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Experimental Testing of a Direct Contact Heat Exchanger for Geothermal Brine

Prepared by:

M. W. Urbanek
DSS Engineers, Inc.
Ft. Lauderdale, Florida

Work performed under P.O. Subcontracts
11X13564 and 62X45736

for
Oak Ridge National Laboratory

OPERATED BY
UNION CARBIDE CORPORATION
FOR THE UNITED STATES
DEPARTMENT OF ENERGY

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National Technical Information Service
U.S. Department of Commerce
5285 Port Royal Road, Springfield, Virginia 22161
NTIS price codes—Printed Copy: A09 Microfiche A01

ORNL/Sub-79/13564/1
and
ORNL/Sub-79/457636/1

DSS-079

EXPERIMENTAL TESTING OF A DIRECT CONTACT HEAT EXCHANGER FOR GEOTHERMAL BRINE

Final Report

July 1, 1978-February 1, 1979

Prepared by:

M. W. Urbanek
DSS Engineers, Inc.
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Work performed under P.O. Subcontracts
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Date Published: December 1979

for

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Contract No. W7405 eng 26

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EXPERIMENTAL TESTING OF A DIRECT CONTACT HEAT EXCHANGER FOR GEOTHERMAL BRINES

Abstract

A series of direct contact heat exchanger (DCHX) experiments were conducted at the East Mesa Geothermal Test Site during the period July 1, 1978 to February 1, 1979. The purpose of these tests was to provide additional data necessary to better understand the thermal and hydraulic characteristics of the DCHX binary cycle loop components that may be used to extract energy from geothermal brines. Isobutane and Isopentane were tested as secondary working fluids.

The analytical and experimental efforts were directed at the problems of working fluid loss in the effluent brine, carryover of water vapor with the vaporized secondary fluid and the free CO_2 content of the feed brine. Tests were conducted by Oak Ridge National Laboratory personnel aimed at evaluating the heat transfer performance of various type tubes installed in vertical shell-and-tube secondary fluid condensers. Data was collected while operating a low temperature isopentane cycle with brine preflashed to 210-212 F; the objective being to gain insight to waste heat recovery applications such as the Arkansas Power and Light project.

Possible alternatives for isobutane recovery from the spent brine were investigated. A system was designed and the economic aspects studied.

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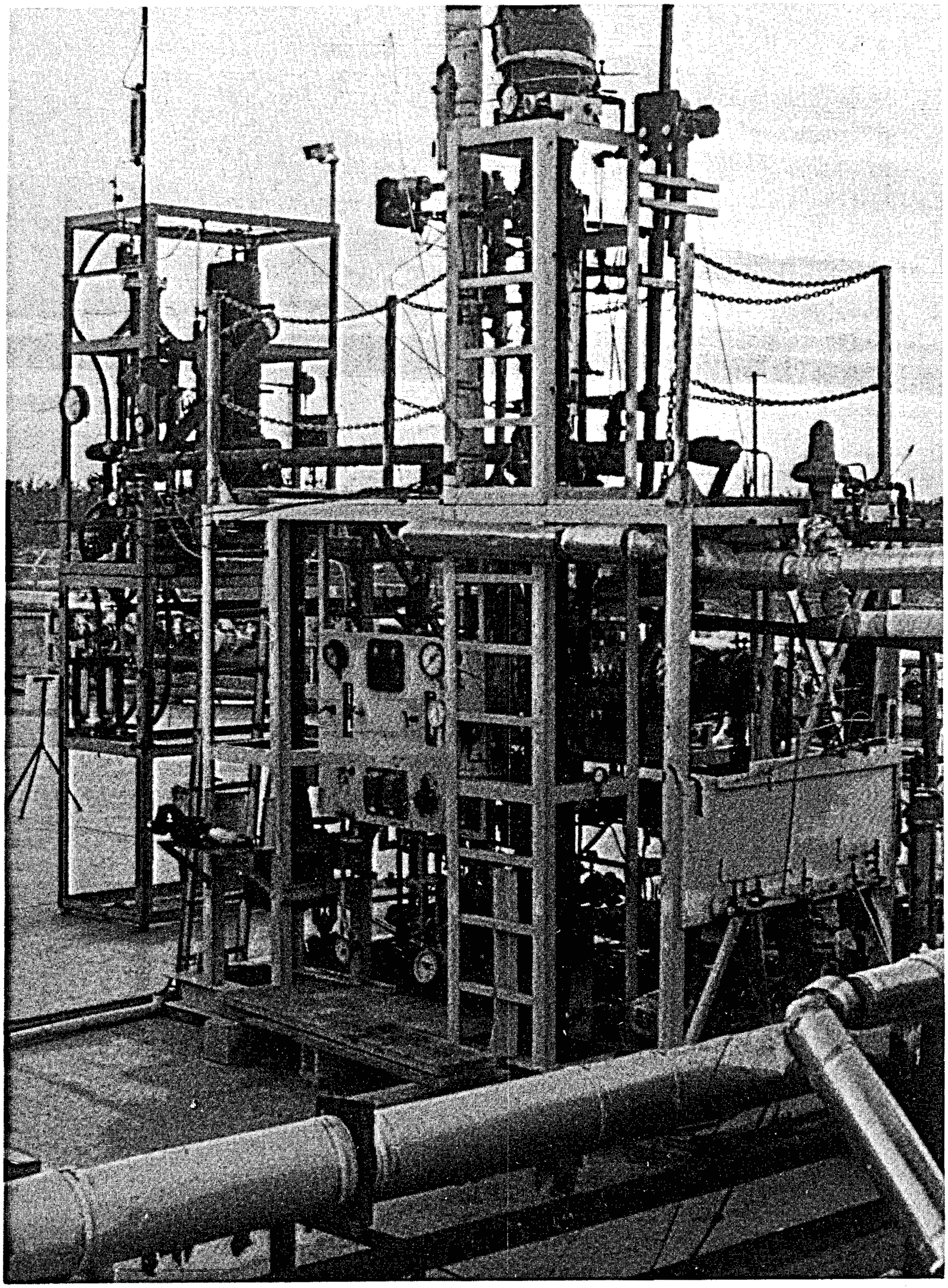

DIRECT CONTACT HEAT EXCHANGER - ORNL CONDENSER LOOP
at
EAST MESA GEOTHERMAL TEST SITE

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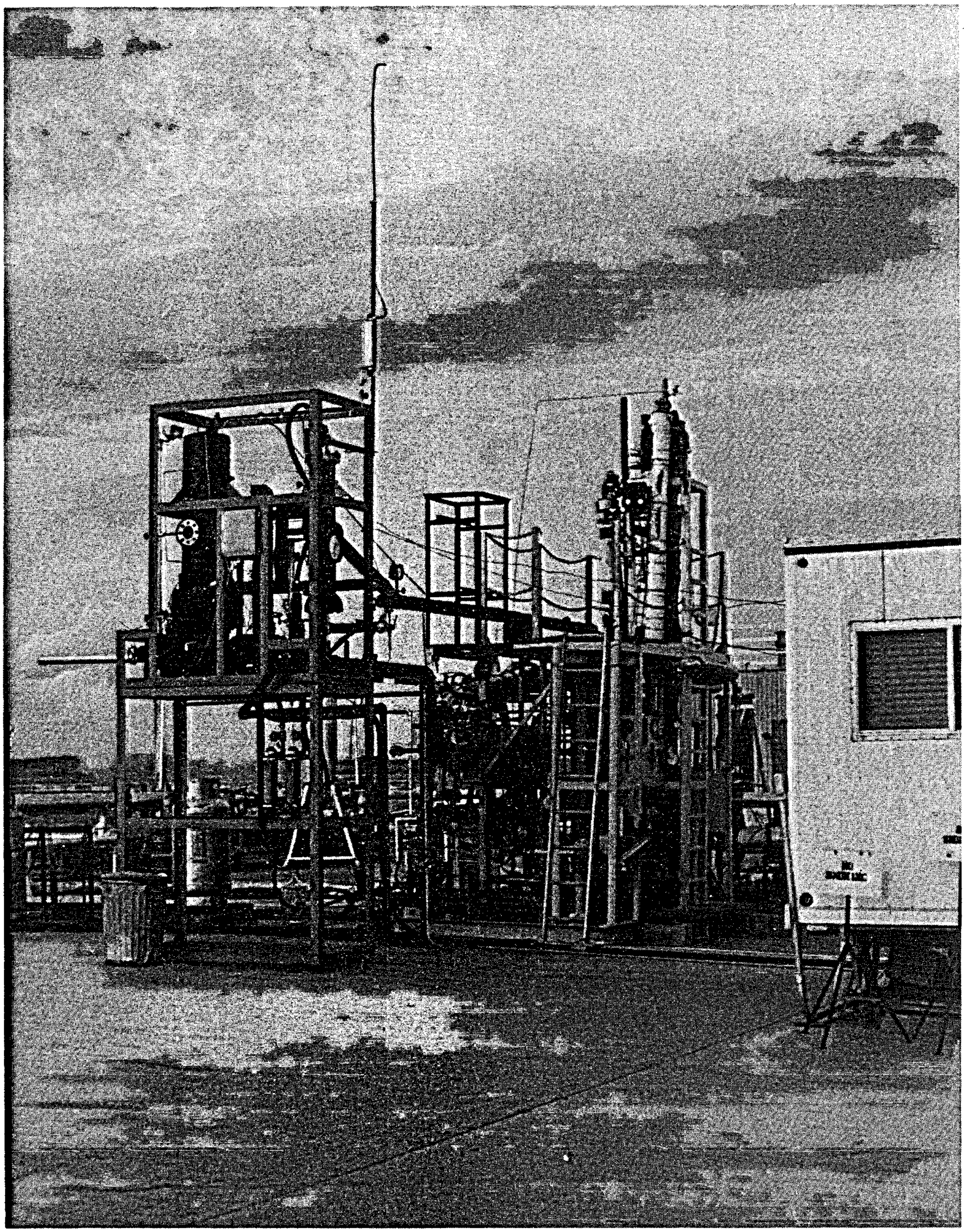


DIRECT CONTACT HEAT EXCHANGER - ORNL CONDENSER LOOP
at
EAST MESA GEOTHERMAL TEST SITE

SECTION 1

SUMMARY & RECOMMENDATIONS

SECTION 1

SUMMARY AND RECOMMENDATIONS

1.1 Summary

A direct contact heat exchanger loop was operated at the East Mesa Geothermal Test Site using Isobutane and Isopentane as working fluids. Tests were conducted from July 1, 1978 to February 1, 1979 under two contractural purchase orders (ORNL 11x13564V and ORNL 62x45736V).

The experimental unit consists of a skid-mounted package containing a complete direct contact binary cycle loop, a hairpin surface type heat exchanger used for calibration, and an auxilliary module piped and instrumented for continuous or intermittent testing of vertical condensers. The test unit provides for closed-loop circulation of a hydrocarbon working fluid. Heat from the geothermal brine is transferred to the hydrocarbon fluid by intimate contact in the DCHX thereby vaporizing the hydrocarbon which can then be expanded through a turbine to produce power. The hydrocarbon is then condensed and returned to the DCHX to complete the loop. An expansion valve was used in lieu of the turbine.

Data necessary for evaluating the heat transfer performance of a 40-tube vertical shell-and-tube condenser and a 2-tube vertical condenser was collected and reduced by ORNL personnel. Field performance data results in lower condensing coefficient values than those calibrated from laboratory conducted experiments. It has not been established whether the problem stems from data collected on the working fluid side or water side of the condenser tubes.

Both high temperature (feed brine at 320-325 F) and low temperature (feed brine at 210-212 F) isopentane cycles were run. Data was collected and reduced for comparison with previously obtained isobutane data in an effort to better understand the problems of working fluid loss to the brine and carryover of water vapor with the vaporized working fluid. No correlation between measured values and operating variables are evident. The data does not suggest common empirical relationships for the two working fluids.

A series of experiments were devised to determine the solubility of free CO_2 in the brine and its effect on overall loop performance. The measured values of free CO_2 dissolved in the brine show a direct linear relationship with pH. Insufficient data was collected to construct CO_2 mass balances or indicate any definite relationship between the soluble CO_2 and thermal performance.

The maximum volumetric heat transfer coefficient obtained for the assumed preheating section of the DCHX using isopentane was approximately 2000 BTU/HR-F-Cu ft. Previously a coefficient of 7300 BTU/HR-F-Cu-ft. was calculated for isobutane. However, it is not definite that the 2000 value represents the maximum that _can_ _be_ obtained before flooding the column.

With adjustment of the Model constants, the Letan-Kehat theoretical model used in liquid-liquid spray column sizing, appears to predict temperature profiles valid for the DCHX. However, it is still not known whether the effective column height predicted by the model can be used to accurately size a DCHX.

A system for recovering isobutane dissolved in the reject brine from a 50 MW direct contact heat exchange power plant was developed and sized. The system uses mechanical compression and refrigeration to recover 90% of the isobutane vapors flashed from the effluent brine steam degassing column. An estimated overall annual savings of $798,000 or .242¢/kwh can be realized with an annual ROI of over 200%.

1.2 Conclusion and Recommendations

The work completed during the seven month experimental program outlined in this report indicates that although there are no apparent major technical obstacles to the development of commercial direct contact heat exchange systems for geothermal power production, there are still some design and optimization studies that should be made. In addition, the concepts of direct contact heat exchange should be expanded and systems developed to include waste heat recovery from low temperature sources such as condensate and bottoming cycles.

Additional experiments are required to further develop the fluted tube and vertical condenser concept of optimizing heat transfer. The effects of the non-condensable gas CO_2 (introduced to the system as soluble CO_2 in the feed brine) on condenser performance should be further explored. Data on fouling resistances associated with the East Mesa brines should be obtained for reference use for the 500 kw pilot unit.

The concept of using direct contact of hydrocarbon vapors with a cooling media to condense the hydrocarbon should be fully explored. A direct contact condenser should be sized, installed and tested on actual geothermal brines.

Analytical work is required to determine a direct, fast and easy method of measuring the "free" or soluble CO_2 content of geothermal brine. Additional experiments are required to determine the overall effect of the free CO_2 on loop components thermal performance.

A system for recovering isobutane from the effluent brine stream using the concepts of compression and refrigeration should be sized, installed and tested at East Mesa.

The relationship between flow rate, dispersed phase drop size and column diameter of the DCHX is still not definitely known. The mathematical models used to predict these in spray column applications, namely, the Minard-Johnson model and the Sakiadis-Johnson model, cannot readily be verified using the existing equipment. A new smaller unit specifically designed and instrumented for the purpose should be built and tested. In addition, this new test unit should be designed to specifically test the Letan-Kehat model for estimating column height, or, provide data necessary to devise a new model.

An important item to reduce costs in the binary cycle that has been neglected to a large extent, is optimization on the heat rejection system. The use of low-cost plastic is possible, either directly for use as condenser material or as surface in dry cooling powers. Concepts and information developed in the OTEC program can be transferred and optimized for this application. A test unit should be designed, fabricated, installed and tested at East Mesa.

Now that the direct contact concept for heating and evaporating a secondary fluid with low salinity geothermal brine has been proven at East Mesa, the design should be tested with high salinity brine. This would require the construction of a new unit located at Niland or a similar high salinity reservoir.

SECTION 2

INTRODUCTION

SECTION 2

INTRODUCTION

There are at present three basic systems in use or under development for recovery of energy from geothermal fluids; (1) the total-flow system, (2) the flashed-steam system, and (3) the binary-cycle system.

The total flow system, which is being developed by Lawrence Livermore Laboratories, shows the greatest efficiency for electrical generation but it also may have the greatest problems with scaling and corrosion.

The second alternative for energy recovery is using multi-stage flash-down with steam to a conventional turbine-generator. One experimental facility with this thermal loop is being operated by San Diego Gas and Electric Co. under contract with DOE in Niland, California. This system also shows much potential but again is plagued by scaling and spent brine disposal problems.

DSS Engineers, Inc., has been under contract with DOE for the development and pilot plant testing of the third alternative or binary-cycle system since 1975. This system uses a secondary working fluid vapor generated in a direct contact heat exchanger to drive a turbo-generator in a closed Rankine cycle for power generation. Most of the experimental testing has been directed at using isobutane as the working fluid. However, recent studies have shown that other working fluids, particularly isopentane, may be more suitable for use with geothermal fluid at some temperatures. The theoretical advantages in using isopentane instead of isobutane include lower working pressures, lower working fluid losses and higher effectiveness with geothermal fluid in certain temperature ranges.

Heat transfer considerations become of increasing economical and technological significance in the utilization of low temperature geothermal energy for power production due to typically low conversion efficiencies. Oak Ridge National Laboratory has initiated a heat transfer development program to improve the condensation heat transfer effectiveness for a variety of hydrocarbon and fluorocarbon working fluids to be used in these low temperature geothermal power cycles.

As part of their geothermal applications program, ORNL conducted a series of condenser performance tests using isobutane vapor generated by the DCHX, during the early months of 1978. The tests were conducted simultaneously with other experimental performance and optimization testing of the DCHX binary loop under contract by Lawrence Berkeley Laboratory (Contract No. 4366002). The ORNL tests were aimed at evaluating the difference in heat transfer performance between smooth tubes and a variety of vertical fluted (Gregorig) tubes installed in a vertical shell-and-tube condenser. The results of this preliminary test program are presented in a Final Engineering Report[1].

In July 1978, DSS Engineers, Inc., was contracted by Oak Ridge National Laboratory (ORNL Order No. 11x-13564V) to operate the Direct Contact Heat Exchanger Loop in support of additional vertical shell-and-tube condenser tests using isobutane as the secondary working fluid. This three month test program also included a series of DCHX performance tests using isopentane.

In October 1978, DSS Engineers, Inc., was awarded a second contract with ORNL (ORNL Order No. 62x-45736V) to conduct a series of follow-up tests and studies over the succeeding four month period. The main objectives of these additional tests were: 1) to modify the DCHX loop for operation with isopentane and low temperature (205-210 F) pre-flashed brine; 2) perform a series of comparison tests using a low temperature-isopentane cycle; 3) determine the water vapor fraction and isopentane dissolved in the effluent brine and attempt to formulate empirical relationships valid for both isobutane and isopentane; 4) quantify and qualify the effects of varying the amount of free CO_2 contained in the feed brine on overall condenser and DCHX performance; and 5) design a system for recovering isobutane from the effluent brine.

This report summarizes the procedures and presents the results of the tests and studies performed under the two ORNL contracts.

SECTION 3

DESCRIPTION OF DIRECT CONTACT HEAT EXCHANGER LOOP COMPONENTS

SECTION 3

DESCRIPTION OF DIRECT CONTACT HEAT EXCHANGER LOOP COMPONENTS

3.1 General Process Description

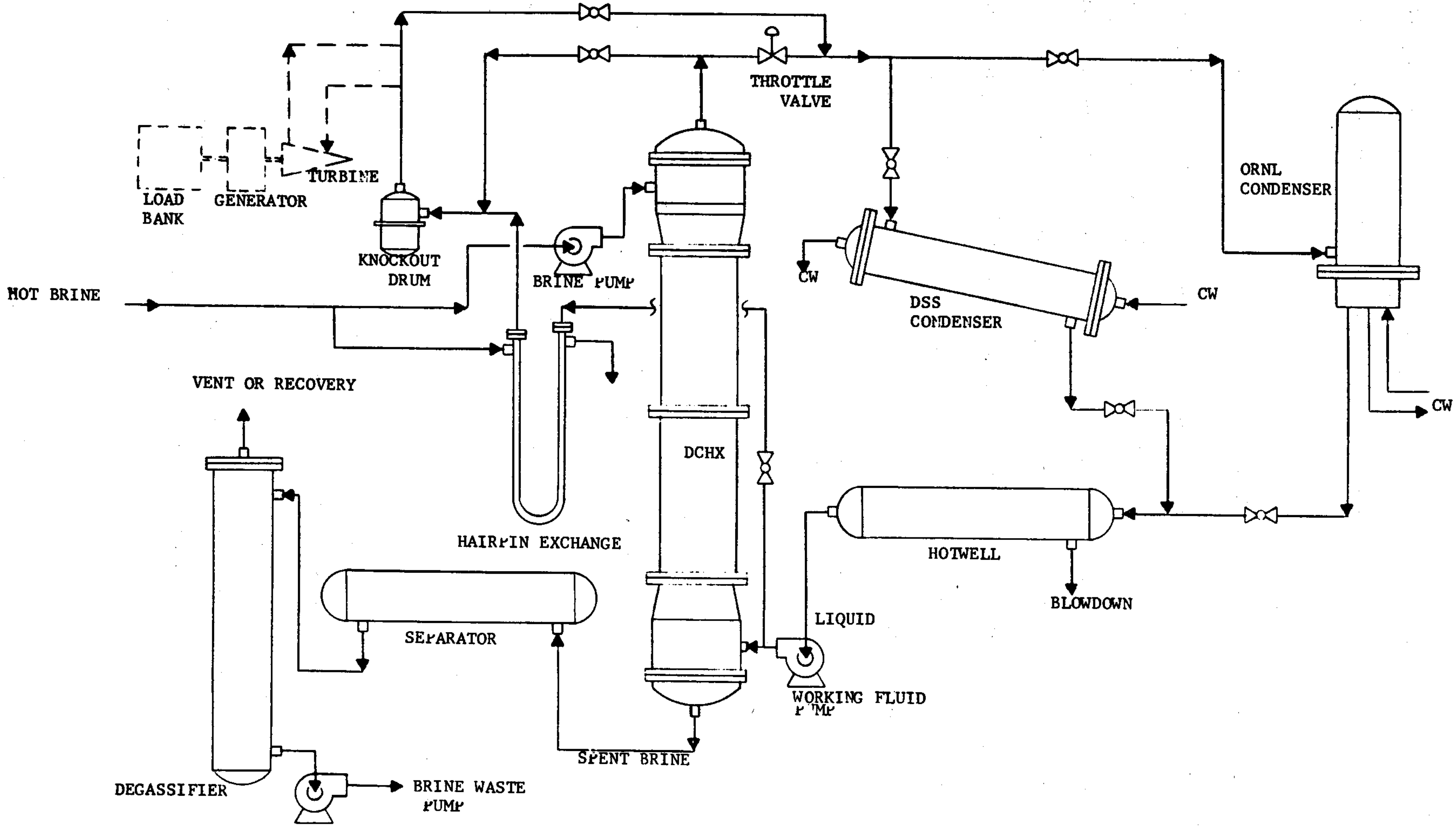
The test unit provided for the continuous circulation, contact and separation of brine and a secondary working fluid. Heat from geothermal brine is transferred to the secondary fluid by intimate contact in a single direct contact heat exchanger combining preheating and vaporization. Other major components are the separator, condensers, hotwell, knockout drum, degassifier (stripper), hairpin exchanger used for calibrations, and the process pumps. The general relationship of these components is illustrated in Figure 3-1. Photographs of the module are shown on pages i and ii.

