Cooling Water	798	610	566	446	342	438
Net	3737	4872	4710	4525	4114	3271
Brine Pump	-	-	-	-	-	-

H = Enthalpy

TABLE 9 (Continued -3)

CYCLE STUDIES - ISOPENTANE

Operating Pressure, psig	205			305		405
Expander Inlet Temperature, °F	**265**	**275**	**300**	**300**	**325**	**325**
Heating Curve						
Boiling Point, °F/H, Btu/lb HC	262.2/102.5	262.2/102.5	262.2/102.5	296.1/127.6	296.1/127.6	322.2/148.8
Dew Point, °F/H, Btu/lb HC	262.1/261.3	262.1/261.2	262.1/261.2	296.1/286.0	296.1/286.0	322.3/304.4
Temperature, °F/H, Btu/lb HC	263.1/263.0	266.4/269.2	274.8/286.0	297.4/288.8	305.7/307.4	323.2/306.6
Temperature, °F/H, Btu/lb HC	264.0/264.7	270.7/277.6	287.4/315.8	298.7/291.6	315.4/331.7	324.1/308.9
Expander Inlet, °F/H	265.0/266.5	275.0/286.5	300.0/253.0	300.0/294.5	325.0/359.8	325.0/311.1
Cooling Curve						
Temperature, °F/H, Btu/lb HC	160.1/228.2	168.2/244.9	187.0/302.2	168.5/246.3	187.9/300.3	168.8/255.0
Temperature, °F/H, Btu/lb HC	140.0/204.1	145.4/214.2	157.9/244.0	145.6/214.6	158.5/245.5	145.8/215.3
Temperature, °F/H, Btu/lb HC	119.9/176.3	122.6/179.8	128.9/189.1	122.7/179.9	129.2/189.4	122.8/180.2
Temperature, °F/H, Btu/lb HC	99.8/156.4	99.8/156.4	99.8/156.4	99.8/156.4	99.8/156.4	99.8/156.4
Water Vaporized, lb/lb HC	0.0531	0.0650	0.1095	0.0662	0.1073	0.0742
Outlet Condensate, lb/lb HC	0.0	0.0	0.0	0.0	0.0	0.0
Brine/HC, lb/lb	2.183	2.437	3.266	4.229	5.798	15.850
Brine Outlet Temperature, °F	224.7	229.5	240.1	275.9	284.0	323.0
Power, kwh/10^6 lb Brine						
Gross	5044	4786	4357	3191	2946	1017
HC Pump	177	159	118	137	265	49
Cooling Water	261	330	254	189	143	51
Net	4606	4297	3985	2865	2538	917
Brine Pump	114	114	114	228	228	341

H = Enthalpy

TABLE 10

CYCLE STUDIES - NORMAL BUTANE

Operating Pressure, psig	138.5	138.5	210	210
Expander Inlet Temperature, °F	176	250	210	300
Brine Inlet Temperature, °F	360	360	360	360
Contactor Temperature Approach, °F	5	5	5	5
Contactor Pressure Drop, psi	3	3	3	3
Flashing Pressure, psig	138.5	138.5	138.5	138.5
Brine Specific Heat	1.0	1.0	1.0	1.0
Expander Efficiency	.85	.85	.85	.85
Condensing Temperature, °F	110	110	110	110
Pump Efficiency	.8	.8	.8	.8
Critical Temperature, °R	765.4	765.4	765.4	765.4
Critical Pressure, psia	550.7	550.7	550.7	550.7
Molecular Weight	58.12	58.12	58.12	58.12
Heating Curve				
Boiling Point, °F/H, Btu/lb HC	173.8/ 38.4	173.8/ 38.4	205.8/ 59.8	205.8/ 59.8
Dew Point, °F/H, Btu/lb HC	173.8/177.6	173.8/177.6	205.8/193.0	205.8/193.0
Temperature, °F/H, Btu/lb HC	174.5/178.3	199.2/203.6	207.2/194.5	237.2/231.7
Temperature, °F/H, Btu/lb HC	175.3/179.0	224.6/238.1	208.6/196.1	268.6/286.8
Expander Inlet, °F/H	176.0/179.6	250.0/287.4	210.0/197/6	300.0/276.2
Cooling Curve				
Temperature, °F/H, Btu/lb HC	138.4/169.9	206.6/269.4	151.8/182.8	230.6/343.9
Temperature, °F/H, Btu/lb HC	128.9/162.0	174.4/213.2	137.8/169.6	190.4/243.9
Temperature, °F/H, Btu/lb HC	119.4/154.9	142.1/175.2	123.9/158.2	150.1/185.5
Temperature, °F/H, Btu/lb HC	109.9/148.2	109.9/148.2	109.9/148.2	109.9/148.2
Water Vaporized, lb/lb HC	0.0145	0.0749	0.0208	0.1316
Outlet Condensate, lb/lb HC	0.0005	0.0	0.0006	0.0
Brine/HC, lb/lb	0.758	1.319	0.887	1.982
Brine Outlet Temperature, °F	128.1	148.5	143.1	179.1
Power, kwh/10^6 lb Brine				
Gross	3675	3909	4796	4679
HC Pump	239	137	363	163
Cooling Water	963	582	834	416
Net	2473	3190	3599	4120
Brine Pump	0	0	76	76

TABLE 11

CYCLE STUDIES - NORMAL PENTANE

Operating Pressure, psig	1385	1385	210	210	105
Expander Inlet Temperature, °F	246	275	278	300	240
Brine Inlet Temperature, °F	360	360	360	360	360
Contactor Temperature Approach, °F	5	5	5	5	10
Contactor Pressure Drop, psi	3	3	3	3	3
Flashing Pressure, psig	138.5	138.5	138.5	138.5	138.5
Brine Specific Heat	1	1	1	1	1
Expander Efficiency	.85	.85	.85	.85	.85
Condensing Temperature, °F	110	110	110	110	100
Pump Efficiency	.8	.8	.8	.8	.75
Critical Temperature, °R	845.1	845.1	845.1	845.1	845.1
Critical Pressure, psia	488.6	488.6	488.6	488.6	488.6
Molecular Weight	72.15	72.15	72.15	72.15	72.15
Heating Curve					
Boiling Point, °F/H, Btu/lb HC	242.1/ 78.6	242.2/ 78.6	274.0/100.4	274.0/100.4	222.7/ 71.1
Dew Point, °F/H, Btu/lb HC	242.1/250.7	242.1/250.7	274.1/275.0	274.1/275.0	222.8/241.9
Temperature, °F/H, Btu/lb HC	243.4/253.1	253.1/272.6	275.4/277.8	282.7/294.3	228.5/225.0
Temperature, °F/H, Btu/lb HC	244.7/255.5	264.0/298.8	276.7/280.7	291.4/316.4	234.3/263.2
Expander Inlet, °F/H	246.0/258.0	275.0/331.1	278.0/283.5	300.0/342.1	240.0/275.5
Cooling Curve					
Temperature, °F/H, Btu/lb HC	160.2/225.8	183.0/290.2	168.3/242.3	185.2/292.3	159.2/242.6
Temperature, °F/H, Btu/lb HC	143.5/205.2	158.7/240.3	148.8/215.9	160.1/244.2	139.3/214.0
Temperature, °F/H, Btu/lb HC	126.7/179.8	134.4/191.5	129.4/183.6	135.1/192.7	119.5/182.2
Temperature, °F/H, Btu/lb HC	110.0/160.8	110.0/160.8	110.0/160.8	110.0/160.8	99.7/160.1
Water Vaporized, lb/lb HC	0.0553	0.1051	0.0672	0.1060	0.0658
Outlet Condensate, lb/lb HC	0.0	0.0	0.0	0.0	0.0
Brine/HC, lb/lb	1.478	2.048	2.046	2.665	1.765
Brine Outlet Temperature, °F	192.5	207.2	228.9	240.3	191.3
Power, kwh/10^6 lb Brine					
Gross	6255	5732	5780	5359	5214
HC Pump	165	119	182	140	114
Cooling Water	550	411	403	313	466
Net	5540	5202	5195	4906	4634
Brine Pump	0	0	76	76	0

TABLE 12

CYCLE STUDIES - ISOBUTANE
(MISCELLANEOUS)