Liquified working fluid is pumped from the hotwell to the direct contact heat exchanger where it is heated and vaporized by counter current contact with brine. The vaporized working fluid and a small amount of water vapor are then expanded to the condensing pressure through a turbine or throttled by a valve in lieu of the turbine. The vapor is then condensed, subcooled, and the liquid returned to the hotwell to repeat the cycle. Water condensate in the knockout drum and hotwell are continuously drained off.

Hot brine is pumped from the goethermal supply to the direct contact heat exchanger where heat is extracted from it to heat and vaporize the working fluid. The cooled brine leaving the heat exchanger enters the separator where any small, entrained drops of working fluid are coalesced and separate out. This high pressure brine is then reduced

FIGURE 3-1

DCHX LOOP COMPONENTS



in pressure to atmospheric pressure by spraying it into a degasifier where most of the working fluid dissolved in the brine is flashed out. Finally, the brine is sent to reinjection.

3.2 General Component Design Details

All test vessels, with exception of the Oak Ridge Condenser Module, were fabricated of carbon steel to Section VIII, Division I of the ASME Boiler and Pressure Vessel Code. Plate, standard weight pipe (sch.40) and welding fittings were used for the shells. Nozzles were either schedule 80 pipe with 300# flanges or 3000# screwed couplings. Overpressure protection was provided by safety valves installed behind stainless steel rupture discs. The DSS Condenser was a standard off-the-shelf item with brass shell and tubes, and cast iron water boxes. Piping was predominately carbon steel hydraulic tubing to facilitate bending and minimize the number of joints. "Swagelok" type fittings were used extensively to allow frequent disassembly and modification of the piping system. The main process piping about the DCHX was changed to schedule 40 for the low temperature cycle runs.

3.2.1 Direct Contact Heat Exchanger

The preheater and boiler are combined into one single column having and overall height of about thirteen feet. The distance between the isobutane distributor at the bottom and the brine distributor at the top is ten feet four inches. The straight section of column is nine feet in length with an inside diameter of six inches. At both ends a conical section with an included angle of 20 degrees flares to a ten inch diameter. The transition at the bottom results in a gradually decreasing downward velocity of the brine and the enlarged diameter at the bottom allows sufficient area in the annulus between the isobutane distributor and the wall to reduce the brine velocity in the straight section. At the top of the column the ten inch diameter provides a large area for mist eliminators. The column is shown schematically in Figure 3-4. Details of the upper boiler assembly are shown in Figure 3-5. Figure 3-6 is a cutaway drawing of the direct contact heat exchanger showing the theoretical preheating and vaporizing sections in a high temperature isobutane cycle application.

Working fluid is introduced near the bottom of the column by a plate containing 390 drilled holes 0.060 inch in diameter. Type 316 stainless steel, 1/8 inch thick was used for the plate. On the bottom face of the plate the edges of the holes were beveled while the upper face of the plate was ground to ensure sharp edges. After machining, the surface of the plate was slightly oxidized to make it more olephobic (less wettable by working fluid) by pickling in concentrated nitric acid.

Brine may be introduced either through a perforated distribution ring or a tangential open pipe inlet, both at the top of the column. The vapors leaving the top of the column must pass through a knit stainless steel wire mesh demister pad six inches thick which minimizes carryover of either liquid working fluid or brine.

FIGURE NO. 3-4

DIRECT CONTACT HEAT EXCHANGER

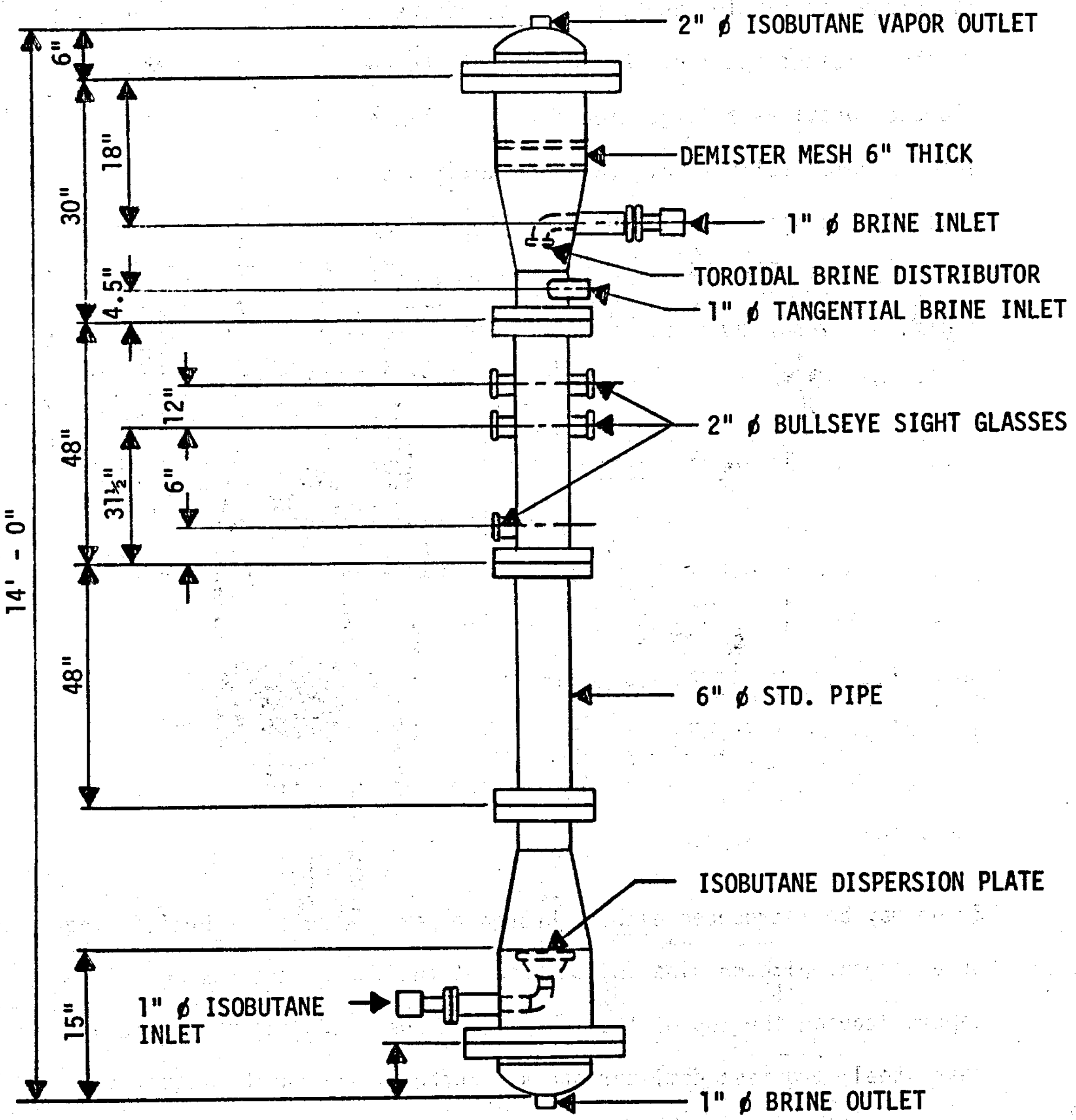


FIGURE 3-5

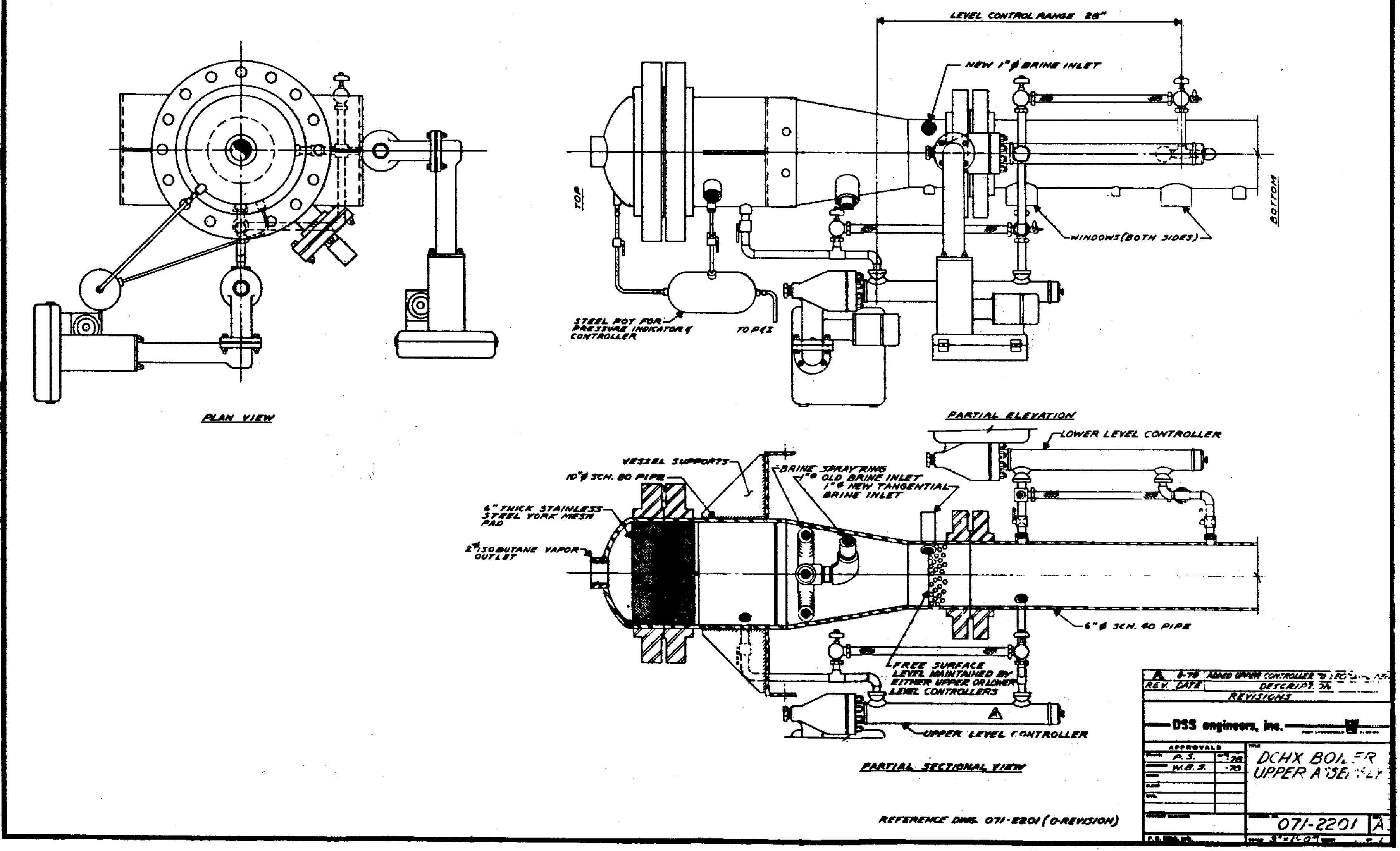
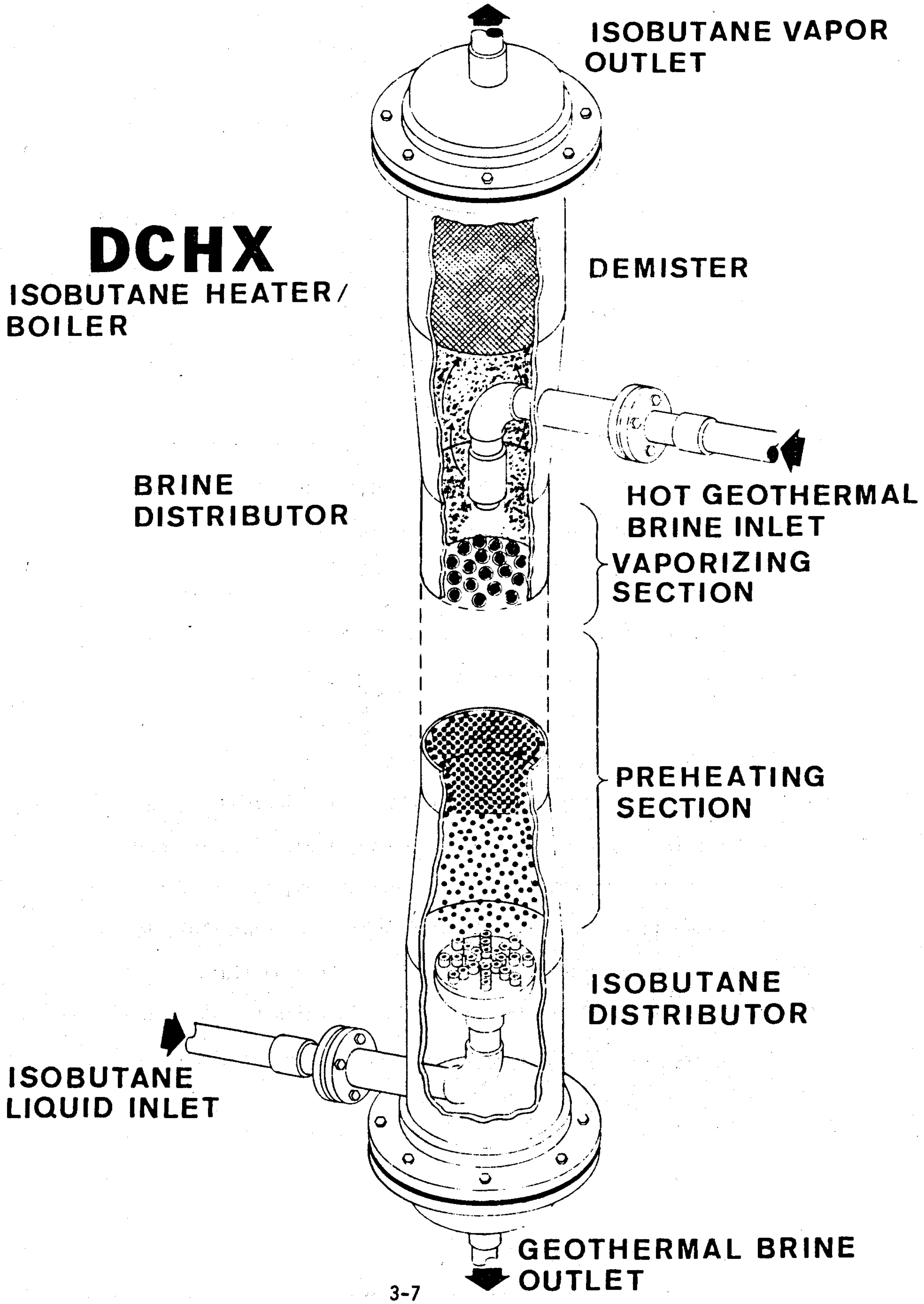

LEVEL CONTROL RANGE 28"
NEW 1"Ø BRINE INLET
TOP
BOTTOM
WINDOWS (BOTH SIDES)
STEEL POT FOR PRESSURE INDICATOR & CONTROLLER
TO P&I
PLAN VIEW
PARTIAL ELEVATION
LOWER LEVEL CONTROLLER
VESSEL SUPPORTS
10"Ø SCH. 80 PIPE
6" THICK STAINLESS STEEL YORK MESH PAD
2" ISOBUTANE VAPOR OUTLET
BRINE SPRAY RING
1"Ø OLD BRINE INLET
1"Ø NEW TANGENTIAL BRINE INLET
6"Ø SCH. 40 PIPE
FREE SURFACE LEVEL MAINTAINED BY EITHER UPPER OR LOWER LEVEL CONTROLLERS
UPPER LEVEL CONTROLLER
PARTIAL SECTIONAL VIEW
REFERENCE DWG. 071-2201 (0-REVISION)
A 8-78 ADDED UPPER CONTROLLER TO
REV. DATE DESCRIPTION
REVISIONS
DSS engineers, inc.
APPROVALS
P.S. -78
W.B.S. -78
DCHX BOILER
UPPER ASSEMBLY
071-2201
A

FIGURE NO.3-6


DCHX
ISOBUTANE HEATER/ BOILER
ISOBUTANE VAPOR OUTLET
DEMISTER
BRINE DISTRIBUTOR
HOT GEOTHERMAL BRINE INLET
VAPORIZING SECTION
PREHEATING SECTION
ISOBUTANE DISTRIBUTOR
ISOBUTANE LIQUID INLET
GEOTHERMAL BRINE OUTLET

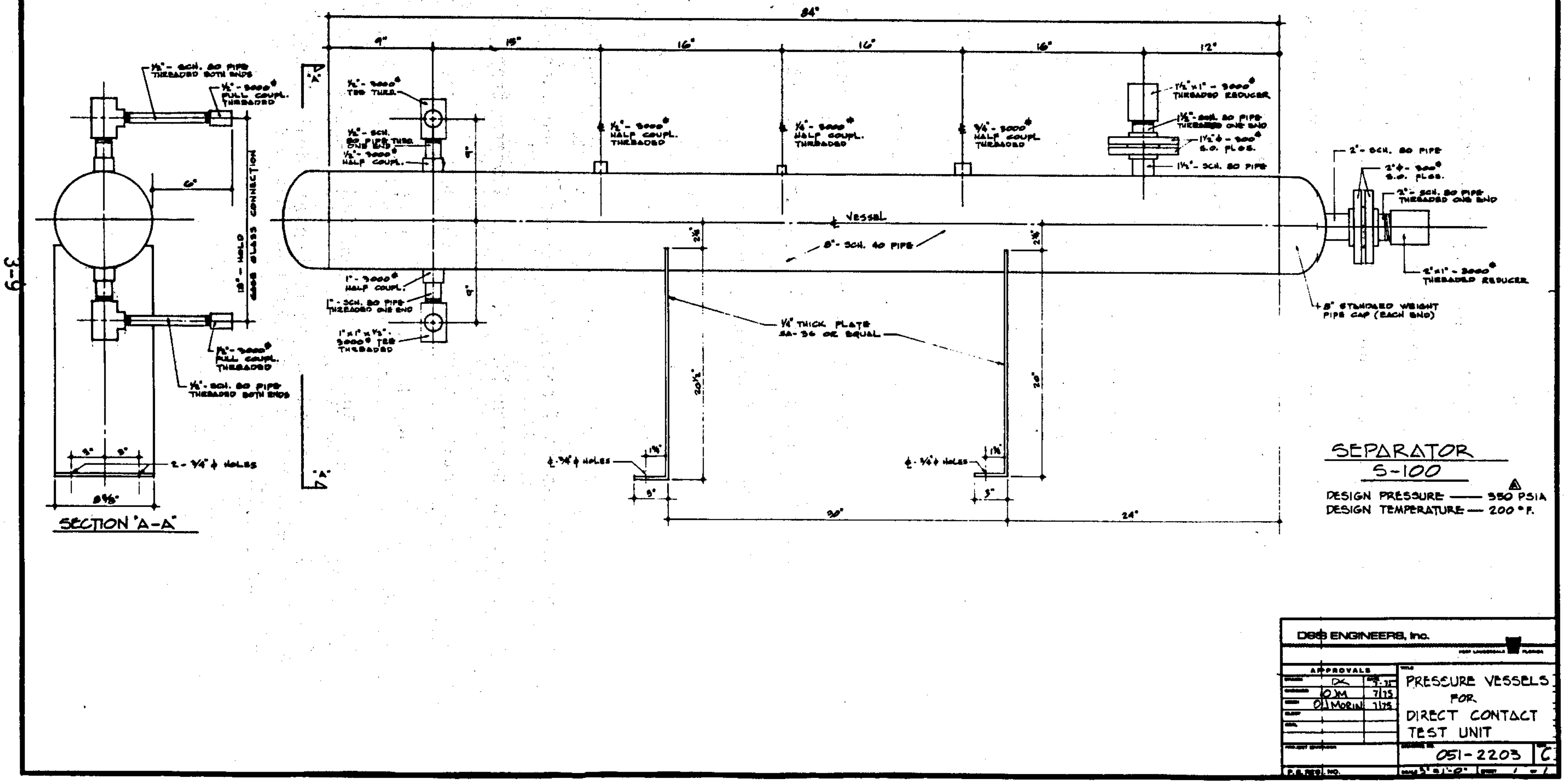
The construction of the column incorporates five pairs of flanges to facilitate inspection and make experimental modifications. Numerous connections are provided on the shell of the column for vents, drains and instrumentation including thermowells, one pressure gage connection, two manometer taps, four high-pressure windows (or "bulls-eyes") for viewing the dispersion, and a pair of gage glass connections.

3.2.2 Separator

Brine leaving the direct contact heat exchanger enters the separator which is a horizontal settler with a coalescer at the inlet end. The vessel is seven feet long and eight inches in diameter, providing a retention time of 3.7 minutes for the brine. It is installed with its longitudinal axis tilted up one degree from the horizontal toward the outlet end. The separator is shown as Figure 3-7.

The coalescer, which fills the entire cross section of the separator at the inlet end, consists of a knit mesh of stainless steel wire and glass fiber. Its function is to entrap and hold the small drops of working fluid entrained in the brine stream so that they can coalesce with other small drops and form drops large enough to rise to the surface. Besides recovering small quantities of dispersed liquid working fluid (as opposed to dissolved working fluid) from the brine discharge stream during normal operating conditions, the separator also provides volume to cope with the large amounts of working fluid entrained in the brine during upset conditions. Working fluid liquid collected is periodically vented to the hotwell which is always at a lower pressure than the separator when the test unit is in operation.

FIGURE 3-7


½" - SCH. 80 PIPE THREADED BOTH ENDS
½" - 3000# FULL COUPL. THREADED
6"
18" - HOLD GAGE GLASS CONNECTION
½" - 3000# FULL COUPL. THREADED
½" - SCH. 80 PIPE THREADED BOTH ENDS
3"
3"
2 - ¾" ϕ HOLES
8⅝"
SECTION "A-A"
"A"
"A"
84"
9"
15"
16"
16"
16"
12"
½" - 3000# TEE THRD.
½" - SCH. 80 PIPE THRD. ONE END
½" - 3000# HALF COUPL.
9"
9"
1" - 3000# HALF COUPL.
1" - SCH. 80 PIPE THREADED ONE END
1" x 1" x ½" - 3000# TEE THREADED
℄ ½" - 3000# HALF COUPL. THREADED
℄ ½" - 3000# HALF COUPL. THREADED
℄ ¾" - 3000# HALF COUPL. THREADED
1½" x 1" - 3000# THREADED REDUCER
1½" - SCH. 80 PIPE THREADED ONE END
1½" ϕ - 300# S.O. FLGS.
1½" - SCH. 80 PIPE
2" - SCH. 80 PIPE
2" ϕ - 300# S.O. FLGS.
2" - SCH. 80 PIPE THREADED ONE END
2" x 1" - 3000# THREADED REDUCER
8" STANDARD WEIGHT PIPE CAP (EACH END)
℄ Vessel
8" - SCH. 40 PIPE
2⅛"
2⅛"
¼" THICK PLATE SA-36 OR EQUAL
20½"
20"
1⅛"
1⅛"
℄ - ¾" ϕ HOLES
℄ - ¾" ϕ HOLES
3"
3"
30"
24"
SEPARATOR
S-100
DESIGN PRESSURE —— 350 PSIA
DESIGN TEMPERATURE —— 200 °F.
DSS ENGINEERS, Inc.
APPROVALS
DK
7-75
OJM
7/75
OJMORIN
7/75
PRESSURE VESSELS FOR DIRECT CONTACT TEST UNIT
051-2203
C
P.E. REG. NO.
3" = 1'-0"
1 of 1

3.2.3 DSS Condenser

Working fluid vapor, after throttling through a valve or expanding through the turbine, is condensed on the shell side of a conventional shell-and-tube heat exchanger by cooling water flowing through a single pass of tubes. The unit is installed with its longitudinal axis tilted up twenty degrees from the horizontal. The vapor inlet is at the upper end of this unit and the condensate outlet is at the lower end. The cooling water flows counter current with respect to the working fluid vapor. The condersate outlet and a vent line are connected to the hotwell in such a way that a liquid level is maintained in the bottom of the condenser to sub-cool working fluid liquid. The total heat transfer area for this condenser is 120 sq.ft. The condenser is shown in Figure 3-8.

3.2.4. Hotwell

The hotwell serves as a receiver for condensed working fluid from which the pump takes suction. As test unit operating conditions change, the amounts of working fluid in the direct contact heat exchanger varies. The hotwell provides the inventory to cope with these operating changes as well as losses through leakage. The vessel is six feet long and 10-5/8 inches in diameter. It is installed with its horizontal axis tiled up one inch from the horizontal toward the suction line of pump. A water trap was installed such that the water condensate in the hotwell may be drained off the lower end of the vessel. Instrumentation on the hotwell includes temperature and pressure indication and one gage glass. A conductivity cell is also installed in the condensate drain line. The hotwell is shown in detail as Figure 3-9.

FIGURE NO. 3-8
SINGLE PASS CONDENSER

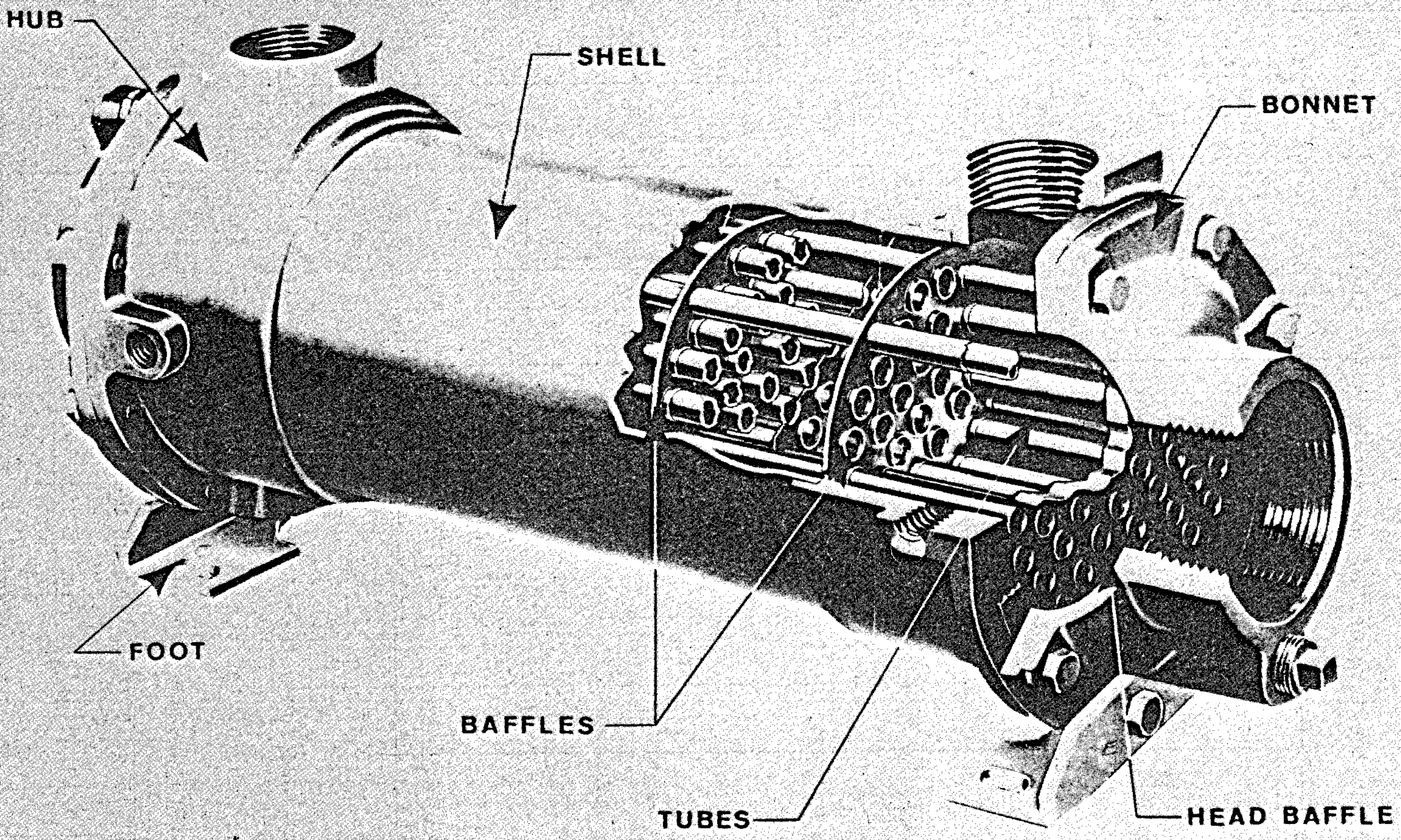
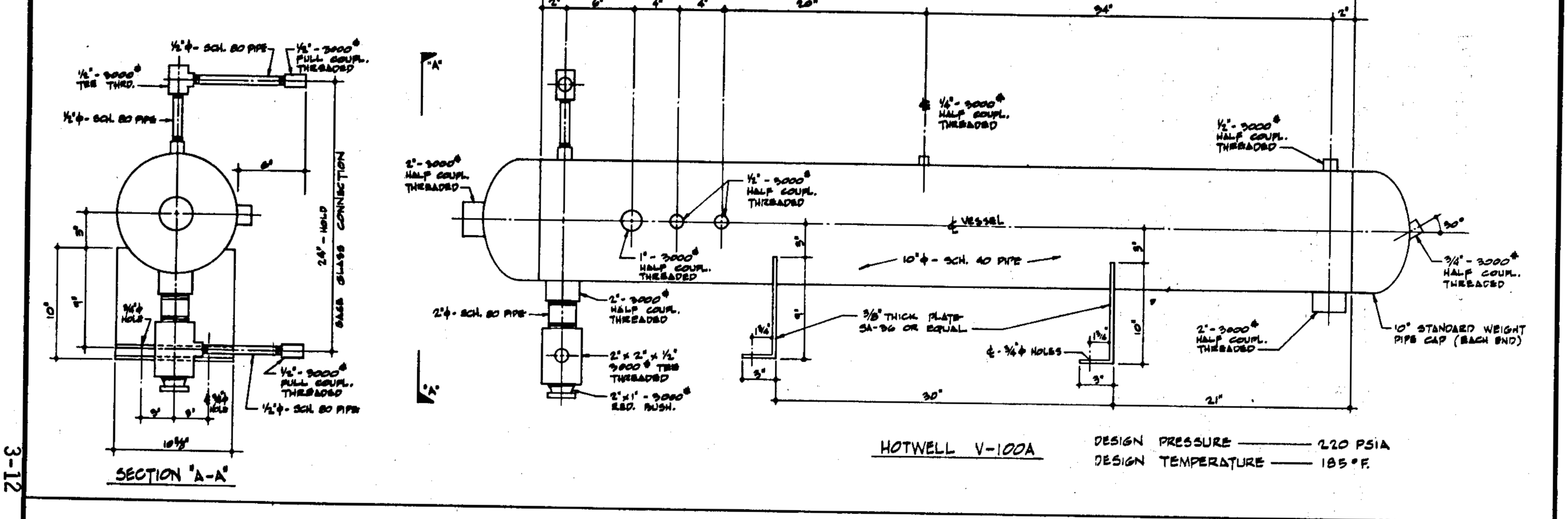


FIGURE 3-9



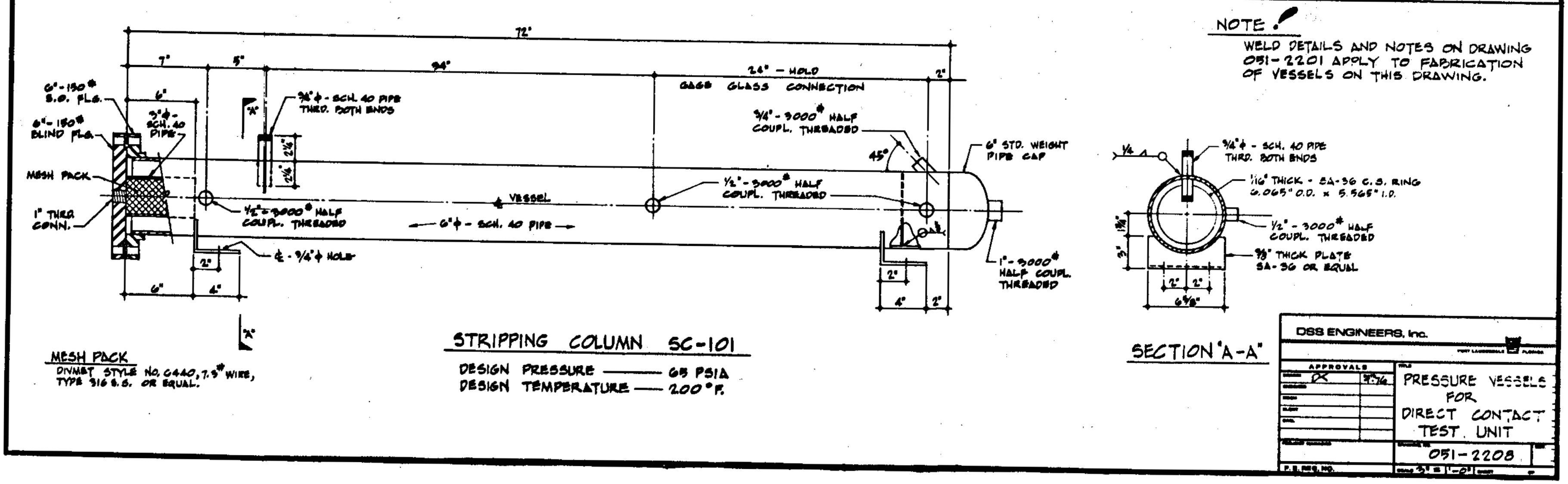


FIGURE 3-10

3.2.5 Degassing Column (Stripping Column)

The high-pressure brine leaving the separator still contains dissolved working fluid. A degassing column was installed in order to flash out this fluid by reducing the pressure. This vessel is 6 ft. in height and 6 ft. 5/8 inches in diameter. The top of the vessel is sealed with a blind flange for accessability. Before the vapors leave the outlet at the top, they pass through a knit mesh demister. The brine exits at the bottom of the vessel to the reinjection system. Instrumentation on the degassing column includes temperature and pressure indication, and a bull's-eye sight glass. The degassing column is shown as Figure 3-10.

3.2.6 Pumps

High Temperature Cycles

The working fluid is circulated by a four cylinder John Bean outside packed plunger pump (Model T-04102) rated at 10 GPM capacity. The brine pump is an outside-packed Triplex plunger pump (John Bean Model M-0910) with a rated capacity of 7 GPM. Both pumps are equipped with a nitrogen filled accumulator on the discharge side used to reduce pressure pulsation. Flow rate is controlled by a valve in a by-pass line from the discharge to suction side and relief valves are installed to prevent overpressure damage to the pump or test unit.

Pumping requirements for the isobutane are especially demanding because of the combination of low available net positive suction head and high discharge head in a single pump. Isobutane liquid also has extremely low density and viscosity placing further restrictions on the type of pump which can be used.

A third pump is required to pump the effluent brine from the stripping or degassing column to the reinjection system. The pump used is a four cylinder outside packed plunger pump similar to the isobutane circulating pump mentioned above.

Low Temperature Cycle

The isopentane is circulated by a Worthington Model D-812 standard cast iron centrifugal pump, size: 1-1/2" x 1" x 6". The impeller was trimmed to deliver 15 gpm at 142 feet TH and 4.7 feet NPSH. The pump is equipped with a back pull-out mechanical seal and 2HP explosion-proof motor.

The low temperature brine circulating pump is a Worthington Model D-512 cast iron centrifigal pump (size: 1-1/2" x 1" x 5") equipped with a "Noryl" impeller, back pull-out mechanical seal, and 2HP explosion-proof motor.

3.2.7 Pre-Flash Separator

To reduce the brine temperature to 210-212 F for the low temperature isopentane cycle a pre-flash separator was installed in the brine inlet line. One of the site's flash drums was modified for this purpose. The drum was piped to maintain a constant head to the suction site of the brine pump. A level gauge was installed to monitor the brine level in the tank.

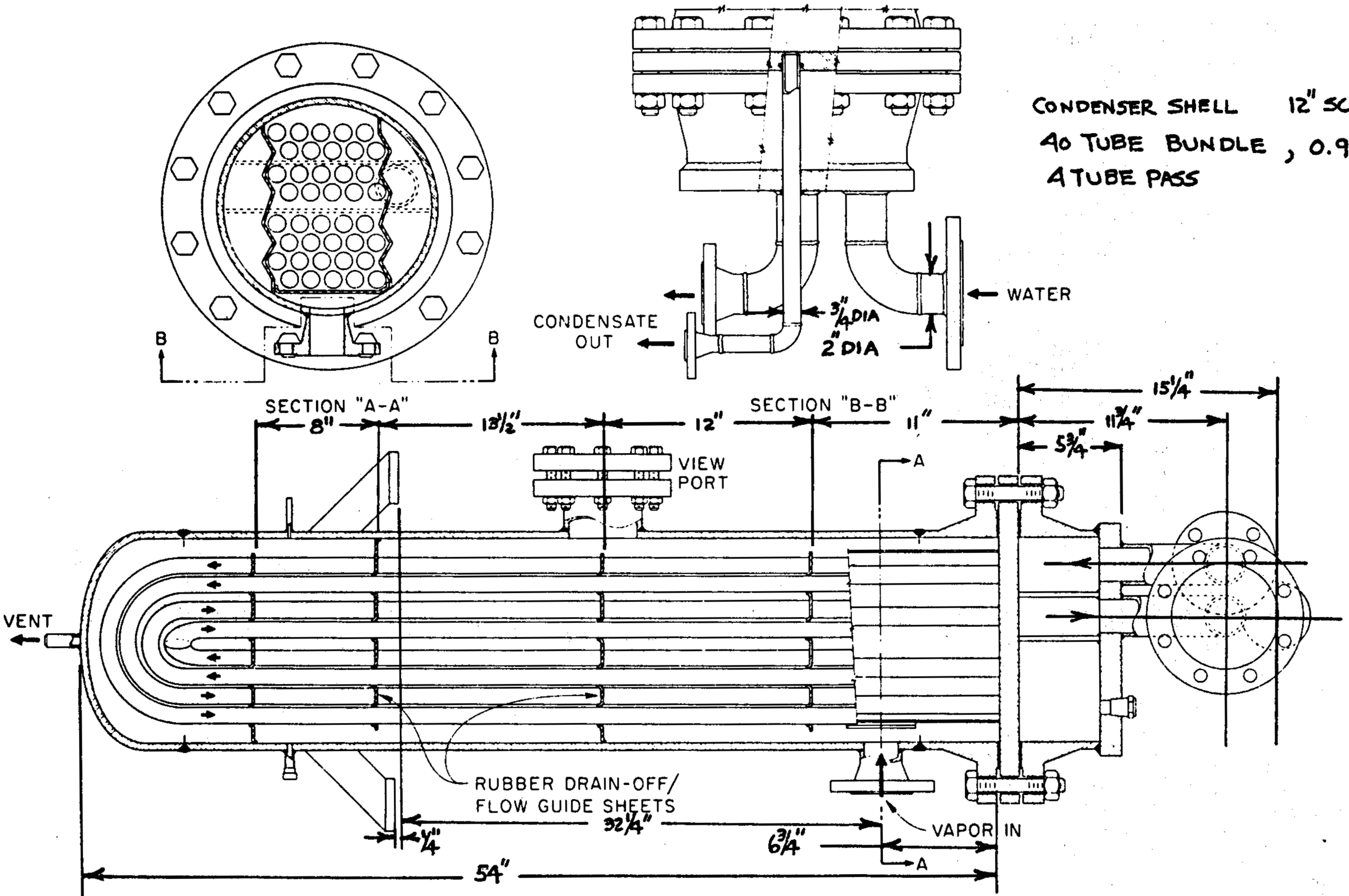
3.2.8 ORNL Condensers

Performance tests were conducted on two separate condensers by ORNL personnel during the time frame of these contracts: 1) a two-tube vertical condenser, and 2) a forty-tube vertical condenser.

The forty-tube Oak Ridge condenser is a stainless steel vertical shell-and-tube type. Both fluted aluminum and smooth tubes were tested. The tubes were four feet long, with an overall total heat transfer area of approximately forty square feet. The shell side (work fluid) of the condenser is designed to handle noncondensables by vapor sweeping across the tube bank in successively smaller flow areas to a purge point. Rubber drain-off tube sheets are used as baffles. The condenser was designed to handle vapor flows of approximately 2000 lbs/hr at a condensing temperature of 110 F. Figure 3-11 (courtesy of Oak Ridge National Laboratory) shows the condenser details. Photographs of the condenser and tube bundle are shown as Figure 3-12 and 3-13. The condenser, together with piping and instrumentation was delivered to the site as a pre-assembled module.

The original ORNL condenser tested during the LBL contractural tests[1] was a two-tube stainless steel vertical shell-and-tube type. It was originally piped parallel to the DCHX condenser to receive part of the isobutane vapors exiting the turbine. These vapors would be condensed on the outer surface of the tubes and returned to the DCHX hotwell by gravity. Figure 3-14 is a photograph of the two-tube condenser during the early testing.

As part of the follow-up test program the two-tube condenser was installed in the forty tube condenser module. Figure 3-15 is a photograph of that module.



CONDENSER SHELL 12" SCH 40 PIPE
40 TUBE BUNDLE, 0.9 in I.D.
4 TUBE PASS

Fig. 3-11 Experimental vertical fluted tube condenser for isobutane.