Operating Pressure, psig	175	240	320	560
Expander Inlet Temperature, °F	175	200	225	300
Brine Inlet Temperature, °F	230	300	300	350
Contactor Temperature Approach, °F	10	5	5	10
Contactor Pressure Drop, psi	3	3	3	3
Flashing Pressure, psig	205	52.4	52.4	120
Brine Specific Heat	1	1	1	1
Expander Efficiency	.85	.85	.85	.85
Condensing Temperature, °F	100	95	95	110
Pump Efficiency	.75	.8	.8	.8
Critical Temperature, °R	734.7	734.7	734.7	734.7
Critical Pressure, psia	529.1	529.1	529.1	529.1
Molecular Weight	58.12	58.12	58.12	58.12
Heating Curve				
Boiling Point, °F/H, Btu/lb HC	169.7/ 45.8	195.0/ 67.6	220.1/ 88.3	275.0/144.2
Dew Point, °F/H, Btu/lb HC	169.7/169.6	195.0/183.3	220.1/193.3	275.0/173.2
Temperature, °F/H, Btu/lb HC	171.5/171.0	196.7/184.9	221.7/195.1	283.3/214.5
Temperature, °F/H, Btu/lb HC	173.2/172.5	198.3/186.5	223.4/197.0	291.7/229.5
Expander Inlet, °F/H	175.0/174.0	200.0/188.1	225.0/198.8	300.0/243.7
Cooling Curve				
Temperature, °F/H, Btu/lb HC	134.0/163.9	140.2/173.7	148.5/181.3	190.8/214.6
Temperature, °F/H, Btu/lb HC	122.7/155.4	125.1/161.6	130.6/165.9	163.8/182.0
Temperature, °F/H, Btu/lb HC	111.3/147.8	110.0/151.1	112.8/153.0	136.9/157.8
Temperature, °F/H, Btu/lb HC	99.9/140.7	95.0/141.7	95.0/141.7	109.9/138.5
Water Vaporized, lb/lb HC	0.0114	0.0147	0.0185	0.0409
Outlet Condensate, lb/lb HC	0.0005	0.0006	0.0012	0.0021
Brine/HC, lb/lb	2.548	1.174	1.422	1.372
Brine Outlet Temperature, °F	162.0	142.8	163.5	179.3
Power, kwh/10^6 lb Brine				
Gross	1142	3526	3550	6081
HC Pump	100	329	379	733
Cooling Water	271	600	501	518
Net	771	2597	2670	4830
Brine Pump	0	200	295	469

TABLE 13

CYCLE STUDIES - ISOPENTANE
(MISCELLANEOUS)

Operating Pressure, psig	75	105			138.5			
Expander Inlet Temperature, °F	195	225	225	225	235	250	250	250
Brine Inlet Temperature, °F	300	341	341	341	360	360	360	360
Contactor Temperature Approach, °F	5	10	10	10	5	10	10	10
Contactor Pressure Drop, psi	3	3	3	3	3	3	3	3
Flashing Pressure, psig	52.4	105	105	105	138.5	138.5	138.5	138.5
Brine Specific Heat	1	1	1	1	1	1	1	1
Expander Efficiency	.85	.85	.85	.8	.85	.85	.85	.85
Condensing Temperature, °F	95	110	120	100	110	100	120	130
Pump Efficiency	.8	.8	.75	.75	.8	.75	.75	.75
Critical Temperature, °R	828.1	828.1	828.1	828.1	828.1	828.1	828.1	828.1
Critical Pressure, psia	490.4	490.4	490.4	490.4	490.4	490.4	490.4	490.4
Molecular Weight	72.15	72.15	72.15	72.15	72.15	72.15	72.15	72.15
Heating Curve								
Boiling Point, °F/H, Btu/lb HC	190.5/ 57.4	212.2/ 62.8	212.2/ 57.0	212.2/ 68.6	231.8/ 75.7	231.8/ 81.5	231.8/ 69.9	231.8/ 64.0
Dew Point, °F/H, Btu/lb HC	190.6/213.9	212.2/219.4	212.2/213.2	212.2/225.5	231.8/233.2	231.8/239.3	231.8/226.9	231.8/220.6
Temperature, °F/H, Btu/lb HC	192.1/215.8	216.5/225.7	216.5/219.5	216.5/231.9	232.9/234.8	237.9/249.3	237.9/236.8	237.9/230.4
Temperature, °F/H, Btu/lb HC	193.5/217.8	220.7/232.4	220.7/226.1	220.7/238.6	233.9/236.6	243.9/260.1	243.9/247.5	243.9/241.1
Expander Inlet, °F/H	195.0/219.8	225.0/239.5	225.0/233.2	225.0/245.7	235.0/238.3	250.0/272.1	250.0/259.3	250.0/252.8
Cooling Curve								
Temperature, °F/H, Btu/lb HC	136.3/198.4	160.2/214.9	166.5/211.2	157.1/220.0	158.2/210.5	164.2/237.9	177.2/230.8	183.7/227.0
Temperature, °F/H, Btu/lb HC	122.5/174.7	143.5/194.8	151.0/192.3	138.0/200.6	142.1/192.7	142.7/209.1	158.2/204.4	165.8/201.9
Temperature, °F/H, Btu/lb HC	108.7/169.2	126.8/172.9	135.5/171.8	118.9/175.1	126.1/172.1	121.3/178.1	139.1/176.3	147.9/175.3
Temperature, °F/H, Btu/lb HC	94.8/156.7	110.1/156.1	120.0/155.6	99.8/156.4	110.1/156.1	99.8/156.4	120.0/155.6	130.0/155.2
Water Vaporized, lb/lb HC	0.0327	0.0468	0.0468	0.0468	0.0436	0.0603	0.0603	0.0603
Outlet Condensate, lb/lb HC	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Brine/HC, lb/lb	1.504	1.413	1.413	1.413	1.239	1.496	1.496	1.496
Brine Outlet Temperature, °F	1.568	176.8	181.0	172.5	174.1	185.6	193.7	197.7
Power, kwh/10^6 lb Brine								
Gross	4097	4883	4365	5095	6438	6398	5339	4818
HC Pump	87	133	127	137	191	173	164	159
Cooling Water	521	560	554	561	647	534	532	530
Net	3489	4190	3684	4397	5600	5691	4643	4129
Brine Pump	24	0	0	0	0	0	0	0

TABLE 13 (Continued -2)

CYCLE STUDIES - ISOPENTANE
(MISCELLANEOUS)

Operating Pressure, psig	170	185	210	
Expander Inlet Temperature, °F	250	257	267	300
Brine Inlet Temperature, °F	300	300	360	360
Contactor Temperature Approach, °F	5	5	5	5
Contactor Pressure Drop, psi	3	3	3	3
Flashing Pressure, psig	170	185	138.5	138.5
Brine Specific Heat	1	1	1	1
Expander Efficiency	.85	.85	.85	.85
Condensing Temperature, °F	95	95	110	110
Pump Efficiency	.8	.8	.8	.8
Critical Temperature, °R	828.1	828.1	828.1	828.1
Critical Pressure, psia	490.4	490.4	490.4	490.4
Molecular Weight	72.15	72.15	72.15	72.15
Heating Curve				
Boiling Point, °F/H, Btu/lb HC	247.2/ 94.9	253.9/ 99.5	264.1/ 98.2	264.1/ 98.2
Dew Point, °F/H, Btu/lb HC	247.3/253.5	253.9/258.3	264.1/256.4	264.1/256.4
Temperature, °F/H, Btu/lb HC	248.2/255.1	254.9/260.1	265.1/258.2	276.1/279.8
Temperature, °F/H, Btu/lb HC	249.1/256.7	256.0/262.0	266.0/260.0	288.0/307.7
Expander Inlet, °F/H	250.0/258.2	257.0/263.9	267.0/261.8	300.0/341.9
Cooling Curve				
Temperature, °F/H, Btu/lb HC	152.8/222.6	154.9/226.3	167.1/225.6	193.0/294.5
Temperature, °F/H, Btu/lb HC	133.5/200.7	134.9/203.1	148.1/202.4	165.4/240.3
Temperature, °F/H, Btu/lb HC	114.2/175.2	114.9/176.1	129.1/175.7	137.7/188.2
Temperature, °F/H, Btu/lb HC	94.8/156.7	94.8/156.7	110.1/156.1	110.1/156.1
Water Vaporized, lb/lb HC	0.0480	0.0506	0.0538	0.1060
Outlet Condensate, lb/lb HC	0.0	0.0	0.0	0.0
Brine/HC, lb/lb	3.200	3.722	1.695	2.414
Brine Outlet Temperature, °F	222.4	232.1	208.4	227.1
Power, kwh/10^6 lb Brine				
Gross	3202	2899	6292	5639
HC Pump	94	88	223	152
Cooling Water	248	216	479	342
Net	2860	2595	5590	5145
Brine Pump	0	0	76	76