COMPLIMENTS OF OAKRIDGE NATIONAL LABORATORY

PHOTOGRAPHS OF THE OAK RIDGE NATIONAL LABORATORY 40-TUBE VERTICAL SHELL-AND-TUBE CONDENSER

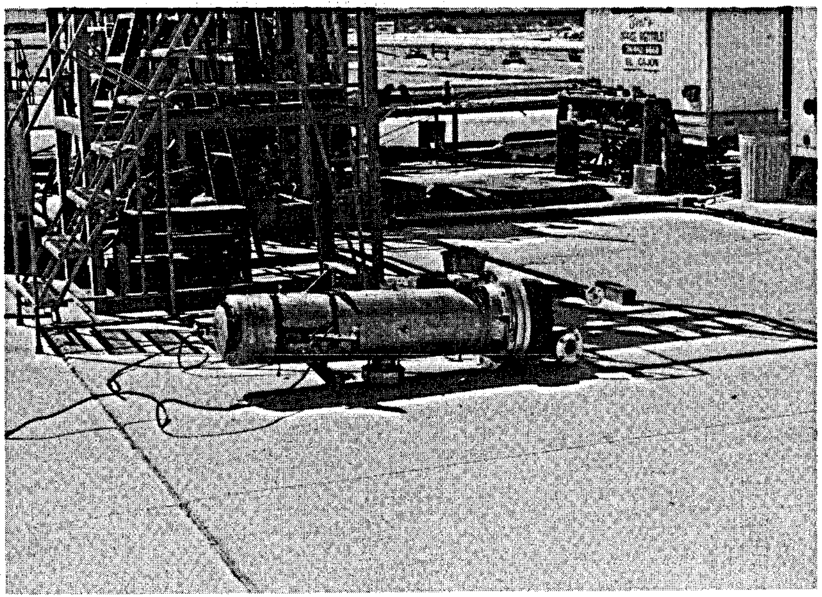


FIGURE 3-12

THE ORNL 40-TUBE CONDENSER ASSEMBLED

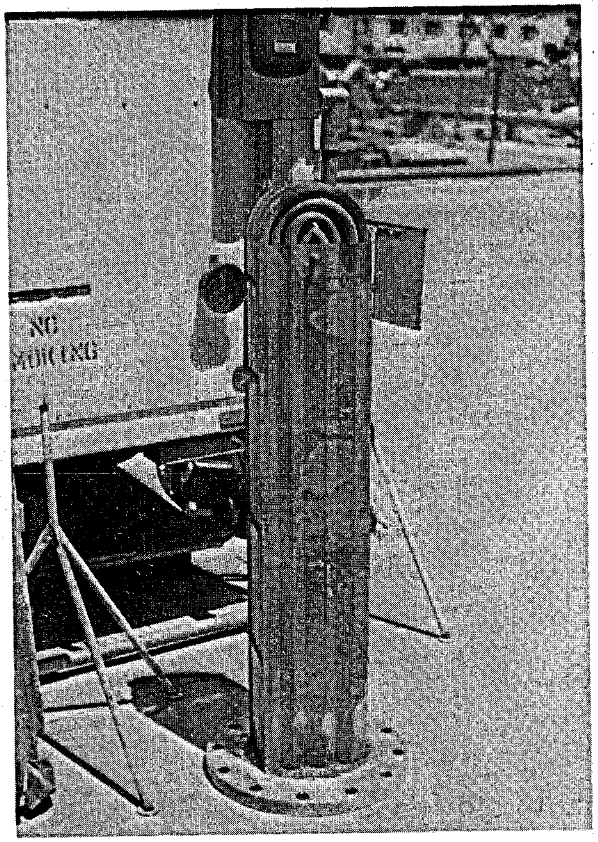


FIGURE 3-13

A SMOOTH TUBE BUNDLE

PHOTOGRAPHS OF THE OAK RIDGE NATIONAL LABORATORY CONDENSERS

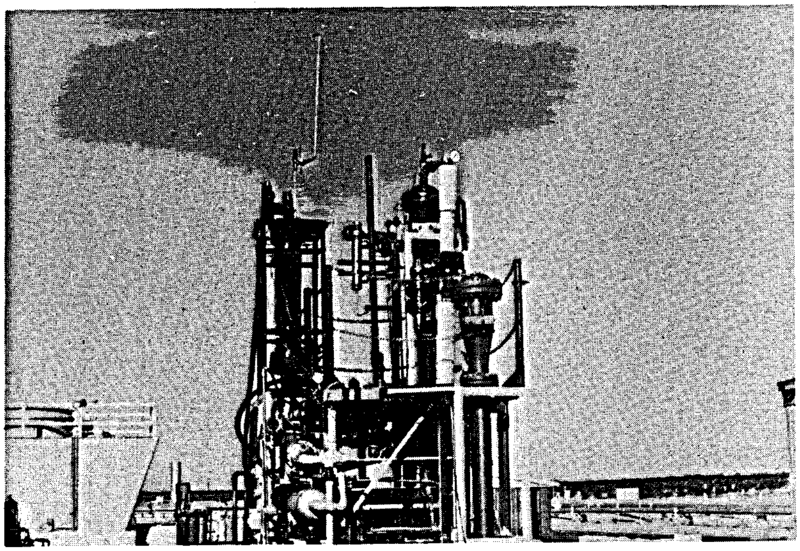


FIGURE 3-14

THE ORNL TWO-TUBE CONDENSER AS INSTALLED DURING THE ORIGINAL TESTING

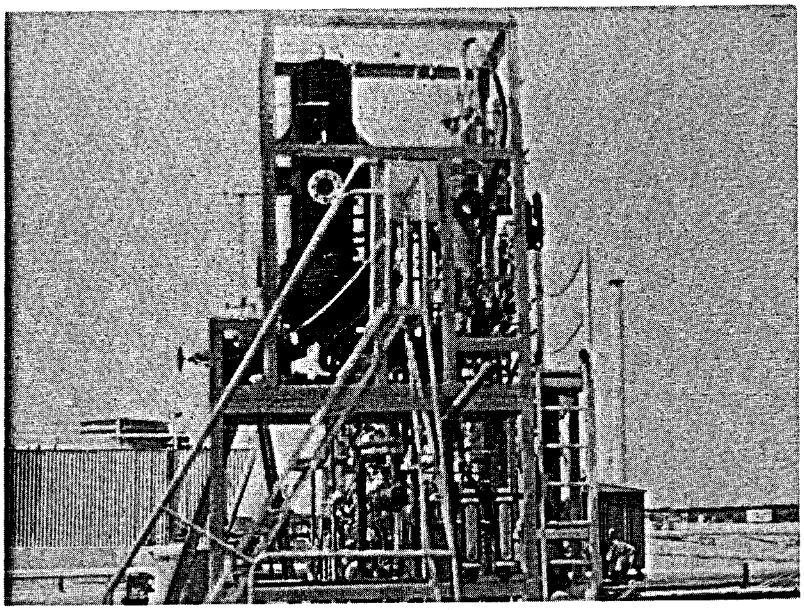


FIGURE 3-15

THE ORNL TWO-TUBE CONDENSER INSTALLED AS PART OF THE 40 TUBE CONDENSER MODULE to approximately atmospheric pressure by spraying into the degassifier where most of the working fluid dissolved in the brine is flashed out. Finally, the brine is sent to the reinjection system.

A "hairpin" surface type heat exchanger is included in the circuit. This can be used to vaporize working fluid in lieu of the DCHX thus providing moisture free, non-condensable free, vapor for calibration.

Basically, two reverse Rankine closed cycles were operated during this series of testing: (1) a high temperature cycle using brine supplied at well head temperatures of 320-325 F, and (2) a low temperature cycle using brine preflashed to 210-212 F. High temperature cycles were run using both isopentane and isobutane as working fluids. The low temperature cycle was run using isopentane. P & I diagrams for the high temperature cycles and low temperature cycle are included as Figures 3-2 and 3-3 respectively.

Some major modifications were necessary in switching from the high temperature cycles to the low temperature cycle. First, a flash drum was piped into the brine inlet line which allowed the hot brine to be flashed to atmospheric pressure at 210-212 F. Because of the suction head limitations characteristic of the positive displacement pumps used for the high temperature-high pressure cycles, both the brine feed pump and working fluid circulating pump were replaced by new centrifugal pumps. The inlet and outlet process piping of the DCHX were changed from one inch tubing to one inch standard schedule 40 pipe to minimize pressure drop. Also, additional thermocouples were installed in the DCHX for a more accurate temperature profile determination.

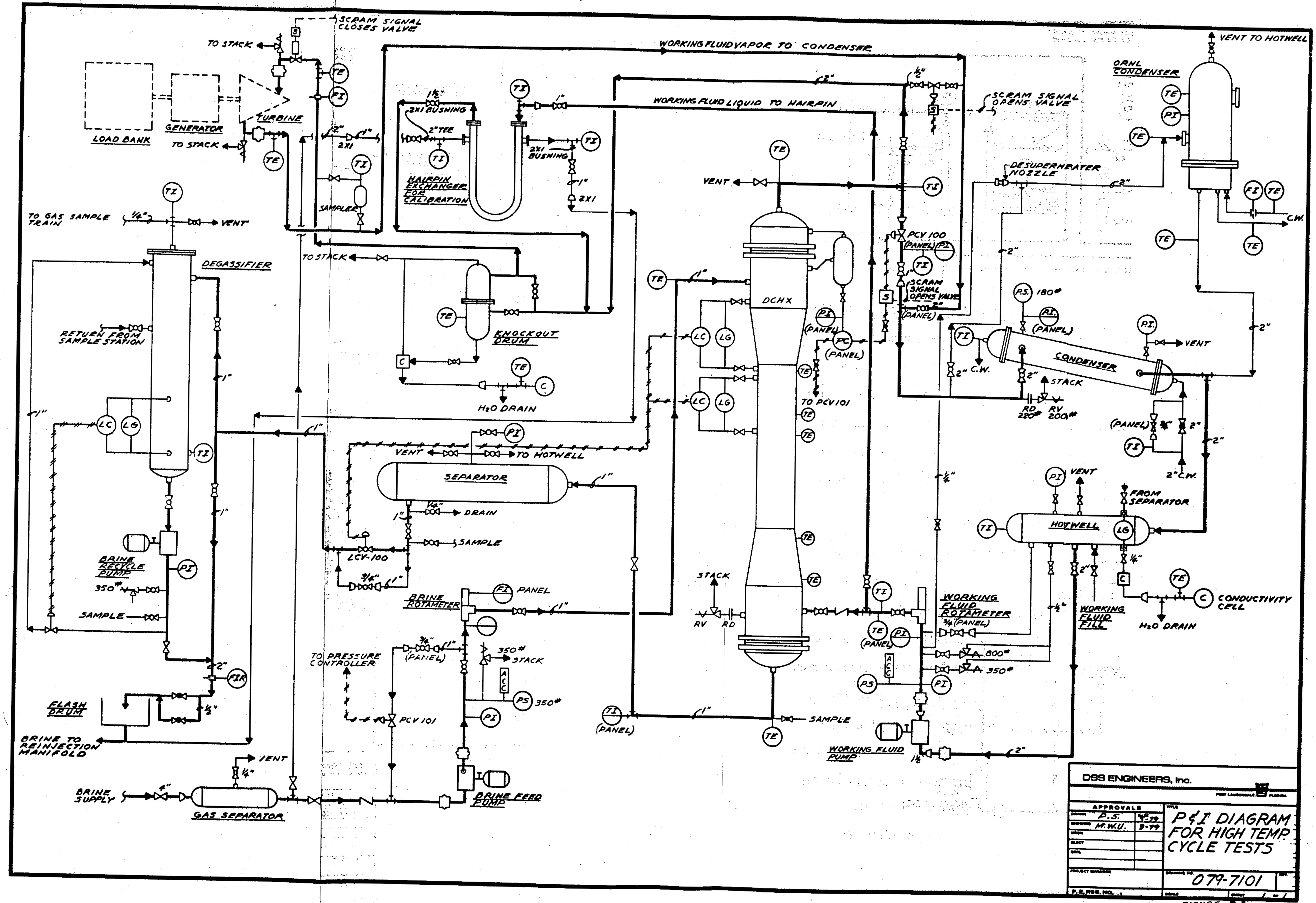

SCRAM SIGNAL CLOSES VALVE
TO STACK
LOAD BANK
GENERATOR
TURBINE
TO STACK
2"
1"
2X1
SAMPLER
1½"
2X1 BUSHING
2" TEE
HAIRPIN EXCHANGER FOR CALIBRATION
2X1 BUSHING
1"
2X1
WORKING FLUID VAPOR TO CONDENSER
WORKING FLUID LIQUID TO HAIRPIN
½"
SCRAM SIGNAL OPENS VALVE
VENT TO HOTWELL
ORNL CONDENSER
C.W.
DESUPERHEATER NOZZLE
TO GAS SAMPLE TRAIN
¼"
VENT
DEGASSIFIER
TO STACK
KNOCKOUT DRUM
H2O DRAIN
RETURN FROM SAMPLE STATION
VENT
DCHX
PCV 100
(PANEL)
SCRAM SIGNAL OPENS VALVE
(PANEL)
PC (PANEL)
TO PCV101
P.S. 180#
P.I. (PANEL)
CONDENSER
2" C.W.
STACK
RD 220#
RV 200#
VENT
(PANEL) ¾"
2" C.W.
VENT
TO HOTWELL
SEPARATOR
¼" DRAIN
SAMPLE
LCV-100
¾"
BRINE RECYCLE PUMP
350#
SAMPLE
FIR
FLASH DRUM
½"
BRINE TO REINJECTION MANIFOLD
BRINE ROTAMETER
FI PANEL
TO PRESSURE CONTROLLER
¾" (PANEL)
350# STACK
ACC
PS 350#
PCV 101
STACK
RV
RD
TI (PANEL)
SAMPLE
TE (PANEL)
WORKING FLUID ROTAMETER
¾ (PANEL)
800#
350#
WORKING FLUID PUMP
1½"
HOTWELL
FROM SEPARATOR
WORKING FLUID FILL
CONDUCTIVITY CELL
H2O DRAIN
BRINE SUPPLY
4"
¼"
GAS SEPARATOR
BRINE FEED PUMP
DSS ENGINEERS, Inc.
FORT LAUDERDALE, FLORIDA
APPROVALS
DRAWN P.S. 3-79
CHECKED M.W.U. 3-79
DESIGN
ELECT
CIVIL
PROJECT MANAGER
P.E. REG. NO.
TITLE
P&I DIAGRAM FOR HIGH TEMP CYCLE TESTS
DRAWING NO. 079-7101
REV.
SCALE
SHEET 1 OF 1

FIGURE 3-2

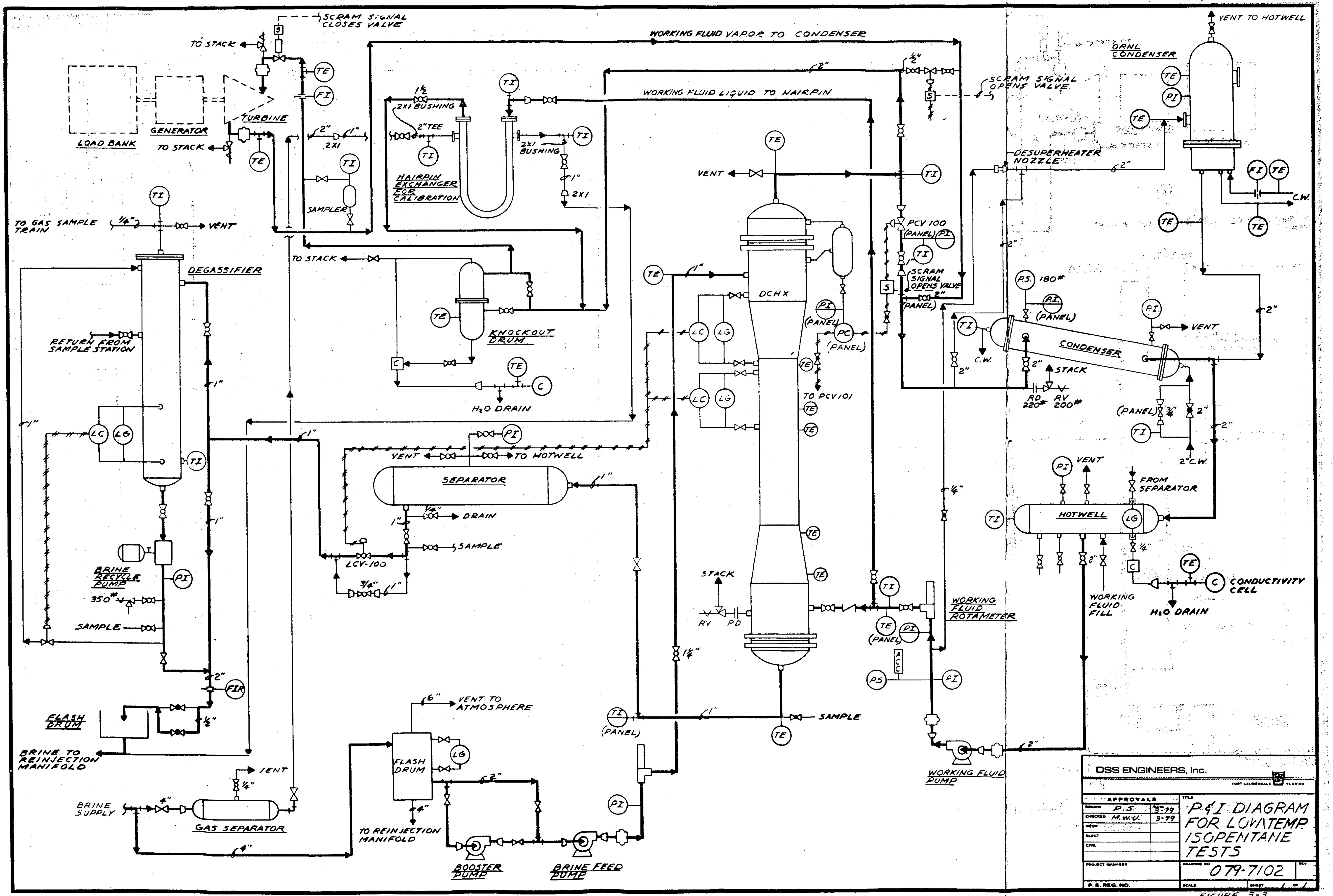

SCRAM SIGNAL CLOSES VALVE
TO STACK
S
TE
FI
LOAD BANK
GENERATOR
TURBINE
TO STACK
TE
2"
1"
2X1
TI
SAMPLER
WORKING FLUID VAPOR TO CONDENSER
WORKING FLUID LIQUID TO HAIRPIN
2"
1½
2X1 BUSHING
2" TEE
TI
TI
TI
2X1 BUSHING
1"
2X1
HAIRPIN EXCHANGER FOR CALIBRATION
TI
TO GAS SAMPLE TRAIN
¼"
VENT
DEGASSIFIER
RETURN FROM SAMPLE STATION
1"
1"
1"
LC
LG
TI
BRINE RECYCLE PUMP
PI
350#
SAMPLE
2"
FIR
½"
FLASH DRUM
BRINE TO REINJECTION MANIFOLD
TO STACK
TE
KNOCKOUT DRUM
C
TE
C
H₂O DRAIN
1"
PI
VENT
TO HOTWELL
SEPARATOR
1"
¼"
DRAIN
1"
SAMPLE
LCV-100
¾"
1"
6"
VENT TO ATMOSPHERE
FLASH DRUM
LG
2"
4"
TO REINJECTION MANIFOLD
VENT
¼"
BRINE SUPPLY
4"
GAS SEPARATOR
4"
BOOSTER PUMP
BRINE FEED PUMP
PI
TE
VENT
TI
TE
1"
DCHX
PI
(PANEL)
PC
(PANEL)
LC
LG
LC
LG
TE
TO PCV101
TE
TE
TE
TE
STACK
RV
PD
1¼"
TI
(PANEL)
1"
SAMPLE
TE
½"
2"
SCRAM SIGNAL OPENS VALVE
S
PCV 100
(PANEL) PI
TI
SCRAM SIGNAL OPENS VALVE
S
2"
(PANEL)
TI
TE
(PANEL)
PI
WORKING FLUID ROTAMETER
ACC
PS
FI
2"
WORKING FLUID PUMP
ORNL CONDENSER
VENT TO HOTWELL
TE
PI
TE
DESUPERHEATER NOZZLE
2"
2"
FI
TE
C.W.
TE
TE
P.S. 180#
PI (PANEL)
TI
C.W.
2"
2"
STACK
RD 220#
RV 200#
CONDENSER
FI
VENT
2"
(PANEL)
¾"
2"
TI
2"
2" C.W.
PI
VENT
FROM SEPARATOR
TI
HOTWELL
LG
2"
¼"
C
TE
C
CONDUCTIVITY CELL
WORKING FLUID FILL
H₂O DRAIN
¼"
DSS ENGINEERS, Inc.
FORT LAUDERDALE FLORIDA
APPROVALS
DRAWN P.S.
DATE 3-79
CHECKED M.W.U. 3-79
MECH
ELECT
CIVIL
PROJECT MANAGER
P.E. REG. NO.
TITLE
P&I DIAGRAM FOR LOW TEMP. ISOPENTANE TESTS
DRAWING NO 079-7102
REV
SCALE
SHEET 1 OF 1

FIGURE 3-3

SECTION 4

ORNL CONDENSER PERFORMANCE EVALUATION TESTING

SECTION 4

ORNL CONDENSER PERFORMANCE EVALUATION TESTING

4.0 INTRODUCTION

The initial ORNL condenser performance tests were conducted using a two-tube stainless steel vertical shell-and-tube condenser. These tests were not specifically contracted, but, by mutual consent, were to be performed simultaneously with the second turbine tests under an LBL contract.

Several operational problems were encountered associated with the ORNL condenser, during these tube tests. The main problem resulting from parallel operation of the DSS and ORNL condensers was that a higher hotwell pressure was experienced than when the ORNL condenser was isolated off line. A possible explanation of this is that the ORNL condenser may have had a lower shell-side pressure drop than the DSS condenser. This, coupled with the fact that no provisions were made to accurately control or measure the vapor flows, could have resulted in the abnormally high hotwell pressure experienced.

As a follow-up condenser test program, DSS was contracted in May of 1978 by Lawrence Berkeley Laboratory (PO #436602-Change #1, Task 10.0) to operate the DCHX-Isobutane system in support of additional ORNL condenser performance tests. These tests involved installation and operation of the ORNL forty-tube condenser. However, because of numerous mechanical, design and operational problems, sufficient data could not be obtained within the time frame allotted for these tests.

Some of the more prominent problems are as follows:

a) Operation downtime was experienced while make-up isobutane was being shipped from Los Angeles. This was due to the fact that bulk storage of isobutane at the site was discouraged by the California State Safety Officer (OSHA).

b) Condenser flooding occurred on a number of occasions. At that time it was thought that this was caused by an undersized condensate return line. The apparent maximum condensate flow rate was 4.5 gpm which restricted operating performance of the DCHX.

c) The overall performance data did not appear to be characteristic of the expected isobutane system. Samples of the working fluid and gas revealed that isobutane was contaminated with approximately 6% propane and at times excessive amounts of non-condensibles.

Based on the inconclusive results obtained during these preliminary tests, follow up experiments were contracted and the results are outlined in this section.

4.1 Scope of Work and Test Loop Preparation

The task was aimed at operating the DCHX loop to provide isobutane vapor to the ORNL condensers in support of the continued performance evaluation test program. It was agreed beforehand that ORNL would set the test conditions and take the necessary data and DSS would attempt operating the loop accordingly. However, spot check data was recorded periodically by DSS personnel to assure loop performance.

The test program as outlined consisted of collecting "baseline" data generated by evaporating isobutane in the "hairpin" surface type heat exchanger and condensing the vapor in the Oak Ridge condenser. Then, the DCHX would be operated to supply vapor at similar pressures and temperatures to the ORNL condenser and the data compared.

Before the tests were initiated some equipment modifications and test loop preparation was required. The 40-tube condenser was dismantled and shipped to San Diego where additional temperature probes were installed along the length of the shell to give a more complete temperature profile. Also, a liquid level indicator was installed such that the onset of any isobutane flooding could be determined before this caused a shut-down condition.

While the condenser modifications were being made, some overhaul and maintenance work was completed about the basic loop to assure safe and continuous operation. All instrumentation was cleaned, inspected and recalibrated. Safety valves, pressure switches, pressure gages and electric relays were tested and/or calibrated. The DSS condenser was dismantled and the tubes were cleaned and inspected. All positive displacement pumps were serviced, repacked and leak tested.

As a final safety check the loop was hydrotested after the ORNL condenser was reassembled and reinstalled. Then all components of the DCHX loop, the condensers, and the "hairpin" surface heat exchanger were evacuated to a vacuum in excess of 28 inches mercury.

This was accomplished in two steps using a compressor to initially evacuate the system to 22 inches vacuum then employing a high capacity vacuum pump to achieve the final vacuum.

As previously mentioned, one of the apparent problems with the data of the initial tests was that it did not agree with laboratory generated data nor did it appear characteristic of an isobutane system. To alleviate this, isobutane certified at a purity of 99+ percent was obtained in sufficient quantity to conduct all anticipated calibration tests. Samples of the isobutane were taken and the purity verified using East Mesa Lab facilities. The analysis results are as follows:

Component	Standard	Cylinder No.1	Cylinder No.2	Cylinder No.3
Isobutane	99.99%	99.950	98.49	99.99
Methane	None	.002	.094	None
Ethane	None	.0007	.054	None
Propane	.00079	.040	1.37	.019

After the system was evacuated, the "pure" isobutane was charged using residual vacuum.

4.2 Condenser Calibration Using Hairpin Exchanger

The calibration loop used to produce the base-line data was: hotwell (liquid isobutane) to hairpin exchanger (liquid to gas) to DSS condenser and/or ORNL condensers (gas to liquid) back to hotwell.

The method used in attempting to collect the base-line data was to initiate isobutane flow through the loop, vaporize in the hairpin exchanger, expand the vapor to the condensing pressure through the "Dummy Nozzle", and condense the vapors in the DSS condenser. When steady state flow conditions were reached, the vapor inlet valve isolating the ORNL condenser from the system would slowly be opened bringing the condenser on line. Then if steady state flow conditions were reached with the condensers operating in parallel, the DSS condenser could be taken off line.

A full set of calibration data was collected by ORNL while spot check data was collected by DSS to ensure operation of the DCHX met the requested test conditions. This spot check data is given in Table 4-1. Photographs of the isobutane condensing in the ORNL condenser are included as Figures 4-2 and 4-3.

4.3 DCHX-ORNL Condenser Performance Tests

When sufficient base-line date was collected, the condenser performance tests were started, using isobutane vapor generated in the direct contact heat exchanger. The test loop used was: hotwell to DCHX to pressure reducing throttle valve to DSS condenser and/or ORNL condenser back to hotwell.

As before, only spot check data was collected by DSS personnel. Typical data is given on Table 4-2.

4.4 General Discussion

Calibration of the ORNL condenser was only partially successful. Although data was obtained while both condensers were being oper-

TABLE 4-1

CONDENSER SPOT CHECK DATA - HAIRPIN - CONDENSER LOOP

						HOTWELL		LIQUID ISOBUTANE		DUMMY NOZZLE DATA				ORNL CONDENSER DATA		
DATE	TIME	BRINE FLOW (lbs/HR)	C_4H_{10} FLOW (lbs/Hr)	BRINE INLET TEMP. (°F)	BRINE OUTLET TEMP. (°F)	HOTWELL PRESS. (psig)	TEMP. (°F)	TEMP. (°F)	PRESSURE (psig)	UPSTREAM VAP.TEMP. (°F)	UPSTREAM VAP.PRESS. (psig)	DOWN-STREAM VAP.TEMP. (F)	DOWN-STREAM VAP.PRESS. (psig)	VAP. TEMP. (F)	CONDENSATE PRESS. (psig)	CONDENSATE TEMP. (F)
					DSS CONDENSER ONLY											
9/8	1:30PM	1050	1234	312	125	52	85	88	140	158	140	153	50	NOT RECORDED		
	2:00PM	1050	1234	312	125	51	85	88	140	165	140	160	50	"	"	"
					DSS CONDENSER PLUS ORNL CONDENSER											
	2:25PM	1050	1234	313	125	51	86	88	125	200	140	167	50	"	"	"
					DSS CONDENSER ONLY											
9/12	2:00PM	1367	1988	315	125	70	100	90	70	167	140	151	70	"	"	"
					DSS CONDENSER (WATER OFF) PLUS ORNL CONDENSER (90 GPM COOLING WATER), NO DESUPERHEATER											
		1057	1234	315	125	66	100	100	120	153	140	140	66	145	60	NOT RECORDED
					DESUPERHEATER ON											
	4:00PM	793	1254	316	125	62	98	74	120	160	140	145	60	136	60	" "
					DSS CONDENSER ONLY											
9/13	8:15AM	1168	1256	318	140	63	96	70	135	263	160	223	60	203	60	" "
					DSS CONDENSER (WATER OFF) PLUS ORNL CONDENSER (50GPM CW), DESUPERHEATER ON											
	8:45AM	1327	1256	318	140	68	100	74	175	214	155	234	65	220	65	" "
	9:05AM	1327	1256	318	140	66	100	71	150	262	170	234	65	220	615	" "

TABLE 4-1 (cont.)

CONDENSER SPOT CHECK DATA - HAIRPIN - CONDENSER LOOP

						HOTWELL		LIQUID ISOBUTANE		DUMMY NOZZLE DATA				ORNL CONDENSER DATA		
DATE	TIME	BRINE FLOW (lbs/HR)	C_4H_{10} FLOW (lbs/HR)	BRINE INLET TEMP. (°F)	BRINE OUTLET TEMP. (°F)	PRESS (psig)	TEMP. (°F)	TEMP. (°F)	PRESSURE (psig)	UPSTREAM VAP.TEMP (°F)	UPSTREAM VAP TEMP (psig)	DOWN-STREAM VAP.TEMP (°F)	DOWN-STREAM VAP.TEMP (psig)	VAP. TEMP (°F)	CONDENSATE PRESS (psig)	CONDENSATE TEMP (°F)
					DSS CONDENSER ONLY (HOTWELL LEVEL 4.5 inches) NO DESUPERHEATER											
9/14	7:50AM	2655	2241	316	158	65	97	87	240	210	265	156	58	NOT RECORDED		
	9:00AM	2655	2241	318	160	72	103	87	250	208	272	170	68	151	68	"
					DSS CONDENSER PLUS ORNL CONDENSER (WITHOUT CW) NO DESUPERHEATER (HOTWELL LEVEL 2-3 inches)											
	9:40AM	2655	2307	320	165	66	105	91	240	211	270	177	68	167	69.6	"
	10:00AM	2655	2307	320	165	68	106	93	250	215	275	175	70	167	69.9	"
					DSS CONDENSER PLUS ORNL CONDENSER (WITHOUT CW) DESUPERHEATER ON											
	12:00noon	2650	2224	322	167	71 55	108	97	260	217	280	170	72 68	167	72.4 70.4	" "
					DSS CONDENSER (VAPOR INLET VALVE 1/2 CLOSED), ORNL COND (WITHOUT CW), DESUPERHEATER ON											
						67										
					DSS COND. (VAPOR INLET VALVE 1/2 CLOSED), ORNL COND. CW @ 37 GPM, DESUPERHEATER ON											
						59									59	"
					DSS COND. (VAPOR INLET VALVE 3/4 CLOSED), ORNL COND CW @ ? , DESUPERHEATER ON											
	1:30PM	2649	2282	325	166	60	100	92	265	221	290	188	62	183	58.1	"
	1:45PM	2915	2262	325	165	56	97	90	260	224	285	186	58	182	58.1	"

TABLE 4-1 (cont.)

CONDENSER SPOT CHECK DATA - HAIRPIN - CONDENSER LOOP

						HOTWELL		LIQUID ISOBUTANE		DUMMY NOZZLE DATA				ORNL CONDENSER DATA		
DATE	TIME	BRINE FLOW (lbs/HR)	C_4H_{10} FLOW (lbs/HR)	BRINE INLET TEMP. (°F)	BRINE OUTLET TEMP. (°F)	PRESS (psig)	TEMP. (°F)	TEMP (°F)	PRESSURE (psig)	UPSTREAM VAP.TEMP (°F)	UPSTREAM VAP.PRESS (psig)	DOWN STREAM VAP.TEMP. (°F)	DOWN STREAM VAP.PRESS (psig)	VAP. TEMP. (°F)	CONDENSATE PRESS. (psig)	CONDENSATE TEMP. (° F)
					DSS CONDENSER ONLY											
9/15	9:25AM	2650	2067	323	162	57	97	87	230	227	258	198	60	192	58.8	-
					DSS CONSENSER: VAPOR INLET VALVE 1/2 OPEN, CW INLET TEMP 62 F, CW OUTLET TEMP 67.5 F ORNL CONDENSER: CW AT 90 GPM, CW INLET TEMP 62 F, CW OUTLET TEMP 66 F NO DESUPERHEATER											
	10:00AM	2649	2087	326	162	59	100	91	240	220	263	202	62	190	61.8	-
					CONDITIONS SAME AS ABOVE BUT WITH DESUPERHEATER VALVE CRACKED OPEN											
	10:15AM	2649	2058	326	163	60	100	93	235	230	263	206	64	188	62.2	-
					CONDITIONS SAME AS ABOVE BUT WITH DESUPERHEATER VALVE FURTHER OPENED											
	10:40AM	2649	2056	328	165	61	101	93	235	235	262	210	65	184	63.8	-
					DSS CONDENSER SAME AS ABOVE, CW INLET 69 F ORNL CONDENSER: CW AT 90 GPM, CW INLET TEMP 69 F, CW OUTLET TEMP 73 F DESUPERHEATER VALVE 1/4 OPEN											
	2:00PM	2649	2051	328	167	66.5	107	97	240	240	266	211	71	133	69.4	-
	3:30PM	2649	1863	328	170	64	105	98	225	280	246	245	70	109.5	66.8	-

TABLE 4-1 (cont.)

CONDENSER SPOT CHECK DATA - HAIRPIN - CONDENSER LOOP

DATE	TIME	BRINE FLOW (lbs/HR)	C_4H_{10} FLOW (lbs/HR)	BRINE INLET TEMP. (°F)	BRINE OUTLET TEMP. (°F)	HOTWELL PRESS. (psig)	HOTWELL TEMP. (°F)	LIQUID ISOBUTANE TEMP. (°F)	LIQUID ISOBUTANE PRESS. (psig)	DUMMY NOZZLE DATA UPSTREAM VAP.PRESS (psig)	DUMMY NOZZLE DATA UPSTREAM VAP.PRESS (°F)	DUMMY NOZZLE DATA STREAM VAP.TEMP (°F)	DOWN STREAM VAP.PRESS (psig)	ORNL CONDENSER DATA VAP. TEMP. (°F)	ORNL CONDENSER DATA CONDENSATE PRESS (psig)	ORNL CONDENSER DATA CONDENSATE TEMP. (°F)
DSS CONDENSER (WITH COOLING WATER) PLUS ORNL CONDENSER (NO DESUPERHEATER)																
9/20	10:15AM	2114	1743	316	156	56	93	88	200	215	225	175	58	175	58	95
CONDITION SAME AS ABOVE WITH DESUPERHEATER VALVE SLIGHTLY OPEN																
	11:00AM	2115	1743	315	155	54	90	88	200	210	225	150	56	150	55.5	94
	11:30AM	2115	1743	314	154	54	90	86	200	210	228	155	55	155	54.5	94
	1:00PM	2112	1745	320	157	56	92	99	200	280	230	160	58	160	58.5	95
	2:45PM	1065	1245	306	123	43.5	83	84	120	160	150	87	45	160	45	95

PHOTOGRAPHS OF ISOBUTANE CONDENSING IN THE ORNL 40 TUBE CONDENSER DURING THE HAIRPIN CALIBRATION RUN

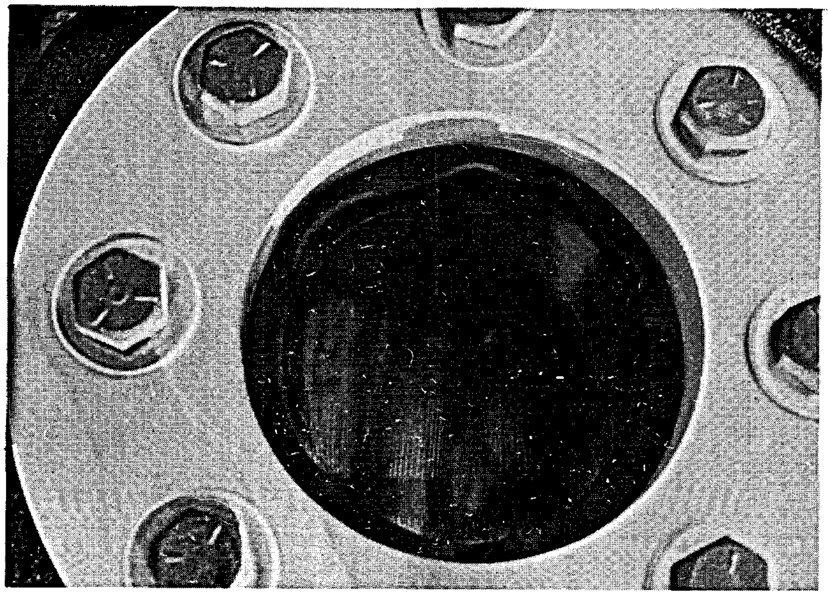


FIGURE 4-2

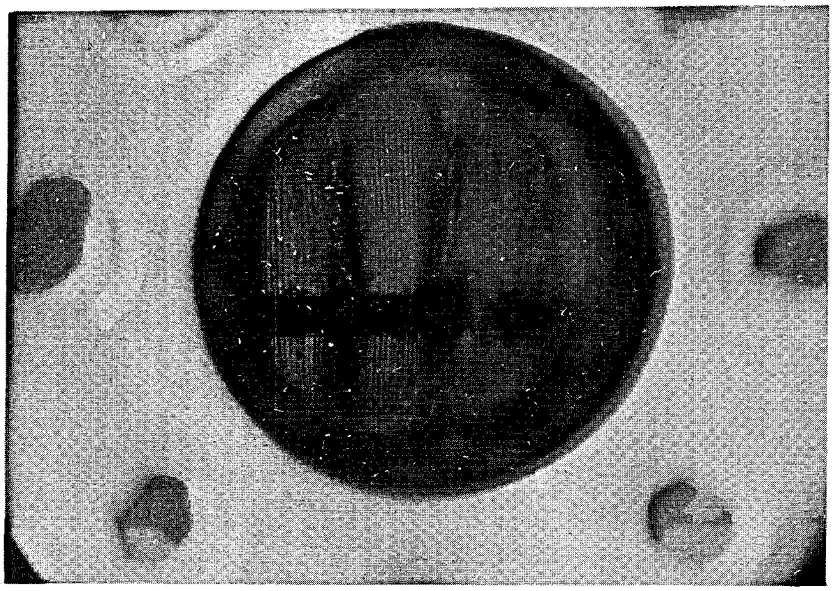


FIGURE 4-3

TABLE 4-2

ORNL CONDENSER SPOT CHECK DATA - BOILER - CONDENSER LOOP

WELL 6-2

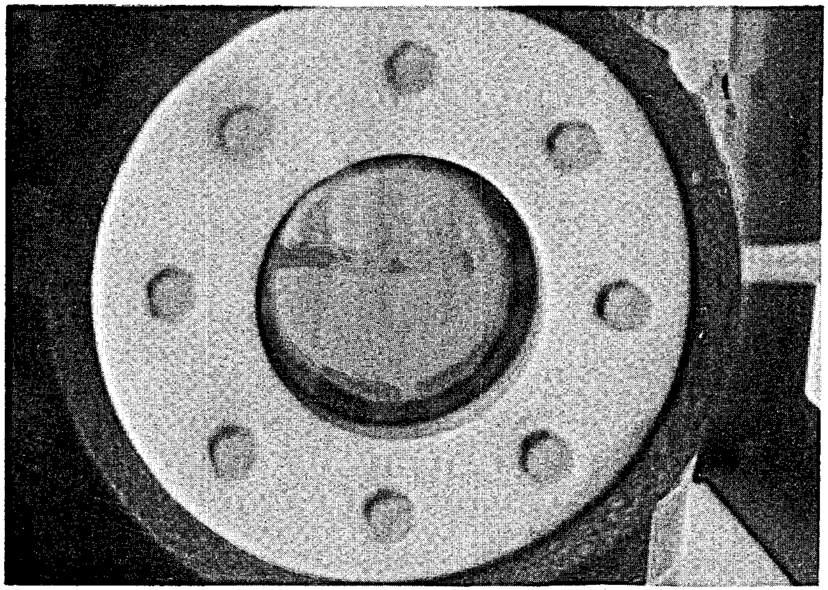
								DSS CONDENSER DATA				ORNL CONDENSER DATA	
DATE DATE	TIME TIME	BRINE FLOW (lbs/HR)	C_4H_{10} FLOW (lbs/HR)	C_4H_{10} BOILER INLET TEMP. (°F)	BOILER VAPOR TEMP. (°F)	BOILER BRINE OUTLET TEMP. (°F)	BOILER PRESS. (psig)	PRESS (psig)	C.W. TEMP. (°F)	C.W. TEMP. (°F)	C.W. FLOW (gpm)	PRESS (psig)	TEMP. (°F)
9/25	2:30PM	2128	2108	93	175	127	150	95	75	80	70	94	160
	3:00PM	2128	2077	115	188	135	170	113	COOLING WATER OFF			109.6	170
			DSS CONDENSER VAPOR INLET VALVE 3/4 OPEN, ORNL ON LINE, DESUPERHEATER SLIGHTLY OPEN										
9/26	10:30AM	1697	1793	88	195	125	145	175	75	70	76	71.4	120

ated in parallel, attempts at isolating the ORNL condenser resulted either in flooding of this condenser or a pressure increase in the hotwell and condenser. The two most plaguing test limiting factors were: 1) the upper design pressure limit of 120 psi on the DSS condenser and hotwell and, 2) upper design temperature limitation of 200 F on the ORNL condenser internals.

Data taken during the actual performance testing using the DCHX was also limited to the case of parallel operation of both the DSS and ORNL condensers. An interesting occurrence at the onset of these tests was the formation of water condensation on the inner side of the 40-tube condenser site glass. A photograph of this is included as Figure 4-4.

Preliminary review of the ORNL condenser field performance data shows much lower values of the condensing coefficient than what had been experienced in the laboratory, under ideal conditions. It had not been established at the time of writing this report whether the problem stems from data collected on the working fluid side or water side of the condenser tubes.

FIGURE 4-4



PHOTOGRAPH OF WATER CONDENSING ON THE INNER SURFACE
OF THE ORNL 40 TUBE CONDENSER SITE GLASS DURING PERFORMANCE TESTING

SECTION 5

DCHX HEAT TRANSFER EXPERIMENTS & CORRELATIONS

SECTION 5

DCHX HEAT TRANSFER EXPERIMENTS AND CORRELATIONS

5.0 INTRODUCTION

Heat transfer performance of direct contact heat exchangers is often expressed as a volumetric coefficient (BTU/Hr-F-Cu ft).

The initial DCHX - Isobutane heat transfer experiments conducted by DSS Engineers, Inc., at Wrightsville Beach, N. C.[2], utilized a separate preheater and boiler so that the performance of each could be evaluated separately. During these early experiments, the volumetric heat transfer coefficient obtained for the preheater varied predominantly with the volumetric isobutane holdup and ranged up to 4000 BTU/HR-F-Cu ft. Boiler heat transfer coefficients ranged as high as 17,000 BTU/HR-F-Cu ft.

Follow up contracts with ERDA and Lawrence Berkeley Laboratory for further development of direct contact heat exchangers in geothermal applications were awarded to DSS Engineers. These involved combining a preheater and boiler into a single continuous column and testing this with isobutane at East Mesa, California. Two objectives of the tests were to determine the volumetric heat transfer coefficients of the preheater and boiler sections and then maximize these by: 1) increasing the mass flow of the two fluids to the "flooding point" while keeping the volume constant and 2) varying the working height of the column keeping the mass flow constant.

The maximum volumetric heat transfer coefficients obtained during the isobutane testing were 7300 BTU/HR-F-Cu ft and 10,200 BTU/HR-F-Cu ft for the assumed preheater and boiler sections respectively. Since these were based on the maximum flow rate of isobutane through the column that could be achieved with the original positive displacement pump, the values may not represent the maximum coefficients that could be obtained for the system.

Since there is no mechanical device or physical barrier separating the vertical column into a preheater and vaporizer and the DCHX is constructed from steel pipe with only a limited number of internal viewing ports, we do not know the exact volumes of the two sections or the corresponding heights required for each section. The boiling and preheating volumetric coefficients above were therefore based on heights calculated using the Letan-Kehat theoretical model for liquid-liquid spray columns (see Appendix A). Observed data during the isobutane runs was in good agreement with this model only after changing the value of a model constant (M) from .83 (suggested by Letan and Kehat) to .70. The test conditions compared were, however, for the relatively small difference in boiler operating level that could be observed.

A major task of this contract was to evaluate the heat transfer performance of the Direct Contact Heat Exchanger using isopentane as the working fluid. This could then be compared with the previous isobutane results. The original task (Task 3.0) made no provisions for the purchase of additional equipment and the test program was devised with the understanding that certain test equipment was not designed for use with isopentane.

One of the objectives of these experiments was to simulate pilot plant process conditions for the Arkansas Power and Light project. This meant evaluating the DCHX-Isopentane performance using low temperature brine, ie, 210-212 F. Accordingly, the East Mesa brine was flashed to within this temperature range. Attempts at trying to pump this low temperature, low pressure brine using the positive displacement brine pump which requires a high suction head were futile. Therefore, a series of high temperature isopentane tests using unflashed brine were run.

An extension to the original contract provided for the procurement of new centrifigal pumps sized to facilitate the low temperature brines (Task 7.0). These were installed and a series of low temperature isopentane cycle tests were then conducted.

Following is a summary of both the high temperature and low temperature isopentane cycle heat transfer experiment results.