TABLE 14

CYCLE STUDIES - MIXED FLUID

Working Fluid Composition, %	(1)	(2)	(3)	(4)	(5)
Operating Pressure, psig	210	210	138.5	125	138.5
Expander Inlet Temperature, °F	310	300	300	205	295
Brine Inlet Temperature, °F	360	360	360	300	360
Contactor Temperature Approach, °F	5	5	5	5	5
Contactor Pressure Drop, psi	3	3	3	3	3
Flashing Pressure, psig	138.5	138.5	138.5	52.4	138.5
Brine Specific Heat	1	1	1	1	1
Expander Efficiency	.85	.85	.85	.85	.85
Condensing Temperature, °F	110	110	110	95	110
Pump Efficiency	.8	.8	.8	.8	.8
Critical Temperature, °R				800.1	913.4
Critical Pressure, psia				529.2	436.9
Molecular Weight				65.14	86.18
Heating Curve					
Boiling Point, °F/H, Btu/lb HC	283.3/108.3	258.8/ 88.8	181.7/ 44.8	188.4/ 55.2	291.2/109.9
Dew Point, °F/H, Btu/lb HC	293.0/311.2	273.5/267.2	250.4/259.7	197.9/204.5	291.1/367.4
Temperature, °F/H, Btu/lb HC	298.7/327.3	282.3/286.9	266.9/299.4	200.3/207.3	292.4/373.0
Temperature, °F/H, Btu/lb HC	304.3/345.1	291.2/309.4	283.5/354.7	202.6/210.1	293.7/378.9
Expander Inlet, °F/H	310.0/365.1	300.0/335.6	300.0/438.3	205.0/213.0	295.0/384.9
Cooling Curve					
Temperature, °F/H, Btu/lb HC	182.4/310.8	200.4/293.2	250.2/413.2	149.5/195.0	159.3/321.9
Temperature, °F/H, Btu/lb HC	168.8/280.7	179.3/252.7	231.0/335.6	135.8/180.8	142.8/322.4
Temperature, °F/H, Btu/lb HC	149.2/222.2	158.2/206.5	211.7/273.8	122.0/167.5	126.3/239.0
Temperature, °F/H, Btu/lb HC	129.6/186.2	137.2/175.6	192.4/232.2	108.3/156.3	109.7/194.0
Water Vaporized, lb/lb HC	0.1223	0.1060	0.1940	0.0278	0.1425
Outlet Condensate, lb/lb HC	0.0	0.0	0.0	0.0	0.0
Brine/HC, lb/lb	3.155	2.319	2.138	1.437	3.744
Brine Outlet Temperature, °F	253.0	224.2	163.9	154.7	265.9
Power, kwh/10^6 lb Brine					
Gross	4945	5251	3380	3602	4836
HC Pump	117	154	81	136	67
Cooling Water	305	392	569	543	269
Net	4523	4705	2730	2923	4500
Brine Pump	76	76	0	77	0

(1) Isopentane 60, N-Hexane 40. (2) N-Butane 25, N-Pentane 50, N-Hexane 25. (3) Isobutane 50, N-Hexane 50. (4) N-Butane 50, Isopentane 50.
(5) N-Hexane 100.

50 MWe PLANT DESIGN

DESIGN BASIS

The following design basis is consistent with the choice of the Heber field as the geothermal reservoir. The Heber field is located in the Imperial Valley of Southern California and is representative of a medium temperature low salinity geothermal resource with excellent commercial potential. The plant size of 50 MWe delivered was chosen on the basis of preliminary economic comparisons made as part of a study for the Electric Power Research Institute by Holt/Procon.

Location: Heber field, California

Plant Size: 50 MWe delivered

Brine Inlet Temperature: 182°C (360°F)

Brine Inlet Pressure: 991 kPa (129 psig)

Brine Specific Heat: 1.0

Brine T.D.S.: 15,000 ppm

Condensing Temperature: 43°C (110°F)

Contactor Temperature Approach: 2.8°C (5°F)

Noncondensable Gases: 0.02% w of brine (CO_2 + H_2S)

Expander Efficiency: 85%

Pump Efficiency: 80%

Working Fluid: Isopentane

DESCRIPTION

Figure 16 is the process flow diagram for a 50 MWe direct contact power plant and shows the major equipment sizes, a material balance, and heat exchange duties. The plant site is the Heber field in the Imperial Valley of California.

Hot geothermal brine from the producing wells enters a flash separator where the noncondensable gases and a small amount of steam are separated from the brine by flashing. Brine from the separator is pumped to the top of the direct contact heat exchange column, or contactor, and flows downward through the column transferring heat to the countercurrent flowing isopentane working fluid. Cold brine from the contactor either flows directly to the field for reinjection, or is sent to a hydrocarbon recovery unit and then to reinjection.

Steam and noncondensables from the flash separator flow through a heat exchanger in order to recover this heat content by preheating working fluid. Condensed steam and noncondensables from the heat exchangers are separated in a condensate separator and the streams are sent individually to disposal.

Isopentane flows through the process in an essentially closed loop. Starting with the expander effluent separator, the isopentane is pumped to the bottom of the contactor and flows upward through the contactor countercurrent to the descending brine. The isopentane is heated and vaporized in the contactor. The vaporized isopentane and a small amount of steam flow from the top of the contactor, through a scrubber, and enter the expander. Vapor from the expander is condensed and flows to the expander effluent separator. Excess steam condensate flows to disposal and isopentane is recycled. Small sidestreams of water and isopentane are pumped from the expander effluent separator to supply the needs of the working fluid-flash vapor exchanger and to supply scrubbing liquid to the working fluid scrubber.

DISCUSSION

Operation without a preflash to remove noncondensables was considered and is believed to be technically feasible. Preliminary calculations, however, indicate that the noncondensable gases are preferentially absorbed by the isopentane and will build up to relatively high concentrations in the working fluid loop. Since the problem can be avoided by using a preflash, and since the heat content of the flashed vapor can be easily recovered, this option was chosen.

The flow sheet shows three 30-foot diameter columns being used for contactors; two required for normal operation and one spare. The use of a spare column is probably not justified for the conditions met in the Heber field, but might be desirable for a field with brines having heavy scaling properties. That is, the use of three columns is conservative. The choice of 30-foot diameter columns requiring field fabrication is a matter of economics. Smaller diameter columns, for instance 12-foot diameter, could be shop fabricated, but would require 13 columns for normal operation and more piping and instrumentation. In this case the use of 30-foot diameter columns appears to be economically justified, and is consistent with commercial practice in the petroleum refining industry.

The method for noncondensable gas disposal is not shown on the flow sheet and the choice of method may vary with location. One method is to route the noncondensable gas to a flare stack; a satisfactory method for the Heber field. Condensate from the condensate separator can be combined with spent brine for reinjection. Condensate from the expander effluent separator can also be disposed of with the spent brine.

A discussion of recovery of hydrocarbon from spent brine is given in the section on Working Fluid Loss in Rejected Brine.

ECONOMICS

PURPOSE

The purpose of the economic study is to provide a means of comparing the direct contact process with two other processes for generating electricity from geothermal brines. The two processes chosen for comparison are the closed loop binary and the two-stage flashed steam. The site chosen for purposes of the comparison is the Heber field located in the Imperial Valley of Southern California. This site is representative of a medium temperature low salinity geothermal source of excellent commercial potential. The values given for the binary and flashed steam processes and for field and transmission costs are taken from a recent study done for the Electric Power Research Institute by Holt/Procon, a joint venture of The Ben Holt Co. and Procon Incorporated.

Thus, the study provides a means of comparing the three processes on a consistent basis.

COST ELEMENTS

In estimating the cost of geothermal power delivered to a utility load, we assume that a privately owned producer will sell thermal energy to an investor-owned public utility who will own and operate the power plant and transmission lines. Thus, there are three elements of cost to be considered.

1. Producer's selling price of thermal energy to the utility, or the utility's energy cost.
2. The utility's cost of generating electricity.
3. The utility's transmission cost to a load center.

Since the methods and factors used in estimating the cost elements are different, they are estimated separately. A computer program was developed for this purpose.

ENERGY COST

In estimating the selling price of energy to a utility, we have used the cost-of-service approach. In this approach, we estimate the capital investment and operating costs associated with the development of the field. We then estimate a selling price for the thermal energy which will give the producer a return on investment commensurate with the risks.

The following procedure is built into the program to determine the cost of energy for a particular case:

1. Reservoir Requirements

The amount of electricity generation desired and the energy conversion process utilized will determine the amount of geothermal fluid required at a given site. A knowledge of the reservoir characteristics will indicate the necessary number of production and injection wells, the field layout and the required collection and distribution piping. With this information, the various costs involved with bringing the field into production can then be estimated.