5.1 Thermal Performance of the DCHX Using Isopentane

5.1.1 High Temperature Isopentane Cycle

Table 5-1 gives operating data and calculated heat transfer factors for the direct contact heat exchanger typical of the high temperature isopentane cycle runs. The isopentane "boiling" temperatures shown on the table were calculated from a heat balance about the liquid-liquid or preheat section of the DCHX. The vapor pressures of isopentane and water are the pressures at saturation calculated assuming an ideal gas relationship. The log mean temperature difference of each section was calculated assuming a hypothetical 5 F pinch point.

TABLE 5-1

TYPICAL ISOPENTANE HEAT TRANSFER OPERATING DATA

and

HEAT BALANCE RESULTS HIGH TEMPERATURE CYCLE

TEST NUMBER	IFF	1CC	5B
Indicated C_5H_{12} Flow (GPM)	3.1	3.1	4.1
Corrected C_5H_{12} Flow (GPM)	2.6	2.6	3.5
C_5H_{12} Flow (lb/hr)	809	809	1076
Indicated Brine Flow (GPM)	3.0	3.0	4.5
Corrected Brine Flow (GPM)	3.5	3.5	5.2
Brine Flow (lb/hr)	1615	1615	2408
Brine Inlet Temp (F)	290	286	286
C_5H_{12} Inlet Temp (F)	74	84	85
Brine Outlet Temp (F)	174	178	181
Boiler Vapor Outlet Temp (F)	231	239	250
Boiler Pressure (psia)	126	125	123
Water Carryover (lb/hr)	19.2	7.6	22.9
Water Vapor Pressure (psia)	15.8	15.8	15.4
C_5H_{12} Sat. Pressure (psia)	108.9	108.9	107.4
C_5H_{12} Boiling Temp (F)*	217	217	214
Q - Preheater (Btu/hr)	68628	63623	83345
LMTD Preheater (F) **	25.4	21.7	25.1
Q Boiler (Btu/hr)	117997	112639	169349
LMTD Boiler (F)	31.7	30.3	30.8
Heat Loss Average (Btu/HR-FT^2-F)***	3.6	3.6	3.6

* Calculated by heat balance about preheat section of DCHX

** LMTD calculated using hypothetical 5 F Pinch Point

***Average heat loss rate experimentally determined for summer month operation

Test 5B is one of the few tests that could be run with higher mass flow rates (ie; 3.5 gpm $C_5H_{12}$and 5.2 gpm brine) and still maintain stable operating conditions. Based on the observed interface level in the DCHX of 10 feet as measured from the isopentane dispersion ring, the volumetric heat transfer coefficient for the preheating section of the column calculated at 1824 BTU/HR-F-Cuft. This was low compared to the previous isobutane runs.

The height of the liquid-vapor interface calculated from the data of test run IFF, Table 5-1, using the Letan-Kehat mathematical model for direct-contact liquid-liquid heat exchangers (see Appendix A) was 10.9 ft (measured above the bottom flange). This compares well with the observed height of 10.7 ft, but this may be somewhat fortuitous. The observed height was not clearly defined because of the presence of bubbles at the interface, and the calculated height involved judicious selection of some of the input parameters, as explained below.

The Letan-Kehat model assumes that the physical properties of the two fluids are constant throughout the process. However, two of the input parameters, R and r, which are functions of the physical properties, are not constant but vary along the column length, as indicated in Figure 5-1. For test IFF the average value of R is 0.838 and r is 0.357. These values might properly be rounded to R = 0.84 and r = 0.36 in view of the nature of the physical property data, but whether the values are rounded is important. The calculated interface height is shown as a function of R and r in Figure 5-2, and at the steep portions of the curves, where the IFF data falls, the

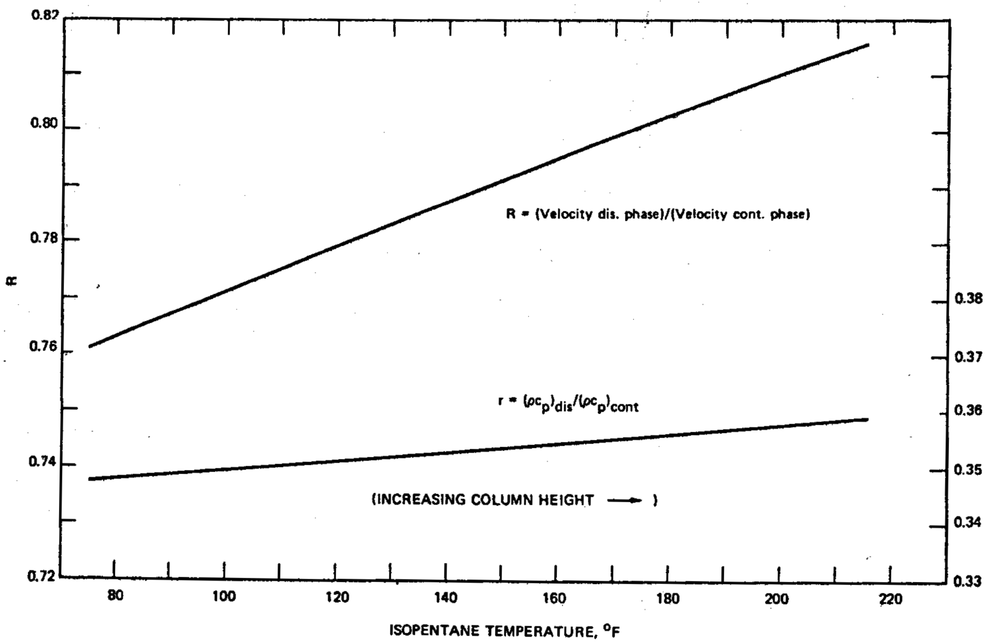


FIGURE 5-1

Variation of R and r with temperature of isopentane and column height in Letan-Kehat model

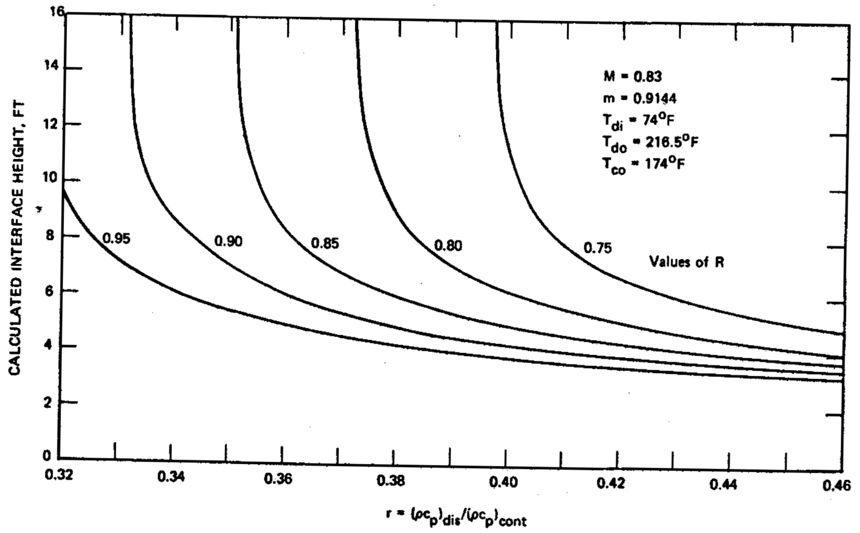


FIGURE 5-2

Calculated interface height as function of R and r, using Letan-Kehat mathematical model

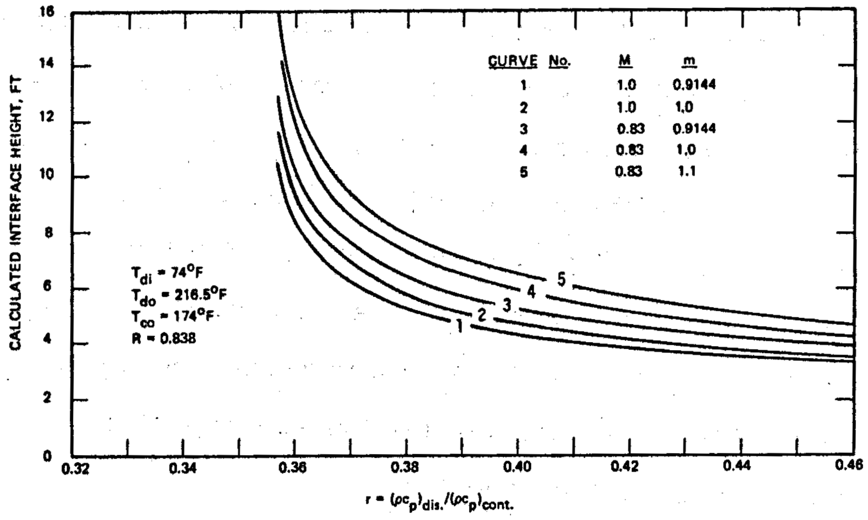


FIGURE 5-3

Effect of constants, M and m, in Letan-Kehat model on calculated interface height in column.

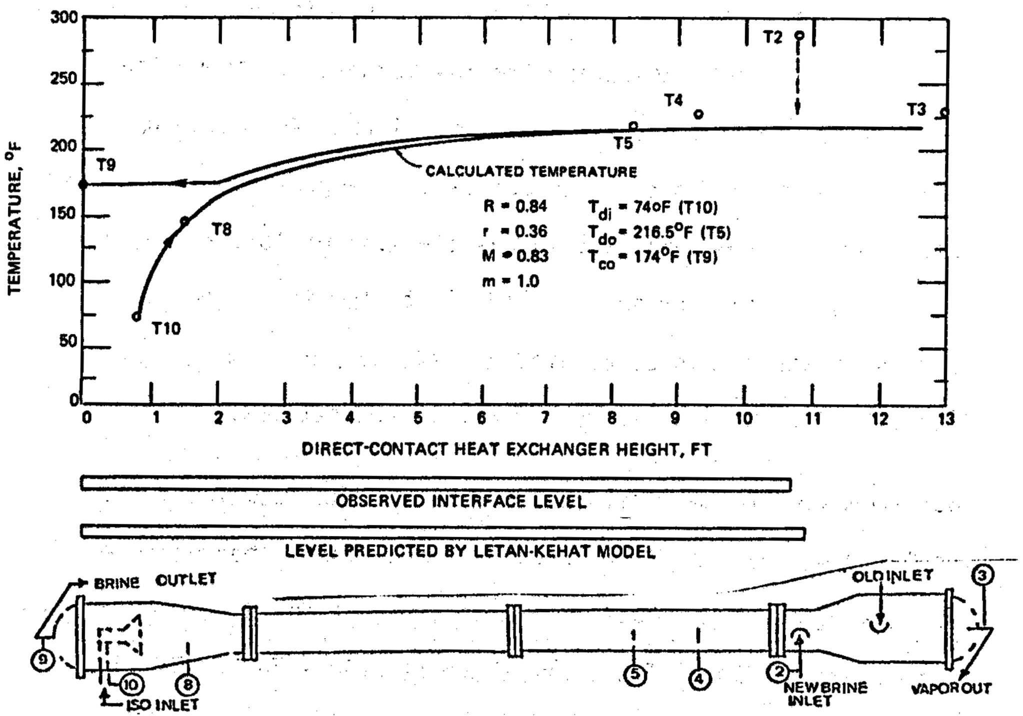


FIGURE 5-4

Observed and calculated temperature relationships and interface levels in direct-contact heat exchanger. (Test IFF, Table 5-1) height is very sensitive to the values of R and r. For example, for $R = 0.838$ and $r = 0.357$, the calculated height is 13.4 ft, but for $R = 0.84$ and $r = 0.34$ the calculated height is 10.9 ft. The effects of two other input parameters, M and m, which are constants having values as suggested by Letan-Kehat, are not as great but are important, as shown in Figure 5-3. The observed and calculated temperature relationships and interface levels for test IFF are compared in Figure 5-4. Although there is fairly good correspondence in this particular set of conditions, the above-mentioned aspects suggest that the Letan-Kehat equation be applied carefully.

5.1.2 Low Temperature Isopentane Cycle

Direct contact heat exchanger operating data from the series of "carryover" tests (see Section 8) was used to determine the thermal performance of the DCHX during the low temperature isopentane cycle runs. Table 5-2 summarizes typical results obtained. Test 7 was included on Table 5 - 2 because, it represents the highest mass flows achieved and test 10 was the test of longest duration. The volumetric heat transfer coefficients for the preheating section of the column calculated as 1742 BTU/HR-F-Cu ft. and 1997 BTU/HR-F-Cu ft for tests 7 and 10 respectively.

Data from Test 10 was fitted to the Letan-Kehat model and the results are shown in Figure 5-5. Two additional thermocouples were installed at available ports in the DCHX for a more accurate temperature profile. These are shown as points T6 and T7 in the figure. As before, the model constants $M = 0.83$ and $m = 1.0$ predicted the level closest to the observed interface.

TABLE 5-2

TYPICAL ISOPENTANE HEAT TRANSFER OPERATING DATA

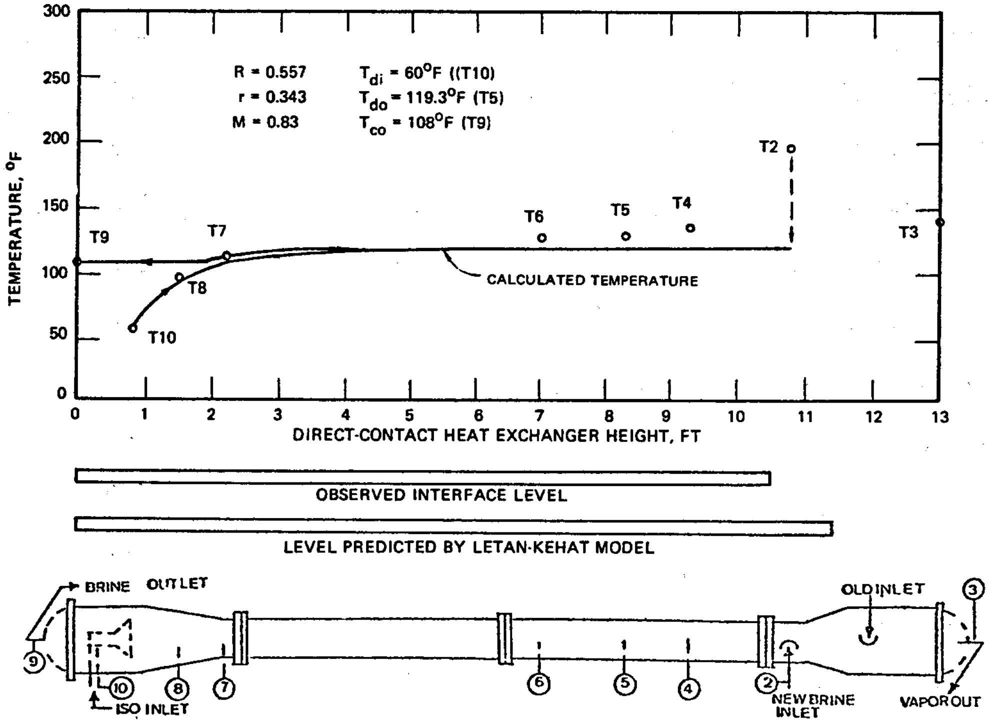
LOW TEMPERATURE CYCLE

TEST NUMBER	7	10
Indicated C_5H_{12} Flow (GPM)	7.5	4.1
Corrected C_5H_{12} Flow (GPM)	7.7	4.1
C_5H_{12} Flow (lb/hr)	2352	1296
Indicated Brine Flow (GPM)	12.5	7.6
Corrected Brine Flow (GPM)	16.9	8.8
Brine Flow (lb/hr)	8142	4248
Brine Inlet Temp (F)	201	201
C_5H_{12} Inlet Temp (F)	85	60
Brine Outlet Temp (F)	113	108
Boiler Vapor Outlet Temp (F)	149	144
Boiler Pressure (psia)	41	40
Water Carryover (lb/hr)	100	78
Water Vapor Pressure (psia)	1.6	1.6
C_5H_{12} Sat. Pressure (psia)	28.2	28.2
C_5H_{12} Boiling Temp (F)*	119	125
Q - Preheater (Btu/hr)	48363	70181
LMTD Preheater (F) **	13.3	22.13
Q Boiler (Btu/hr)	660968	319620
LMTD Boiler (F)	20.1	21.36
Heat Loss Average (Btu/HR-FT^2-F)***	65.4	65.4

* Calculated by heat balance about preheat section of DCHX

** LMTD calculated using hypothetical 5 F Pinch Point

***Average heat loss rate experimentally determined for summer month operation



<u>FIGURE 5-5</u>

Observed and calculated temperature relationships and interface levels in direct-contact heat exchanger in low-temperature isopentane cycle. (Test No. 10, Table 5-2)

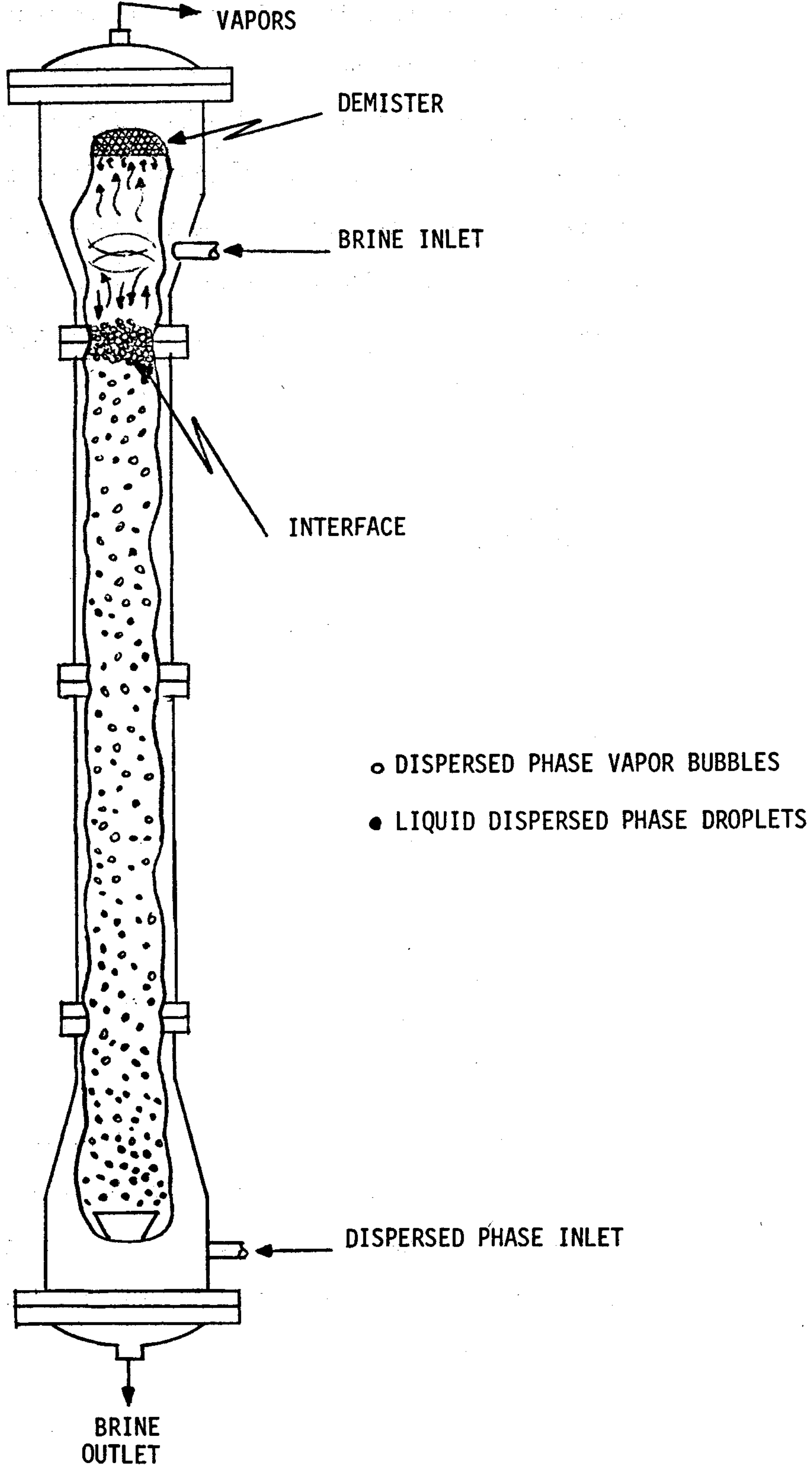
Figure 5-6 shows a hypothetical sketch of how droplets of liquid working fluid may be "vaporized" in the column. In the sketch, droplets of the dispersed phase liquid randomly vaporize to gaseous bubbles and rise through the continuously down flowing brine to a so called interface where liquid-vapor constant changes to vapor-vapor contact. This mechanism has not been verified. at East Mesa because the column has a limited number of viewing ports and these are mostly located at the top. Also, it would probably be difficult to distinquish between droplets of liquid and gaseous bubbles.

In this case, it appears rather senseless to attempt to distinguish between a preheating and boiling section. Rather, for column sizing, an overall volumetric heat transfer coefficient for the combined liquid-liquid and liquid-gas contact section is more meaningful. This should be experimentally determined under laboratory conditions for various working fluids where the maximum value can explicitly be related to "flooding" or "holdup" of the dispersed phase. Then, sizing of the top section or vapor release area becomes a trivial matter.

The temperature profiles predicted by the Letan-Kehat theoretical model for liquid-liquid spray columns appear to be valid for the DCHX. However, the model constants must be adjusted to set the predicted height at the observed interface level. It is not yet known whether the adjusted model will be valid for different operating levels of the DCHX. In addition, the model equations should be adjusted to account for radiant temperature losses and temperature approach at the ends of the column.

Figure 5-6

HYPOTHETICAL MODEL OF PREHEATING AND VAPORIZATION IN A DCHX



5.2 Discussion of Results

The maximum volumetric heat transfer coefficient for the assumed preheating section of the direct contact heat exchanger obtained during the isopentane testing was approximately 2000 BTU/HR-F-Cu ft. This is considerably lower than the coefficient of 7300 BTU/HR-F-Cu ft calculated for the previous isobutane tests using the same equipment. However it is not definite that the 2000 BTU/HR-F-Cu ft represents the maximum that can be obtained without flooding the column.