2. Capital Investment

The capital investment for a geothermal project is the money required by the project for which a return on investment is expected. For the purposes of this investigation, it is assumed that none of these funds are obtained by borrowing. The components of the capital investment are as follows:

a. Exploration and Land Acquisition Costs

This represents the money spent in geological and geophysical research, exploratory drilling, bonus payments and other costs involved with establishing the presence of an exploitable geothermal reservoir. Since the reservoir will typically be sufficiently large to supply more than one power plant, only a proportional amount of this charge is assigned to the project under consideration. These costs are incurred prior to the field development phase.

b. Well Drilling Costs

The costs of a drilling program to provide the required production and injection wells are continuously disbursed during the duration of the program. The program culminates in the startup of the power plant.

c. Working Capital

Working capital is required as of the startup of the power plant. The working capital is sufficient to pay one month's expenses, and is returned at the termination of the project.

d. Capital Additions

An additional annual drilling cost is required to provide additional wells in the field to offset the effects of a declining reservoir. Typical decline rates and project lives indicate a capital requirement equal to the initial drilling costs, but disbursed evenly over the life of the project.

The capital investments listed above are expected to return a profit commensurate with the risks involved in a geothermal venture. To account for the time value of money, the discounted-cash-flow method is used to determine the annual revenue requirements.

3. Expenses

Several types of expenses are incurred during the operation of a producing geothermal field. These fall into two classes: cash expenses and book expenses. The book expenses are not deducted from the revenues but are used in determining the taxable income for federal and state income taxes.

a. Cash Expenses

(1) Royalty Payment

Royalty payments are typically 12.5% of the gross revenues.

(2) Operating Expenses

Annual operating expenses include labor, maintenance, supplies, utilities, etc.

b. Book Expenses

(1) Depreciation

Depreciation for tax purposes is calculated by the sum-of-the-years-digits method for a tax life of 15 years.

Depreciation for bookkeeping purposes is calculated by the sum-of-the-years-digits method over the life of the project.

(2) Intangible Drilling Expenses

These expenses are deducted in the year incurred. Since most of the drilling activities occur before any taxable income is produced by the project, it is assumed that the producer can take advantage of this deduction elsewhere to the credit of the project. Although this deduction is currently in use in the oil industry, there is some question as to whether it will also be available to the geothermal industry.

(3) Depletion Allowance

This deductible expense, similar to depreciation, we assume to be 22% of gross revenue. However, it has not been ruled applicable to geothermal in general.

4. Taxes

Federal, state and local taxes will be paid by the project. The federal and state income taxes are paid according to the net taxable income. A federal investment tax credit of 10% is deducted from the federal income tax. The credit is based on the tangible, depreciable assets of the project. The state and local real property taxes are generally charged according to the value of the real property. This is judged to be a function of the revenues produced by that property.

Thus, these taxes are usually a percentage of the gross revenues. They are treated as an ad valorem tax of 10% in this study.

5. Annual Cash Flow

The cash flow is the sales revenues minus taxes and expenses not including depreciation. Since there is also an annual capital addition, this is also subtracted from the cash flow. The cash flow is then discounted at the desired rate of return to a present worth at the beginning of power plant operations. The annual sales revenue is adjusted iteratively until the sum of the discounted cash flows equals the capital investment at the start of power plant operations. The sales revenue then becomes the cost of energy to the power plant.

Table 15 presents an estimate of the initial capital requirements to develop the reservoir for each of the three processes. The estimates include the cost of producing wells, injection wells, dry holes and surface installations.

The cost of the Heber wells of $300,000 each is pretty close to our understanding of actual well costs in the relatively easy drilling characteristics of the Imperial Valley.

In each case, we have assumed that about 20% of the development wells will be dry holes.

Total well costs vary from a low of $5.5 million for the Heber binary process to $7.5 million for the other two options.

Table 16 presents estimates of field operating and maintenance costs for the three cases. The estimate of the field staff portion is shown in Table 17. The annual cost of the field staff including salaries, benefits, field office burden and G&A is estimated to be $253,000.

Producing well maintenance costs are estimated as follows:

1. Each producing well is acidized once per year at a cost of $10,000.

2. Major remedial well work is done once every four years for each well at a cost of $80,000.

3. Two of the original wells will be abandoned by the end of the project at a cost of $50,000 per well.

Injection well maintenance costs were estimated on the following basis:

1. Each injection well is stimulated once per year at a cost of $25,000.

2. Major remedial well work is done once every two years for each well at a cost of $80,000.

3. One injection well will be abandoned by the end of the project at a cost of $50,000.

Annual surface-installation maintenance (labor and materials) is calculated at 4% of the initial capital cost and down-hole surveys are figured at $1,000 per year per well.

The cost of pumping electricity was figured at 2.0 cents/kwh. For the pumped wells this cost represents about one-third of the total operating and maintenance expense.

COST OF GENERATING ELECTRICITY

Capital cost estimates for the three Heber conversion options and for one field installation are presented in Tables 18, 19, 20 and 21. These estimates are made on the basis that a single contract would be let for design, procurement and construction and, therefore, represent the installed cost ready for operation.

Major equipment costs (i.e., pressure vessels, heat exchangers, pumps, cooling tower, turbine and generator) were based on vendor quotations when possible. Construction items (i.e., concrete, piping, structural, instruments, painting, electrical, insulation, paving, roads, fencing and buildings) were based on material takeoffs and current unit prices of such materials. Indirect field costs and home office services are based upon experience in building facilities of similar size and complexity.

The power plant costs for each process at Heber are as follows:

	Million Dollars
Flashed Steam	26.8
Binary	28.5
Direct Contact (Case 1)	29.25

These costs exclude the cost of land and any costs incurred by the owner associated with design and construction.

If the spare contactor column is omitted for the direct contact plant, the capital cost is reduced to 27.6 million dollars. If a stripping unit is added for hydrocarbon recovery from spent brine the capital cost would be increased 0.4 million dollars to a total of 28.0 million dollars (Case 2). Thus it is reasonable to say that the capital cost of a direct contact plant is about equal to that of a binary plant, and could be slightly less. A two-stage flashed steam plant can be expected to have a somewhat lower capital cost.

Table 22 presents estimates of power plant operating and maintenance costs for the three cases. Table 23 is an estimate of the cost of power plant labor, including salaries, benefits, field office burden and G&A expense.

Annual maintenance costs are figured as 2% of the initial plant cost. Cooling water makeup is purchased at $3.50 per acre-foot. Cooling water treatment chemicals are estimated to cost $20,000 per month, based on Imperial Irrigation District's expenses at their El Centro plant.

The cost of generating electricity is estimated by a method used by public utilities. The total cost is the sum of operating and maintenance costs and fixed charges. The fixed charges are each expressed as a percentage of the invested capital. They are:

1. Return on Investment

   Current capital requirements are in the range of 11 to 13%.

2. Income Tax

   The method used takes the expected annual taxes over the life of the power plant including provisions for investment tax credit and interest deductions and converts them to a uniform annual "levelized" expense. An interest rate of 9% is used with a 50-50 debt/equity ratio.

3. Depreciation

   The depreciation expense is often calculated by the straight-line method, but for economic analysis the sinking-fund method is generally used. The program can use either method, but the base case uses the sinking-fund method (at the rate of return).

4. Ad Valorem Tax

   This accounts for the various property and ad valorem taxes. A typical value is 2.5% of capital cost.

5. Administrative and General Expense

   This is typically one percent of capital cost.

6. Insurance

   This is typically one-tenth percent of capital cost.

TRANSMISSION COST

Capital costs for transmission are estimated at $500,000 and operating and maintenance costs at $5,000/year. The fixed charge factors are the same as for generating electricity.

COST OF DELIVERED POWER

The delivered power cost is the sum of the costs of energy, generation, and transmission. The program gives the annual delivered power cost and also uses the load factor (typically 85%) and the plant size to determine the unit power cost (mils/kwh). Also included is a printout of the yearly cash flows for the geothermal field.

Table 24 is the program printout for the direct contact process Case 1, and includes the values used for factors such as project life and rate of return. The capital cost for the power plant in this case is $29,250,000 and the operating cost includes $670,000/year for hydrocarbon losses at $0.06/lb. The total power cost is 40.4 mils/kwh. Case 2 (Table 25) has a lower power plant capital cost, $28,000,000; a reduction of $1,650,000 for elimination of a spare direct contact heat exchange tower, and an addition of $400,000 for a hydrocarbon stripping unit. An operating cost credit of $570,000/year is taken for the recovery of 85% of the hydrocarbon lost in Case 1. The total power cost is reduced to 38.3 mils/kwh.

COST COMPARISON

Table 26 compares power costs for the direct contact, binary and two-stage flashed steam processes. The binary process has the lowest power cost, 35.2 mils/kwh. The flashed steam process at 38.1 mils/kwh is 0.2 mils/kwh less than the direct contact process Case 2 and 2.3 mils/kwh less than Case 1.

It would appear that where the scaling properties of the brine permit the use of shell and tube heat exchangers, the binary process is to be preferred. For scaling brines the choice would have to be made between the direct contact and the flashed steam processes. This study indicates little difference in power costs between the direct contact and flashed steam processes. The brine rate for the direct contact process is slightly lower, however, and the process is in an earlier stage of development. It is possible that a direct contact process can be developed that will be superior to the flashed steam process.

The near term (by 1980) viable alternative energy sources available to a utility in the southwest are coal-based or oil-based. It is our impression that coal-based power is somewhat less expensive than oil-based power and that a fairly typical cost would be about 30 mils/kwh, while oil-based (burning low-sulfur fuel oil) would be somewhat higher, about 35 mils/kwh. The costs for geothermal power at a field like Heber appear to be competitive.

The geothermal power costs presented here can be expected to provide good relative values, but the absolute values are less valid. The effects of changes in key variables are explored in the sensitivity analysis presented in Table 27.

TABLE 15

ESTIMATED INITIAL FIELD CAPITAL COSTS

(ALL COSTS IN $ K)

	HEBER		
ITEM	DIRECT CONTACT	BINARY	FLASH
PRODUCING WELLS			
No. of Wells x Cost/Well	16 x 300	12 x 300	16 x 300
Cost	4,800	3,600	4,800
INJECTION WELLS			
No. of Wells x Cost/Well	8 x 300	6 x 300	8 x 300
Cost	2,400	1,800	2,400
DRY HOLES			
No. of Wells x Cost/Well	3 x 250	2 x 250	3 x 250
Cost	750	500	750
WELL COST	7,950	5,900	7,950
SURFACE INSTALLATION*	7,410	5,900	7,800
TOTAL FIELD COST	15,360	11,800	15,750

---

*Includes down-hole pumps.

TABLE 16

ESTIMATED FIELD OPERATING AND MAINTENANCE COSTS

(ALL FIGURES IN $ K/YEAR)

	HEBER		
ITEM	DIRECT CONTACT	BINARY	FLASH
FIELD LABOR (INCLUDING OVERHEAD & G&A)	253	253	253
PRODUCING WELL MAINTENANCE	491	368	491
INJECTION WELL MAINTENANCE	524	393	524
SURFACE INSTALLATION MAINTENANCE	256	236	256
DOWN HOLE SURVEYS	24	18	24
MISCELLANEOUS SUPPLIES	53	40	53
PURCHASED POWER	678	665	885
TOTALS	2,279	1,973	2,486

TABLE 17

ESTIMATED FIELD STAFF COSTS

POSITION	NO. OF HIRES	RATE $/MONTH	RATE $/MONTH
FIELD OPERATORS	4	1,000	4,000
ROUSTABOUT	1	1,000	1,000
ELECTRICIAN INSTRUMENT SPECIALIST MECHANIC	2	1,200	2,400
FOREMAN	1	1,500	1,500
OFFICE MANAGER	1	1,000	1,000
MECHANICAL ENGINEER	0.5	1,800	900
PRODUCTION ENGINEER	0.5	1,800	900
	10.0		11,700
OVERHEAD*			9,360
TOTAL MONTHLY COST			21,060
ANNUAL COST			$253,000

---

*Overhead includes fringe benefits, field burden and G&A expense.

TABLE 18

THE BEN HOLT CO.

ESTIMATE SUMMARY SHEET

CUSTOMER

LOCATION Heber, California

JOB NO.

PLANT Direct Contact

REV. NO. 0 DATE 5/19/76

ACCOUNT	Materials	Subcontract	Labor	TOTAL
1100 Columns (incl. trays)	635,400	1,710,000		2,345,400
1200 Pressure Vessels				
1300 Heat Exchangers	1,236,000			1,236,000
1400 Furnace/Heaters				
1500 Pumps	863,500			863,500
1600 Boilers				
1700 Cooling Towers		2,070,000		2,070,000
1800 Compressors-Turbine Generator	3,400,000			3,400,000
1900 Tanks				
2800 Other	214,400			214,400
Labor			207,700	207,700
TOTAL MAJOR EQUIPMENT	6,349,300	3,780,000	207,700	10,337,000
3100 Concrete	215,000		220,000	435,000
3200 Pipe, Valves, Fittings	1,700,000	10,000	1,100,000	2,810,000
3300 Structural Steel	425,000		210,000	635,000
3400 Instruments	400,000	80,000	40,000	520,000
3500 Painting		30,000		30,000
3600 Electrical	1,275,000	685,000		1,960,000
3700 Insulation		260,000		260,000
3800 Paving, Roads, Fences & Misc.	20,000	90,000	100,000	210,000
3900 Buildings		200,000		200,000
TOTAL CONSTRUCTION ITEMS	4,035,000	1,355,000	1,670,000	7,060,000
DIRECT FIELD COSTS	10,384,300	5,135,000	1,877,700	17,397,000
Indirect Field Costs (pg. 2)	415,700	256,000	3,029,300	3,701,000
TOTAL FIELD COSTS	10,800,000	5,391,000	4,907,000	21,098,000
8200 Home Office Services				3,319,000
SUB-TOTAL				24,417,000
9500 Sales Tax on Material				623,000
9200 Fee & Contingency				2,110,000
Escalation				2,100,000
TOTAL SELLING PRICE				$29,250,000

TABLE 19

THE BEN HOLT CO.

ESTIMATE SUMMARY SHEET

CUSTOMER

LOCATION Heber, California

JOB NO.

PLANT Binary

REV. NO. 0 DATE 4/22/76

ACCOUNT	Materials	Subcontract	Labor	TOTAL
1100 Columns (incl. trays)				
1200 Pressure Vessels	213,000			213,000
1300 Heat Exchangers	2,700,000			2,700,000
1400 Furnace/Heaters				
1500 Pumps	1,132,000			1,132,000
1600 Boilers				
1700 Cooling Towers		1,800,000		1,800,000
1800 Turbine & Generator	3,700,000			3,700,000
1900 Tanks				
2800 Other	155,000			155,000
Labor			200,000	200,000
TOTAL MAJOR EQUIPMENT	7,900,000	1,800,000	200,000	9,900,000
3100 Concrete	200,000		210,000	410,000
3200 Pipe, Valves, Fittings	1,640,000	60,000	1,000,000	2,700,000
3300 Structural Steel	400,000		200,000	600,000
3400 Instruments	360,000	60,000	40,000	460,000
3500 Painting		50,000		50,000
3600 Electrical	1,225,000	650,000		1,875,000
3700 Insulation		255,000		255,000
3800 Paving, Roads, Fences & Misc.	20,000	80,000	100,000	200,000
3900 Buildings		200,000		200,000
TOTAL CONSTRUCTION ITEMS	3,845,000	1,355,000	1,550,000	6,750,000
DIRECT FIELD COSTS	11,745,000	3,155,000	1,750,000	16,650,000
Indirect Field Costs (pg. 2)	471,000	145,000	2,208,000	2,824,000
TOTAL FIELD COSTS	12,216,000	3,300,000	3,958,000	19,474,000
8200 Home Office Services				3,119,000
SUB-TOTAL				22,593,000
9500 Sales Tax on Material				707,000
9200 Fee & Contingency				2,600,000
Escalation				2,600,000
TOTAL SELLING PRICE				$28,500,000

TABLE 20

THE BEN HOLT CO.

ESTIMATE SUMMARY SHEET

CUSTOMER

JOB NO.

PLANT Flashed Steam

LOCATION Heber, California

REV. NO. 0 DATE 4/22/76

ACCOUNT	Materials	Subcontract	Labor	TOTAL
1100 Columns (incl. trays)				
1200 Pressure Vessels	283,000			283,000
1300 Heat Exchangers	1,280,000			1,280,000
1400 Furnace/Heaters				
1500 Pumps	451,000			451,000
1600 Boilers				
1700 Cooling Towers		1,600,000		1,600,000
1800 Turbine & Generator	6,880,000	200,000		7,080,000
1900 Tanks				
2800 Other - Vacuum Equip. and etc.	406,000			406,000
Labor			400,000	400,000
TOTAL MAJOR EQUIPMENT	9,300,000	1,800,000	400,000	11,500,000
3100 Concrete	200,000		350,000	550,000
3200 Pipe, Valves, Fittings	750,000	10,000	400,000	1,160,000
3300 Structural Steel	200,000		100,000	300,000
3400 Instruments	300,000	100,000	30,000	430,000
3500 Painting		50,000		50,000
3600 Electrical	950,000	425,000		1,375,000
3700 Insulation		200,000		200,000
3800 Paving, Roads, Fences & Misc.	10,000	100,000	90,000	200,000
3900 Buildings		200,000		200,000
TOTAL CONSTRUCTION ITEMS	2,410,000	1,085,000	970,000	4,465,000
DIRECT FIELD COSTS	11,710,000	2,885,000	1,370,000	15,965,000
Indirect Field Costs (pg. 2)	468,000	144,000	1,749,000	2,361,000
TOTAL FIELD COSTS	12,178,000	3,029,000	3,119,000	18,326,000
8200 Home Office Services				2,971,000
SUB-TOTAL				21,297,000
9500 Sales Tax on Material				703,000
9200 Fee & Contingency				2,400,000
Escalation				2,400,000
TOTAL SELLING PRICE				$26,800,000

TABLE 21

THE BEN HOLT CO.