A problem lies in determining the actual "boiling" temperature of the working fluid. This is critical in determining the so called "pinch point" of the working cycle, the log mean temperature difference for the preheating and boiling sections, the heat transfer coefficients, and in general, the overall cycle thermal performance. The method used in this work was to construct a general heat balance for the liquid-liquid contact section using the known mass flow rates, working fluid inlet temperature and brine outlet temperature and solve for a common "boiling" temperature. This assumes a zero pinch point and infinite heat transfer at the point separating the preheater and boiler.

This calculated "boiling" point temperature does not always agree with the temperature measured at the observed interface. Perhaps because the interface does not represent a true physical division between preheating and boiling in the column.

SECTION 6

ISOPENTANE CONTENT OF EFFLUENT BRINE-DEGASSING COLUMN EXPERIMENTS

SECTION 6

ISOPENTANE CONTENT OF EFFLUENT BRINE-DEGASSING COLUMN EXPERIMENTS

6.0 INTRODUCTION

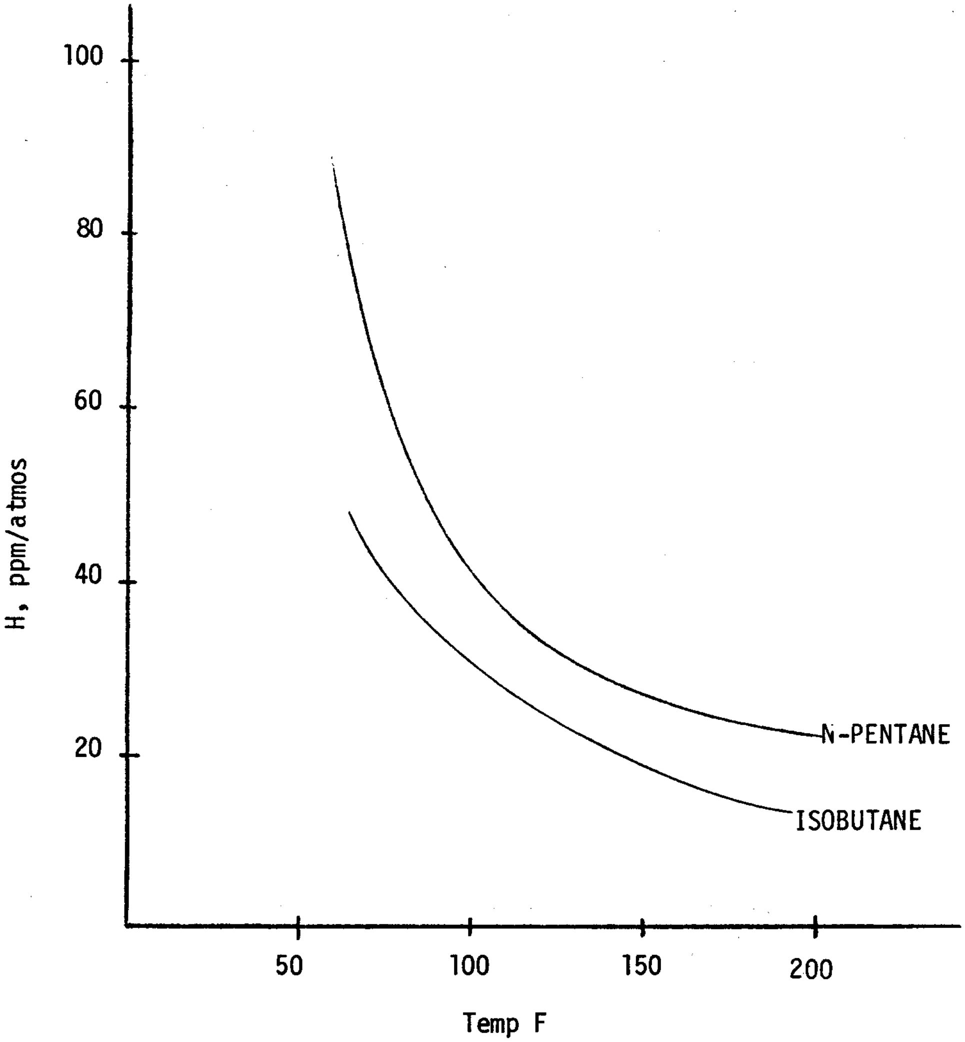
A major concern in development and utilization of direct contact heat exchangers for geothermal power generation is the amount of working fluid contained in the effluent brine. Although the theoretical equilibrium solubility of both isobutane and pentane in the brine existing the DCHX is low at the normal operating temperatures and pressures for these fluids (isobutane-360 ppm and pentane-60 ppm for fresh water and low saline solutions), there is concern that even these small amounts would constitute a significant loss and operating expense in the full scale power plant.

Work has been conducted at the University of Utah[4], to determine the solubility of selected secondary fluids in variable salinity brines. Figure 6-1, constructed from this work, shows the theoretical solubilities of isobutane and n-pentane in low saline solutions as a function of brine temperature and partial pressure of the working fluid above the brine. To our knowledge, the theoretical solubility of isopentane in brine has not yet been determined under laboratory conditions. Therefore, any reference to theoretical solubility of isopentane in this report will be based on the curves for n-pentane shown in Figure 6-1.

Besides solubility, another factor known as"column flooding" may contribute to the loss of working fluid with the effluent brine. For a fixed flow rate of continuous phase (brine), as the

FIGURE 6-1

SOLUBILITIES OF ISOBUTANE AND N-PENTANE IN FRESH WATER



flow rate of dispersed (working fluid) is increased, the rising droplets increase in concentration within the column to a point where more dispersed phase cannot be forced through the column without entrainment of droplets by the continuous phase. This situation is referred to as flooding.

Regardless of the mechanism, a degassing column was designed and installed on the test unit to reduce the amount of working fluid dissolved in the reject brine. This column allows for a drop in brine pressure thus releasing more of the dissolved working fluid as a flashed vapor.

During previous experiments with isobutane, the degassing column operated at an average C_4H_{10} volatalization efficiency of 98%. The amount of dissolved isobutane in the degassing column brine outlet stream ranged between 10.8 and 12.4 ppm. This closely approached the theoretical isobutane solubility predicted by the curve shown in Figure 6-1, ie, 8.4 to 13.2 ppm for the operating conditions of the degassing column.

It was the purpose of this set of experiments to measure the amount of dissolved isopentane in the brine leaving the DCHX and compare these to theoretical values, and if possible, formulate an empirical relationship that my be used to predict these values.

6.1 Equipment and Experimental Procedures

Both high and low temperature isopentane cycles were tested in determining the isopentane concentrations of the effluent brine. A series of tests were run in which the major DCHX operating variables such as flow rates, levels and mass flow ratio were changed and monitored and samples of the DCHX reject brine were collected and analyzed for isopentane. Then a series of experiments were conducted using the degassing column.

The basic test loop flow pattern was: brine to DCHX to separator to degassing column to final flash drum to reinjection manifold.

A physical description of the degassing column is given in Section 3.2.5. The column is shown in Figure 3-10. The apparatus used in the degassing sampling system is shown in Figure 6-2.

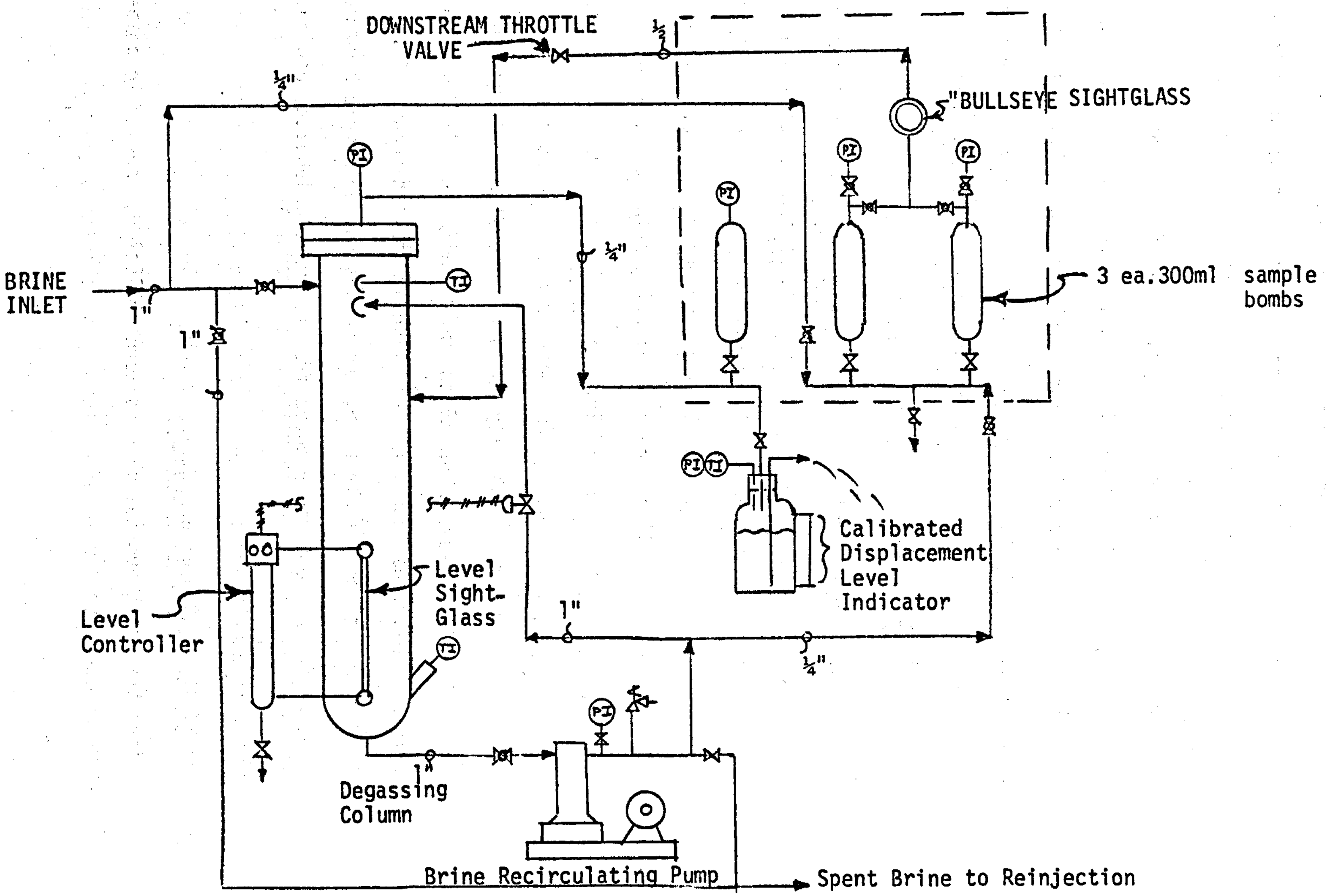
To initiate a degassing column test, brine was started flowing through the direct contact heat exchanger with the reject brine by-passing the degassing column. Once stable operating conditions had been established, the degassing column brine inlet valve was opened dispersing brine into the top of the column through a spray nozzle. The by-pass valve was then closed.

A knit mesh pack of stainless steel was installed in the top of the column to help coalesce any small drops of brine entrained in the vapors. The column contained 5/8 inch polyethylene pall rings to increase the wetted surface area.

Working fluid vapors were released by the reduction of pressure and vented to atmosphere through the gas outlet line. This line terminated at the top of a 5 gallon water bottle calibrated at known volumetric increments. The bottle was fitted with a separate immersion leg with the inlet of the leg at the inside bottom of the

FIGURE 6-2

P&I DIAGRAM FOR DEGASSING COLUMN & SAMPLING TRAIN



bottle and the outlet open to a atmosphere. A 3-hole laboratory cork was used to assure an air-tight seal. This setup was used to measure gas flow rate by water displacement.

Samples were collected in 300 ml stainless steel bombs at three locations to quantify and qualify the operational performance of the column. These consisted of a liquid brine inlet sample, a liquid brine outlet sample and a vapor sample taken from the top of the column. A sample panel was constructed to facilitate the taking of brine samples. A photograph of this is shown as Figure 6-3.

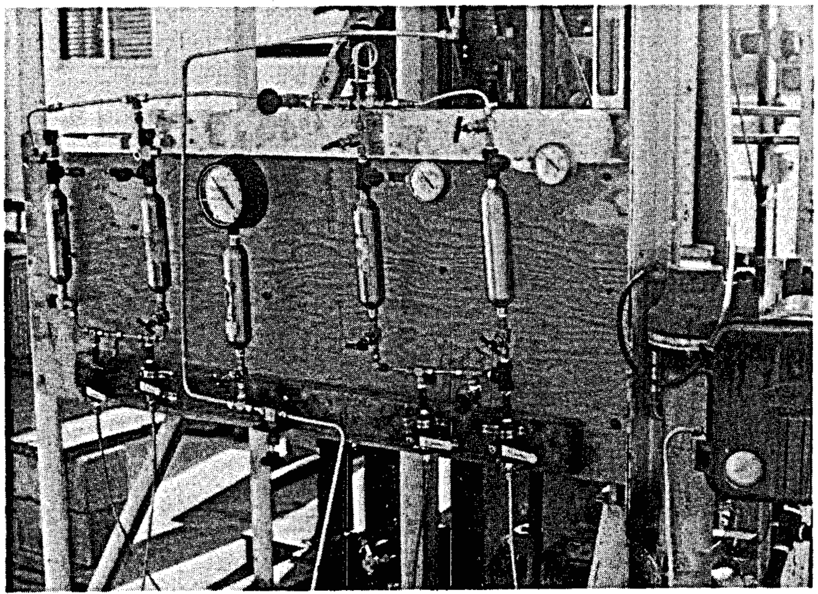
When taking brine samples, a sample bomb was mounted on the panel using swagelok fittings. Like the gas bomb, these sample bombs had been previously evacuated in the site's chemistry lab to approximately 28-29 inches of Hg vacuum. The ball valve for the particular line being sampled was opened and the line purged by allowing brine to flow out the drain valve on the sampling manifold. The drain valve was then closed and the valves on the sampling bomb were opened. The downstream throttling valve was then opened slightly so that brine flowed through the sample bomb. The bull's-eye sight glass was observed to make sure that no bubbles were present, thus ensuring that flashing did not occur. After brine flowed through the sample bomb, the valves on the sample bomb were then closed and the bomb was removed from the panel, labeled, and operating data recorded.

6.2 Laboratory Analysis Procedures

After the brine and gas samples were collected, the sample bombs

FIGURE 6-3

DEGASSING COLUMN SAMPLING STATION



SAMPLES WERE COLLECTED IN 300 ml STAINLESS STEEL SAMPLE BOMBS were taken to the East Mesa Test Site Chemistry lab for analysis.

Each brine sample bomb was weighed empty and then, after sampling, reweighed on the laboratory balance and the net sample weight recorded. The bomb was then connected to a second 300 ml sample bomb which had been evacuated to 28-29 inches Hg in the lab. The brine sample was then placed in the laboratory oven and heated to 100-110 C. After a period of heating, the valves between the sampling bomb and the evacuated bomb were opened allowing gas to flash into the evacuated bomb. The final equilibrium pressure (still under vacuum) of the two bombs was then recorded. After flashing, the bombs were disconnected and the original sample bomb reweighed to determine the amount of vapor flashed. This was assumed to be water vapor. Since the bomb containing the flashed gases was usually under vacuum or slight positive pressure, it had to first be pressurized to 30 psi with helium before connecting to the gas chromatograph for hydrocarbon analysis. The helium provided the driving force necessary to introduce the sample and, since this was used as carrier gas for calibration, was not detected by the G.C. The G. C. analysis along with the laboratory flash information and column operating information at time of sample collection, were then used to calculate the moles of hydrocarbon flashed and concentration in the flowing stream.

The gas samples collected at the degassing column vent were at a pressure of 3-5 psig. Since each sample was injected into the G.C. three times for three separate analysis, the 3-5 psig was borderline low and, therefore, each bomb was pressurized with helium to 30 psi thus providing the necessary driving force. The bomb pressure before and after adding the helium was noted and recorded. This was crucial in determining the total moles of gas in the bomb and consequently the molar hydrocarbon composition.

6.3 Results and Discussion

Because of the limited time alloted for the high temperature isopentane test program only a few runs were made at this condition to determine isopentane solubility. Table 6-1 shows the results of a typical set of data taken about the degassifier column.

As indicated, the isopentane mass balance shows a 94.2% closure. However, the actual measured isopentane solubility values appear much lower than those indicated by the theoretical relationship (Figure 6-1).

A series of tests were run using a low temperature isopentane cycle to determine the amount of isopentane contained in the DCHX effluent brine and the results compared to the theoretical solubilities. Table 6-2 gives this comparison.

Referring to Table 6-2, the measured concentrations ranged from 38 ppm to 274 ppm with an arithmetic mean of 142 ppm. Based on the effluent brine temperature and saturated pressure of pentane for these tests, the theoretical solubilities were determined using Figure 6-1. These ranged from 56 to 70 ppm with a mean of 61 ppm.

A possible explanation for the difference between the measured and theoretical values was that the DCHX may have been operated at or near its "flooding point".

TABLE 6-1

TYPICAL DEGASSING COLUMN RESULTS

HIGH TEMPERATURE ISOPENTANE CYCLE

DCHX Outlet Brine Flow (lbs/hr)	2686
C_5H_{12} Content of DCHX Inlet Brine (ppm)	28.2
C_5H_{12} Content of DCHX Outlet Brine (lbs/hr)	$7.89\text{x}10^{-2}$
Degassifier Inlet Brine Temperature (F)	170
Degassifier Brine Outlet Flow (lbs/hr)	2686
C_5H_{12} Content of Outlet Brine from Col. (ppm)	7.2
C_5H_{12} Content of Outlet Brine (lbs/hr)	$2.0\text{-X}10^{-2}$
Outlet Brine Temperature (F)	164
Measured Degassifier Vapor Flow @ STP (ml/sec)	4.502
C_5H_{12} Vapor Flow Rate (lbs/hr)	$5.42\text{X}10^{-2}$

Closure on C_5H_{12} Mass Balance

$$\% \text{ Closure} = 100 - \frac{(7.89 - 7.43)}{7.89} \times 100$$

$$\% \text{ Closure} = 94.2$$

C_5H_{12} Solubility - Theoretical vs Measured

	Theoretical	Measured
DCHX Brine Outlet	136	28.2
Degassing Column Brine Outlet	20	7.2

TABLE 6-2

THEORETICAL SOLUBILITY OF C_5H_{12} IN DCHX EFFLUENT BRINE

VS

MEASURED CONCENTRATIONS

TEST NO.	MEASURED CONCENTRATIONS (ppm)	THEORETICAL* SOLUBILITIES (ppm)
1	136	56
2	60	56
3	66	70
4	51	62
5	38	58
6	117	60
7	171	62
8	144	62
9	192	61
10	243	58
11	210	62
12	274	61

* Solubility of Pentane (Figure 6-1) at Temperature of Effluent Brine. Pressure Equals Pressure of Saturated Isopentane at Temperature of Effluent Brine.

The two most popular theoretical equations used to predict the "flooding point" of a spray column are: 1) the Minard-Johnson model and the Sakiadis-Johnson model. These relate column diameter and dispersed phase droplet diameter to the superficial velocities of both phases in the column. Both models are presented in Appendix B to this report.

Since both models are dependent on the dispersed phase droplet size, it was necessary to determine an average droplet size to use. This was done by welding a .25 inch diameter cylinder to the inner sides of one of the "bullseye" site glass couplings. When installed in the DCHX, the relative size of the dispersed phase droplets could be determined by comparison with this reference. Photographs of the reference cylinder are shown as Figures 6-4 and 6-5.

Using the reference cylinder, it was found that the average isopentane droplets were between .125 and .150 inches in diameter. Figure 6-6 is a photograph of typical isopentane droplets flowing passed the reference.

To correlate flooding conditions predicted by the models with the direct contact heat exchanger performance, the theoretical equations were rearranged into a form suitable for plotting and then fitted with the appropriate DCHX operating data. The resulting curves and experimental data points are shown in Figure 6-7.

The Sakiadias-Johnson model plotted in Figure 6-7 predicts that the DCHX was being operated at flooding for tests 1,3,7,8,10 and 11. The Minard-Johnson model predicts flooding during tests 3 and 9 only.