ESTIMATE SUMMARY SHEET

CUSTOMER

JOB NO.

LOCATION Heber, California

PLANT Binary Production & Injection

REV. NO. 0

DATE 4/22/76

ACCOUNT	Materials	Subcontract	Labor	TOTAL
1100 Columns (incl. trays)				
1200 Pressure Vessels	70,000			70,000
1300 Heat Exchangers				
1400 Furnace/Heaters				
1500 Pumps	1,411,000			1,411,000
1600 Boilers				
1700 Cooling Towers				
1800 Compressors				
1900 Tanks				
2800 Other	1,000			1,000
Labor			53,000	53,000
TOTAL MAJOR EQUIPMENT	1,482,000		53,000	1,535,000
3100 Concrete	13,000		18,000	31,000
3200 Pipe, Valves, Fittings	884,000	20,000	677,000	1,581,000
3300 Structural Steel				
3400 Instruments	281,000	149,000	25,000	455,000
3500 Painting				
3600 Electrical	190,000	440,000		630,000
3700 Insulation		10,000		10,000
3800 Paving, Roads, Fences & Misc.		217,000		217,000
3900 Buildings				
TOTAL CONSTRUCTION ITEMS	1,368,000	836,000	720,000	2,924,000
DIRECT FIELD COSTS	2,850,000	836,000	773,000	4,459,000
Indirect Field Costs (pg. 2)	114,000	41,000	727,000	882,000
TOTAL FIELD COSTS	2,964,000	877,000	1,500,000	5,341,000
8200 Home Office Services				888,000
SUB-TOTAL				6,229,000
9500 Sales Tax on Material				171,000
9200 Fee & Contingency				700,000
Escalation				700,000
TOTAL SELLING PRICE				$7,800,000

TABLE 22

ESTIMATED PLANT OPERATING AND MAINTENANCE COSTS

(ALL FIGURES IN $ K/YEAR)

	HEBER		
ITEM	DIRECT CONTACT	BINARY	FLASH
LABOR COST	327	327	327
MAINTENANCE (LABOR AND MATERIALS)	536	570	536
COOLING WATER AND CHEMICALS	1,008	253	258
MISCELLANEOUS	50	50	50
TOTALS	1,921	1,200	1,171

TABLE 23

ESTIMATED POWER PLANT LABOR COSTS

POSITION	NO. OF HIRES	RATE $/MONTH	RATE $/MONTH
OPERATORS	9	1,000	9,000
LABORER	1	750	750
ELECTRICIAN INSTRUMENT SPECIALIST MECHANIC	2	1,200	2,400
OFFICE MANAGER	1	1,000	1,000
SUPERINTENDENT	1	2,000	2,000
	14		15,150
OVERHEAD*			12,120
TOTAL MONTHLY COST			27,270
ANNUAL COST:			$327,000

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*Overhead includes fringe benefits, field burden and G&A expense.

TABLE 24

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                    T H E   B E N   H O L T   C O .


                     GEOTHERMAL PROJECT ECONOMICS
                               06/02/76


HEBER DIRECT CONTACT BINARY CYCLE PLANT - CASE 1


INPUT DATA

     PROJECT LIFE, YEARS                                  25
     EXPLORATION COST                               $800,000.
     LEASE BONUS PAYMENT                                  $0.
     WELL DRILLING COST                           $7,950,000.
     SURFACE INSTALLATION                         $7,410,000.
     WORKING CAPITAL                                $326,718.
     ANNUAL CAPITAL ADDITIONS                       $318,000.
     POWER PLANT INVESTMENT                      $29,250,000.
     TRANSMISSION SYSTEM INVESTMENT                 $500,000.
     FIELD O&M EXPENSE                            $2,279,000.
     PLANT O&M EXPENSE                            $1,921,140.
     TRANSMISSION O&M EXPENSE                         $5,000.
     INTANGIBLE DRILLING EXPENSE                        .517
     DEPLETION ALLOWANCE                                .220
     STATE TAX RATE                                     .090
     FIELD RATE OF RETURN                               .150
     POWER PLANT RATE OF RETURN                         .120


POWER COST SUMMARY

     ENERGY COST                                  $7,296,061.    (48.4%)

     FIXED CHARGES
        RETURN ON INVESTMENT       $3,510,000.
        INCOME TAX                   $992,681.
        DEPRECIATION                 $219,374.
        AD VALORUM TAX               $702,000.
        ADM. & GENERAL               $292,500.
        INSURANCE                     $29,250.

        TOTAL                                     $5,745,805.    (38.1%)

     OPERATING & MAINT. EXPENSE                   $1,921,140.    (12.8%)

     TRANSMISSION COST
        FIXED CHARGES                 $98,219.
        O & M  EXPENSE                 $5,000.

        TOTAL                                       $103,219.    ( 0.7%)
                                                 ------------
     POWER COST                                  $15,066,225.

     POWER RATE = 40.440 MILLS/KWH
```

TABLE 24 (Continued)

FIELD ECONOMICS SUMMARY

ENERGY SALES	$7,296,061.
EXPENSES	$3,191,008.
MARGIN	$4,105,054.

PAYOUT 6.1 YEARS

YEAR	MARGIN	DEPRECIATION	TAXES	CASH FLOW	CASH POSITION
0					-16665
1	4105	865	992	2795	-13870
2	4105	843	1003	2784	-11086
3	4105	820	1015	2772	-8314
4	4105	798	1027	2760	-5554
5	4105	776	1039	2748	-2806
6	4105	753	1197	2590	-216
7	4105	731	1306	2481	2265
8	4105	709	1327	2460	4725
9	4105	686	1349	2438	7163
10	4105	664	1371	2416	9579
11	4105	642	1394	2393	11972
12	4105	619	1417	2370	14342
13	4105	597	1441	2346	16688
14	4105	575	1464	2323	19011
15	4105	552	1488	2299	21310
16	4105	530	1513	2274	23584
17	4105	508	1513	2274	25858
18	4105	485	1513	2274	28132
19	4105	463	1513	2274	30406
20	4105	441	1513	2274	32680
21	4105	418	1513	2274	34954
22	4105	396	1513	2274	37228
23	4105	374	1513	2274	39502
24	4105	351	1513	2274	41776
25	4105	329	1513	2274	44050

(IN THOUSANDS OF DOLLARS)

TABLE 25

```
                    T H E   B E N   H O L T   C O .
                    -------------------------------


                        GEOTHERMAL PROJECT ECONOMICS
                                06/04/76


HEBER DIRECT CONTACT BINARY CYCLE PLANT - Case 2


INPUT DATA

     PROJECT LIFE, YEARS                                          25
     EXPLORATION COST                                       $800,000.
     LEASE BONUS PAYMENT                                          $0.
     WELL DRILLING COST                                   $7,950,000.
     SURFACE INSTALLATION                                 $7,410,000.
     WORKING CAPITAL                                        $326,718.
     ANNUAL CAPITAL ADDITIONS                               $318,000.
     POWER PLANT INVESTMENT                              $28,000,000.
     TRANSMISSION SYSTEM INVESTMENT                         $500,000.
     FIELD O&M EXPENSE                                    $2,270,000.
     PLANT O&M EXPENSE                                    $1,351,250.
     TRANSMISSION O&M EXPENSE                                 $5,000.
     INTANGIBLE DRILLING EXPENSE                                 .517
     DEPLETION ALLOWANCE                                         .220
     STATE TAX RATE                                              .090
     FIELD RATE OF RETURN                                        .150
     POWER PLANT RATE OF RETURN                                  .120


POWER COST SUMMARY

     ENERGY COST                                          $7,296,061.   (51.2%)

     FIXED CHARGES
        RETURN ON INVESTMENT        $3,360,000.
        INCOME TAX                    $950,258.
        DEPRECIATION                  $209,999.
        AD VALORUM TAX                $672,000.
        ADM. & GENERAL                $280,000.
        INSURANCE                      $28,000.

        TOTAL                                             $5,500,258.   (38.6%)

     OPERATING & MAINT. EXPENSE                           $1,351,250.   ( 9.5%)

     TRANSMISSION COST
        FIXED CHARGES                  $98,219.
        O & M  EXPENSE                  $5,000.

        TOTAL                                               $103,219.   ( 0.7%)
                                                         ------------
     POWER COST                                          $14,250,788.

     POWER RATE = 38.252 MILLS/KWH


FIELD ECONOMICS SUMMARY

               ENERGY SALES            $7,296,061.
               EXPENSES                $3,191,008.
                                      ------------
               MARGIN                  $4,105,054.

               PAYOUT       6.1 YEARS
```

TABLE 26

ESTIMATED GEOTHERMAL POWER COSTS — THREE CONVERSION OPTIONS

		Fuel Cost		Power Cost (mils/kwh)				
Case	Brine Rate K lb/hr	¢ per K lb Brine	¢ per M Btu	Fuel	Fixed Charges	Operating & Maintenance	Transmission	Total
Direct Contact (Case I)	9,584	10.2	53.7	19.6	15.3	5.2	0.3	40.4
Direct Contact (Case II)	9,584	10.2	53.7	19.6	14.8	3.6	0.3	38.3
Binary	6,942	12.0	57.5	16.7	15.0	3.2	0.3	35.2
Flashed Steam	10,010	10.3	71.7	20.6	14.1	3.1	0.3	38.1

TABLE 27

SENSITIVITY ANALYSIS - GEOTHERMAL POWER COSTS

BASIS: DIRECT CONTACT BINARY

		Power Cost - mils/kwh				
	Conditions	Fuel	Fixed Charges	Operating & Maintenance	Transmission	Total
1.	Base Case (includes depletion and intangibles write-off)	19.5	15.4	5.2	0.3	40.4
2.	Lower Field Capital and Overhead and Maintenance - 20%	16.5	15.4	5.1	0.3	37.3
3.	Higher Field Capital and Overhead & Maintenance - 20%	23.3	15.4	5.2	0.3	44.2
4.	Field Decline - 0%	18.4	15.4	5.2	0.3	39.3
5.	Field Rate of Return - 10%	15.9	15.4	5.1	0.3	36.7
6.	Field Rate of Return - 20%	24.1	15.4	5.2	0.3	45.0
7.	No Depletion & Intangible	24.8	15.4	5.2	0.3	45.7
8.	Depletion Only	20.7	15.4	5.2	0.3	41.6
9.	Intangibles Only	23.6	15.4	5.2	0.3	44.5
10.	Project Life - 20 Years	20.1	15.6	5.2	0.3	41.2
11.	Project Life - 30 Years	19.2	15.4	5.2	0.3	40.1
12.	Power Plant Rate of Return - 10%	19.6	12.9	5.2	0.2	37.9
13.	Power Plant Rate of Return - 14%	19.6	18.0	5.2	0.3	43.1

PILOT PLANT PROGRAM

CONCLUSIONS FROM PHASE I

Laboratory work and design studies show that the use of the direct contact heat exchange process in the production of electric power is technically feasible. Preliminary cost estimates indicate that the process is competitive with other geothermal power processes, particularly where the closed loop binary process cannot be used due to scaling or corrosion problems. The most important remaining question requiring experimental confirmation is the effect of operation with actual geothermal brine on process performance. The possible effect of scale formation is of particular importance.

Design studies and very preliminary cost estimates lead to the conclusion that best practice will be to operate the heat exchange column at about wellhead pressure. The working fluid of choice appears to be a pentane cut. Operating at relatively low pressure tends to produce somewhat less gross power but more net power, and equipment costs are expected to be substantially lower.

OBJECTIVE

The pilot plant program consists of the design, construction, and operation of a skid-mounted pilot unit. The unit is to be operated in the field using actual geothermal brine. The objective of this effort is to demonstrate the technical and economic feasibility of the direct contact binary process. It is hoped that sufficient information will be obtained to permit the design of a full scale demonstration unit without undue technical or economic risk.

Phase I work has demonstrated that a direct contact binary process is technically feasible. Preliminary cost estimates indicate that electric power can probably be produced at a competitive price, especially for scaling brines. The process has not, however, been operated with geothermal brines. Operation in the field with actual geothermal brines would not only reduce the risk of commercialization, but would also provide a checkout of an integrated production well and plant system.

The nature of geothermal brines, especially with respect to scale forming tendencies, corrosion, and noncondensable gas content, is such that laboratory testing is of limited use. On the other hand, brine characteristics can be expected to have an important effect on process performance. As an example, the rate of scale formation and the physical characteristics of the scale may well dictate the choice of heat exchange column internals. Field testing the process with actual geothermal brines is therefore highly desirable.

Since the properties of geothermal brines vary widely with location, it would be useful to be able to perform tests at several locations. The test device should be a skid-mounted unit and should be designed to provide a reasonable amount of operating flexibility.

Operation of a skid-mounted pilot plant can be expected to provide answers to the following:

1. The rate of scale formation, the properties of the scale, where scale is formed within the equipment, and the effect of scale on equipment performance.
2. The foaming characteristics, if any, of the brine working fluid system.
3. The effect of noncondensables on equipment performance.
4. The amount of hydrocarbon lost in various effluent streams, and the effectiveness of recovery methods.
5. Possible sources of environmental contamination and the effectiveness of control methods.
6. Corrosion characteristics of the brine.
7. Turbine performance with mixed hydrocarbon-steam vapor.

In addition, useful information will also be obtained on items such as ease of process control.

PROGRAM

The program consists of the design, fabrication, and operation of a skid-mounted direct contact binary process pilot unit. The detailed design will include enough flexibility so that the unit can be operated at other sites at a later date, if that proves to be desirable. One or more skids will be used as needed to provide ease of fabrication, modification and maintenance, and ease of transport.

The unit will be moved to the field and connected to existing geothermal production and reinjection wells. Operation of the unit will be continued over a period of time sufficient to provide information on the effect of operating variables on performance and to establish the reliability of the operation at a chosen set of conditions. At the end of operations the unit will be placed in standby condition.

On completion of the field work, technical and cost information will be updated, and a technical and economic evaluation will be prepared.

A more detailed description of the program follows:

Task 1 - Site Selection

Although the pilot unit should be designed so that it can be used at a variety of locations, the basic design should be for a specific site. The specific design can then be reviewed and modified to provide the desired flexibility. All available design data for the selected site will be acquired and organized into a design basis. These data should include information such as brine composition, temperature and pressure, availability of utilities, and interfacing requirements for brine supply and disposal.

The test facility now being constructed at East Mesa would appear to be a logical choice of site.

Task 2 - Design of Skid-Mounted Unit

The preliminary design of Phase I will serve as a starting point. An optimum cycle will be developed for the specific site conditions, along with a heat and material balance. Process flow and piping and instrument diagrams will be prepared and used to develop final skid layout and piping drawings. The design will be reviewed to provide the desired flexibility, both during the design process and at the end.

A field test program will be developed. The program will be set forth in sufficient detail to ensure adequate data collection to meet the objective of the program.

Task 3 - Laboratory Tests

Laboratory tests will be performed, as required, to obtain answers to design problems. This effort will be of limited extent, and may not be required at all.

Task 4 - Fabrication of the Unit

Maximum use will be made of standard purchased components, and outside vessel fabricators will be used where this leads to greater efficiency and less expense. Hydrostatic tests and instrument and control checks will be made in the shop.

Task 5 - Operation of Pilot Unit

The skid-mounted pilot unit will be transported to the field and connected to the brine supply and disposal lines and to the field utilities. These activities and the operation of the unit will be coordinated with the operators of the geothermal field. It may be possible to carry out start-up activities and many of the tests on one shift per day. Continuous runs will require around the clock operation.

The results of the pilot unit operation will be followed closely and results will be evaluated as they are obtained. It is anticipated that some minor unit modifications may be required; and it is important to be able to modify the test program, if necessary, in response to test results as they are developed.

At the conclusion of the test program, the pilot unit will be placed in standby condition.

Task 6 - Technical and Economic Evaluation

The field test results will be analyzed and used to update the technical information. Additional cost information will be obtained as needed to conform to any process or equipment changes resulting from the Phase II work. A technical and economic evaluation of the direct contact process will be prepared. Recommendations for further development of the process will be made if, as expected, the results of the pilot plant work justify further effort.

DESCRIPTION OF SKID-MOUNTED PILOT UNIT

The final design of the skid-mounted pilot unit will depend to some extent on the choice of site, although the unit will be designed with a reasonable degree of flexibility to permit operation at other sites. Figure 18 presents a preliminary flow sheet for the unit. The sizes shown are consistent with the production of 50 kw (gross) of electric power.

Brine is brought into the unit under flow control, using a brine inlet pump if necessary. Hot brine enters the top of the heat exchange column and leaves the bottom under level control. Cooled brine flows to the top of the stripper column, leaves the bottom under level control, and is pumped back to the field for reinjection.

Hydrocarbon from the hydrocarbon accumulator is pumped and metered into the bottom of the heat exchanger column, where it is heated and vaporized by the hot brine. Hot hydrocarbon vapor, along with some vaporized water from the brine, flows from the top of the column to the expander. Flow to the expander is controlled and measured, any excess vapor being bypassed by a pressure controller. Vapor from the expander, and any bypassed vapor, flows to the main condenser. Condensed hydrocarbon and water are separated in the hydrocarbon accumulator. Water is removed through a level control valve; the flow rate is measured and the water is discarded. Hydrocarbon is sent back to the heat exchange column, closing the hydrocarbon loop.

Cold brine entering the stripper is contacted with noncondensable gases, or nitrogen, metered into the bottom of the stripper. Stripped hydrocarbon vapor, stripper gas, and some water vapor flow from the top of the stripper to the stripper condenser. Condensed hydrocarbon and water are separated in the stripper accumulator, with water removed intermittently for disposal. The hydrocarbon is returned to the hydrocarbon loop at the inlet to the hydrocarbon pump, under level control. The stripper gas is returned to the stripper by means of the stripper blower. Excess noncondensable gas, if present, is vented through a pressure controller.

Any noncondensable gases entering the system with the hot brine will accumulate in the hydrocarbon accumulator. Accumulator pressure is maintained by removing noncondensables with the noncondensables compressor. The compressor discharges into the stripper gas loop, ahead of the stripper condenser.

Power produced by the generator is measured and dissipated using an air cooled resistor load.

Where flashed steam is available along with the hot brine, the steam is used to vaporize hydrocarbon in the vaporizer. The hydrocarbon vapor is combined with the heat exchange column vapor and flows to the expander. Condensate from the vaporizer can be combined with the cold brine, or disposed of separately. Noncondensable gases, if present in the flashed steam, are vented from the vaporizer under pressure control. A hydrogen sulfide absorber, or other control method, may be required depending on the amount and composition of these noncondensables.

Part of the flexibility to be incorporated in the design will be the ability to test different column internals, such as sieve trays or packing, in the heat exchange and stripper columns. Provision for changing column height will be provided by flanged column sections.

An expander-generator set has been included in the design although the concept could be tested without it. The expander-generator set should be included because of the lack of commercial know-how available on the use of mixed hydrocarbon-water vapor feed in expanders.

ESTIMATED COST AND SCHEDULE

The estimated cost of the pilot plant program is $273,900 and is expected to require one year. A summary of the cost elements is as follows:

Purchased Parts	$ 50,000
Direct Labor	73,000
Overhead	73,000
Craft Labor	24,000
Transportation	23,000
Other Direct Costs	6,000
Total	249,000
Fee (10%)	24,900
Total Plus Fee	$273,900

The proposed schedule is shown in Figure 19.

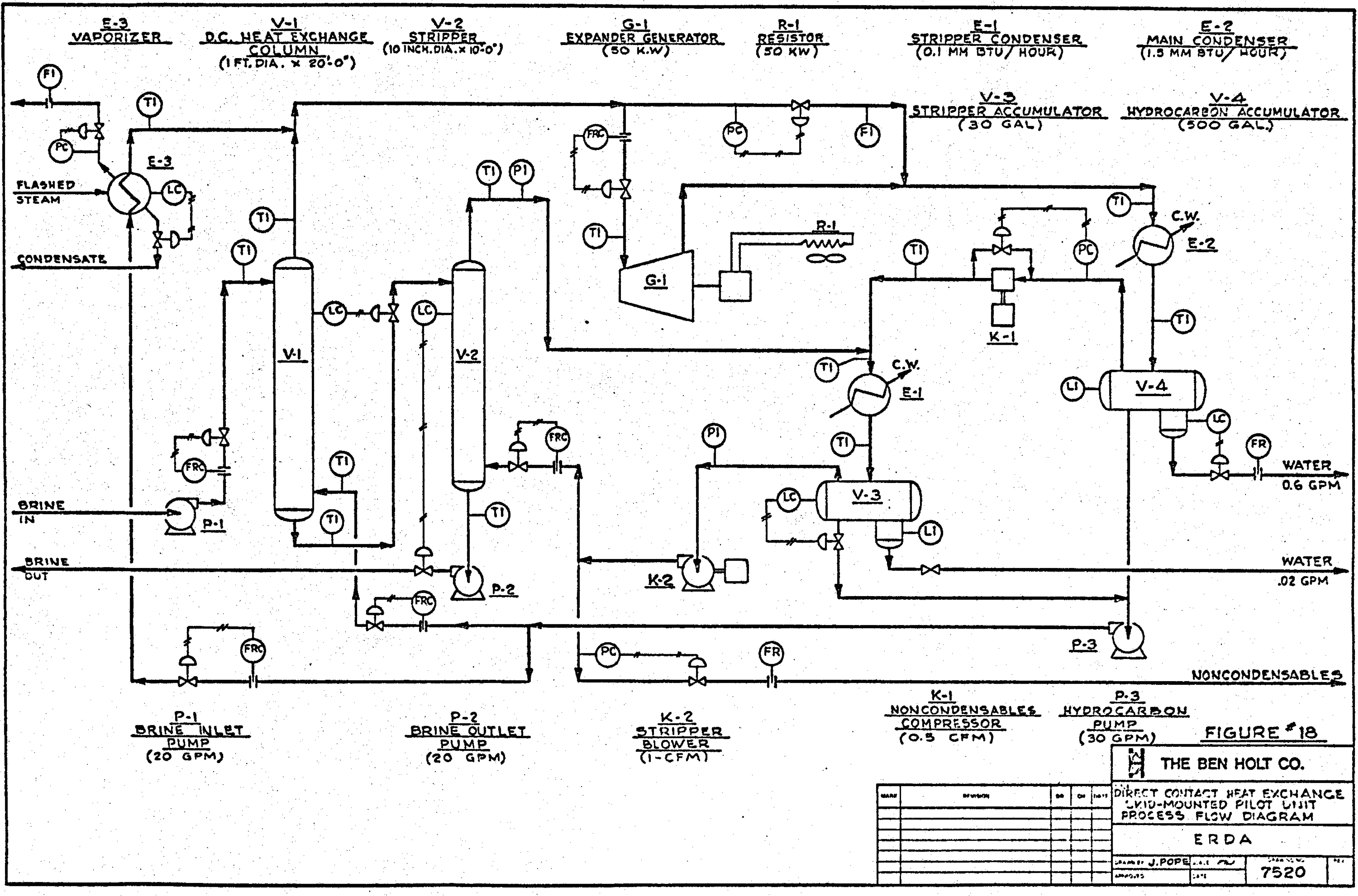


FIGURE #18

THE BEN HOLT CO.

DIRECT CONTACT HEAT EXCHANGE
SKID-MOUNTED PILOT UNIT
PROCESS FLOW DIAGRAM

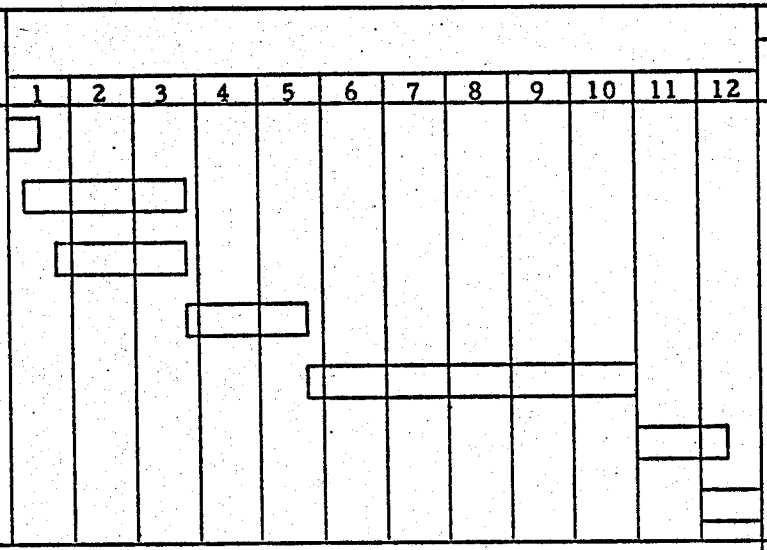
ERDA

7520

FIGURE 19

DIRECT CONTACT BINARY PROCESS

PILOT PLANT SCHEDULE



TASK	MAN-HOURS Engineers	Draftsmen/ Designers	Technicians
1. Site Selection	86	-	-
2. Design	950	280	-
3. Lab Tests	173	-	173
4. Fabrication	346	-	-
5. Operation	1,730	-	1,730
6. Evaluation	390	-	-
7. Final Report	173	20	-
	3,848	300	1,903

Man-Hours, Subtotal	6,051
Administrative Review	60
Project Manager	520
Purchasing	173
Secretarial	500
	7,304
Craftsmen	1,038
TOTAL MAN-HOURS	8,342

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