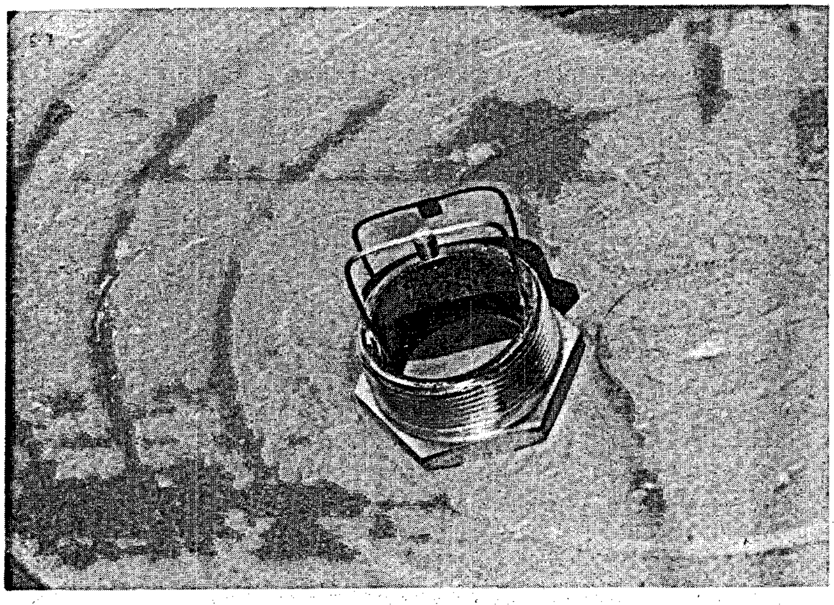


FIGURE 6-4 - .25 INCH ID REFERENCE CYLINDER WELDED TO INNER SIDES OF "BULLSEYE" SITE GLASS

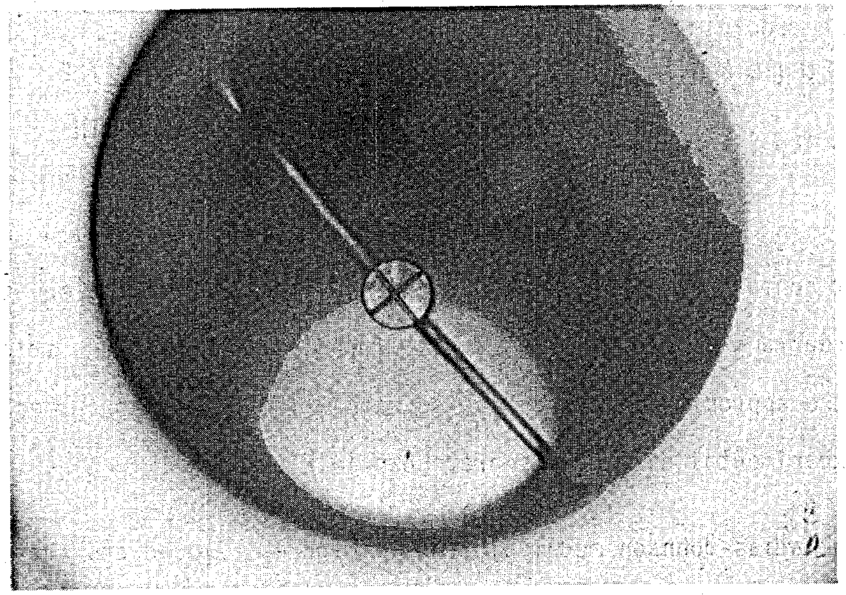


FIGURE 6-5 - REFERENCE CYCLINDER INSTALLED IN DCHX

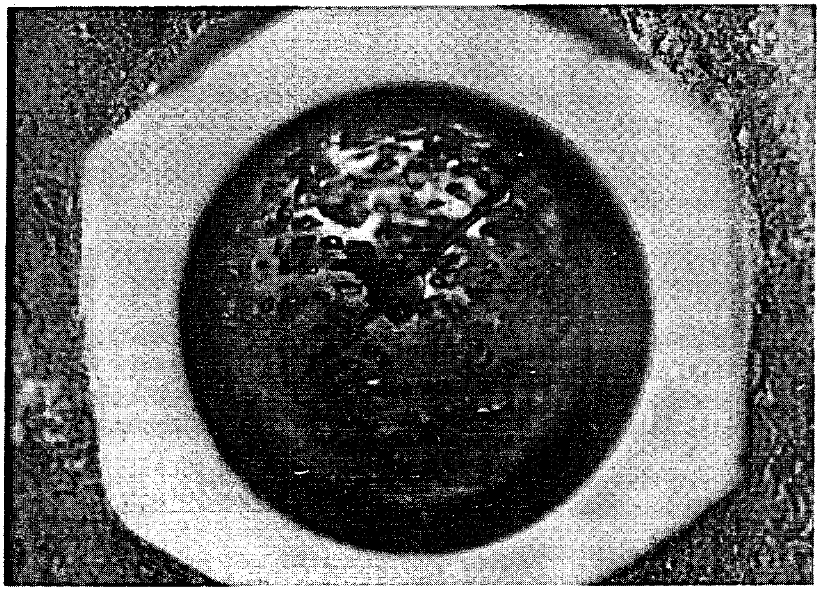
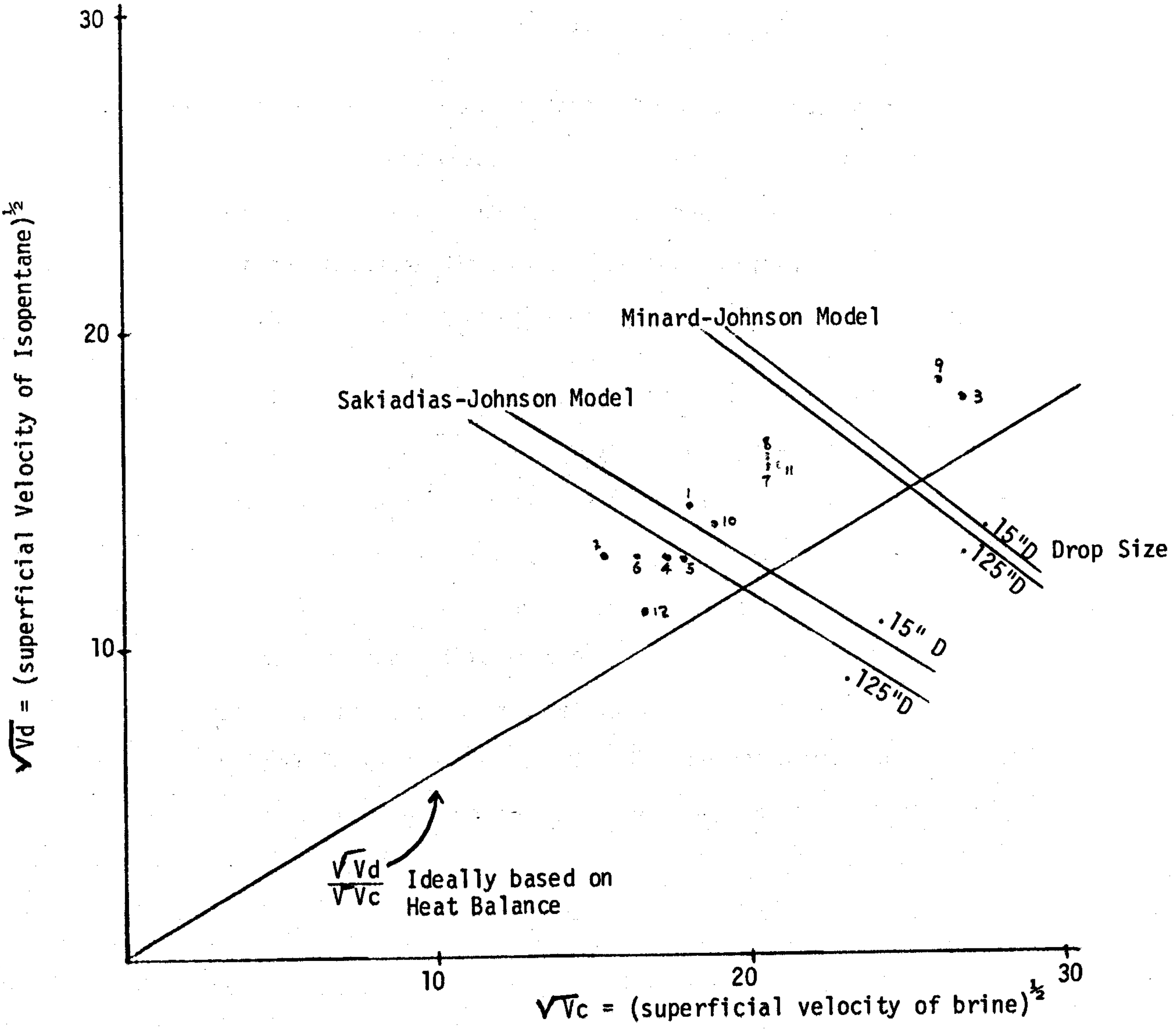


FIGURE 6-6 - DROPLETS OF ISOPENTANE FLOWING PASSED THE .25 INCH ID REFERENCE. TYPICAL DROPLETS RANGED BETWEEN 0.125 AND 0.150 INCHES IN DIAMETER

FIGURE 6-7

SUPERFICIAL VELOCITIES OF ISOPENTANE AND BRINE AT FLOODING



However, the concentrations of isopentane actually measured vary and do not support either model.

An attempt was made to relate the measured isopentane concentration values with various DCHX operating variables such as vapor pressure, temperatures, mass flow rates, levels, etc. Table 6-3 is a summary of these values. No correlation is apparent.

Data was collected about the degassing column during the low temperature isopentane cycle tests. The concentration of isopentane in the feed and brine reject stream was determined and compared to the theoretical predicted values based on the temperatures and operating pressure. Table 6-4 is a summary of these tests.

The isopentane concentrations of the degassing column inlet stream was higher than theoretical in four out of the five tests. The fifth test shows an abnormally low concentration of isopentane in the inlet brine sample. However it was reported that the sample bomb leaked during the laboratory flash of this test. The measured concentrations of isopentane in the degassifier outlet brine were all less than theoretically predicted; in three of the five tests no isopentane was detected. This indicates that the degassing column was doing an adequate job of removing the isopentane as flashed vapors.

TABLE 6-3

DCHX OPERATING VARIABLES AND CONCENTRATIONS OF ISOPENTANE IN EFFLUENT BRINE

Test No.	Mb Brine Flow (lbs/min)	$\dot{M}$F IC$_5$ Flow (lbs/min)	$\frac{\dot{M}F}{Mb}$	Boiler Level* Inches	Boiler Vap Press (psia)	Boiler Vap Temp (F)	Effluent Brine Temp (F)	IC$_5$** Concentration (ppm)
1	65.4	24.3	.371	-8	36	140	102	136
2	46.7	18.8	.403	-11	36	133	103	60
3	141.2	35.8	.254	-4	50	168	127	66
4	60.7	18.5	.304	-1	40	154	113	51
5	64.1	18.5	.288	-1	36	142	107	38
6	52.3	18.5	.353	-1	42	140	112	117
7	84.1	28.4	.338	-3	40	152	113	171
8	84.0	29.0	.345	-2	42	147	113	144
9	135.7	39.2	.288	-5	45	145	118	192
10	70.8	21.6	.305	-4	40	144	108	243
11	84.7	29.0	.342	-3	43	156	114	210
12	55.1	14.6	.264	-4	41	152	115	274

* measured from reference point on DCHX (see Figure

* * measured concentration of isopentane in effluent brine from DCHX

TABLE 6-4

DEGASSING COLUMN TEST RESULTS

LOW TEMPERATURE ISOPENTANE CYCLE

	DEGASSING COLUMN INLET					DEGASSING COLUMN OUTLET			
				ISOPENTANE CONCENTRATION				ISOPENTANE CONCENTRATION	
Test No.	Brine Flow (lbs/min)	Temp. (F)	Pressure (psia)	Measured (ppm)	Theoretical (ppm)	Temp. (F)	Pressure (psia)	Measured (ppm)	Theoretical (ppm)
1	135.7	118	45	192	61	104	15	ND	38
2	70.8	108	40	243	58	105	19	21.8	48
3	84.7	114	43	210	62	106	19	20.6	48
4	55.1	115	41	274	61	108	19.5	ND	47
5	55.0	117	40	16*	60	120	20	ND	43

* leaky sample bomb reported by lab

SECTION 7

FREE CO_2 CONTENT OF BRINE vs CONDENSER & DCHX PERFORMANCE

SECTION 7

FREE CO_2 CONTENT OF BRINE VS CONDENSER & DCHX PERFORMANCE

7.0 INTRODUCTION

It is a well known fact that carbon dioxide, like other "non-condensing" gases found dissolved in geothermal brines, impedes heat transfer throughout the direct contact heat exchanger loop. In particular, the CO_2 released by flashing in the vapor-liquid section of the DCHX travels with the vaporized working fluid and tends to collect in the condenser adversely affecting the film and condensing coefficients. As an example, Othmer[7] has shown that, when as little as one (1) percent of air by volume is mixed with steam entering a surface type condenser, the condensing coefficient falls from 2000 to 1100 with a ΔT between the steam and cooling medium of 20 F. It is also found that during steady stable transfer, the air tends to surround the tube surface setting up a resistance through which the steam must diffuse.

During the initial stages of testing the 40 tube and 2 tube condensers with isobutane at East Mesa, ORNL found that the field performance data did not compare favorably with data collected for the condensers operated under laboratory conditions. One possible explanation was the presence of varying amounts of CO_2 on the shell side of the condenser during the field tests.

For many gases the relationship between the concentration of gas dissolved in a liquid and the equilibrium partial pressure of the gas above the liquid surface may be expressed by Henry's Law. If this were the case at East Mesa, the DCHX feed brine could be preflashed to release Co_2 based on Henry's Law. However, at high pressures or for gases of relatively high solubilities such as CO_2, the direct proportionality of the gas-solubility laws breaks down. The situation is further complicated by the fact that the brines contain soluble carbonates and bicarbonates.

Based on the above, DSS Engineers, Inc., was awarded a one month contract extention by ORNL to operate the DCHX-Isobutane Loop in support of a series of experiments conducted by DSS and Barber Nichols Engineering Company. These experiments were directed at the questions:

a) How does the amount of free CO_2 in the inlet brine to the DCHX affect condenser and DCHX performance?
b) How does the solubility of free CO_2 in the brine vary with pressure and temperature?
c) How can we readily measure the free CO_2 content of the brine?

The following is a summary of the work performed.

7.1 Methodology

The basic procedure for determining the affect, if any, of the free CO_2 content of the feed brine was: preflash the brine under controlled conditions; operate the basic DCHX loop with the ORNL condenser on line; take samples from "key" points in the loop; analyze for CO_2 and; then try to relate the calculated CO_2 concentrations to operating and performance data.

A small flash-separator unit was installed in the brine supply line to remove CO_2 and some steam by flashing to the atmosphere. A control valve was installed on the separator vent line, instrumented and operated to provide a controlled rate of venting in proportion to the overall temperature drop of the inlet and outlet brine streams. Sample taps were installed in the inlet and outlet brine lines of the separator to facilitate sampling and pH measurements. Photographs of the separator and equipment are shown as Figure 7-1 and Figure 7-2.

Figure 7-2 shows a small booster pump installed in the separator brine outlet line. This was added when it was realized that venting the separator lowered the pressure of the brine below the suction head requirement of the DCHX positive displacement brine feed pump.

Figure 7-3 shows the basic test flow scheme and sample point locations.

In-line pH probes were installed in the inlet and outlet brine lines to the separator. This enabled continuous pH monitoring of these streams for correlation with free CO_2 content.

All high pressure samples were collected in 300 ml stainless steel sample bombs. The basic sampling procedure outlined in Section 6-1 was employed.

The bombs containing the vapor samples taken from the top of the direct contact heat exchanger, condenser vents, and new brine inlet flash-separator vent, were connected directly to the gas chromatograph for CO_2 analysis.

Determination of the free CO_2 content of the liquid samples was made using a multi-step process. First, a bomb containing a liquid sample was connected to a second bomb which had previously been purged and evacuated. The bombs were then placed in the laboratory oven and heated to the maximum temperature allowed by the oven. After a period of heating, the valve between the bombs was opened and the system

PHOTOGRAPHS OF THE FLASH-SEPARATOR USED TO VARY THE AMOUNT OF FREE CO_2 IN THE BRINE FEED TO THE DCHX BY CONTROLLED VENTING

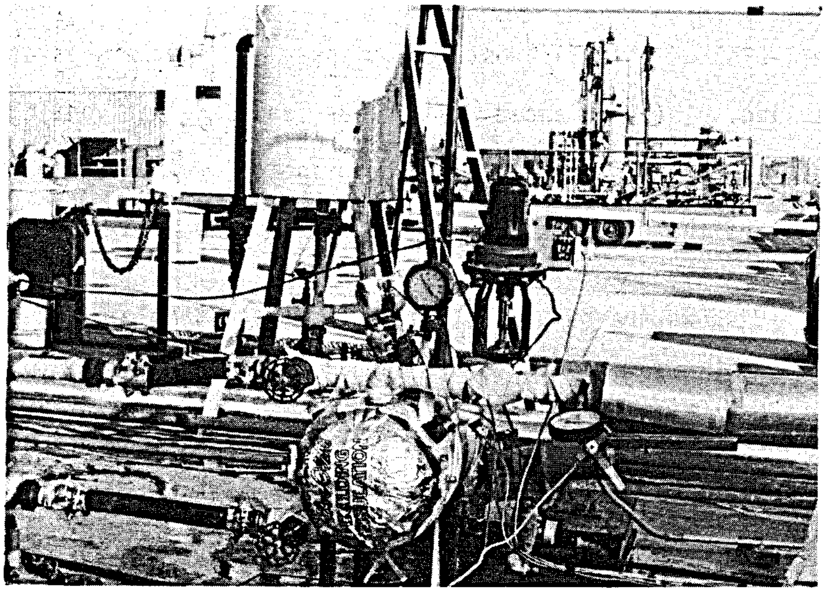


FIGURE 7-1

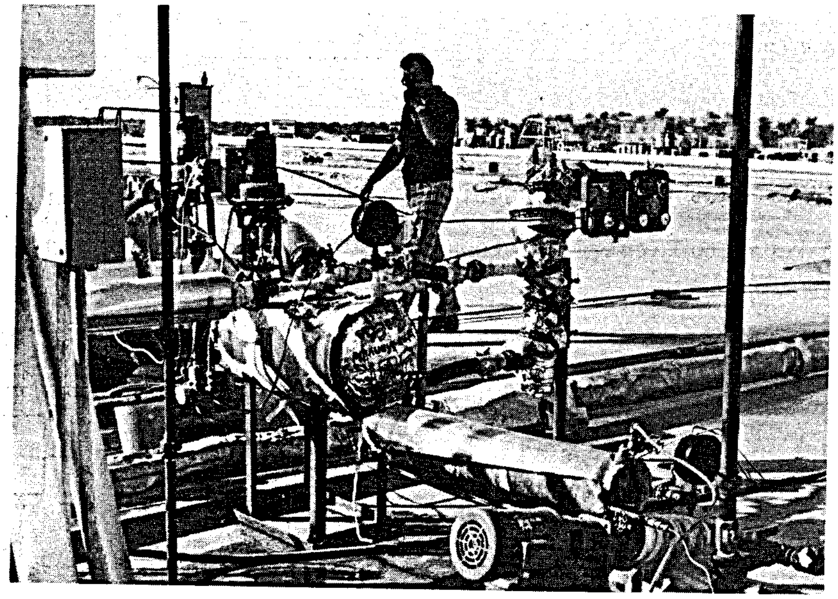
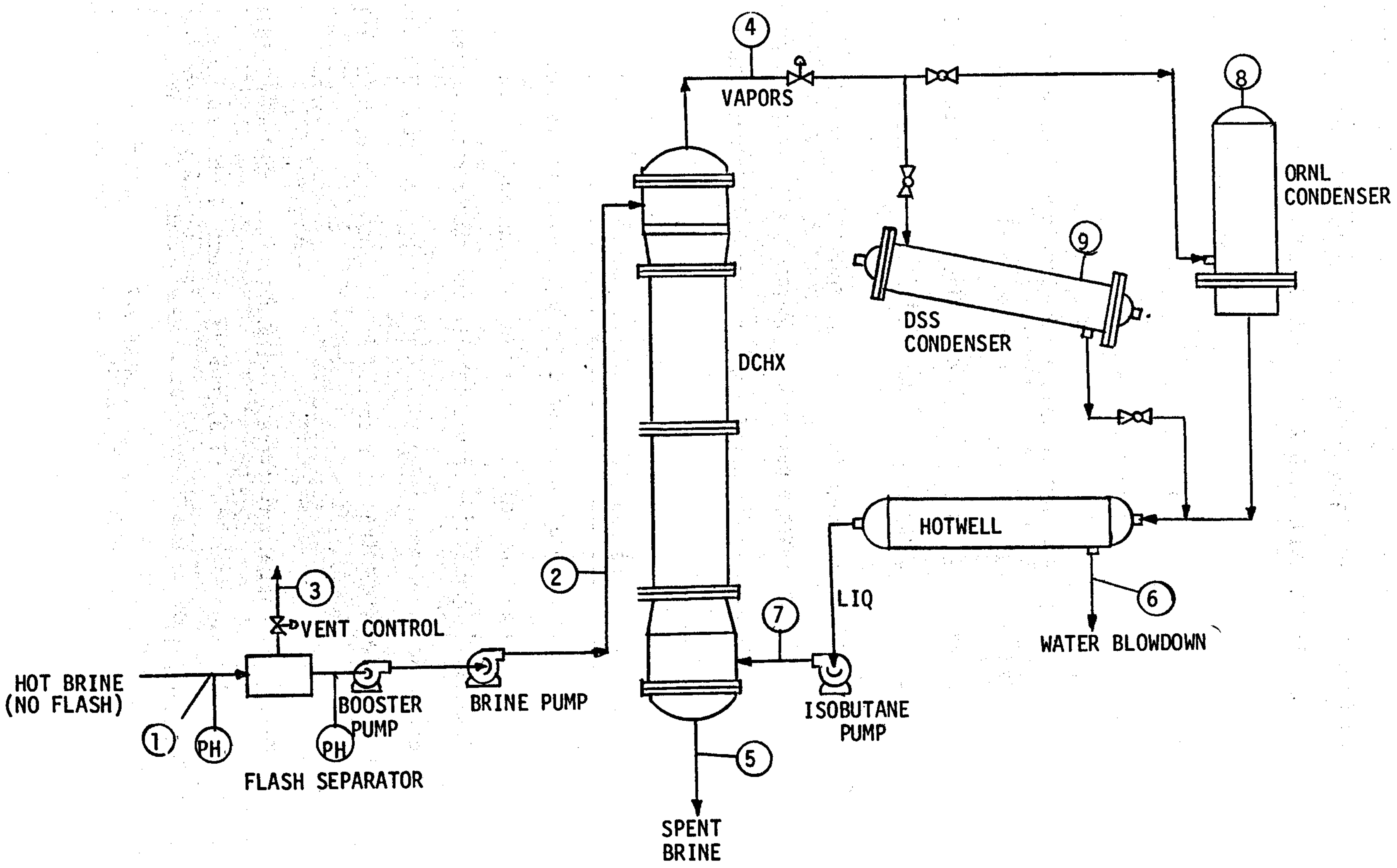


FIGURE 7-2

FIGURE 7-3

BASIC TEST LOOP FOR FREE CO_2 DETERMINATION

SHOWING SAMPLE POINTS



allowed to reach an equilibrium state. The temperature and pressure were then recorded. The valves separating the bombs were then closed and the bombs disconnected from one another.

The bomb containing the flashed vapors usually was at a low pressure or under a vacuum. This was pressurized with helium to 30 psi before connecting to the gas chromatograph for analysis.

After flashing the bomb containing the liquid portion of the original sample was drained to a beaker, the amount of sample remaining was measured and recorded and the pH determined. A pH of below 8.3 indicated the presence of bicarbonate and/or residual CO_2 species.

A method for determining this residual free CO_2 was not definitely known at the onset of these experiments. Therefore, to provide timely results, two methods were investigated using the actual samples. These are outlined below.

Back-Titration Method for Determining Free CO_2

This method is based on standard lab procedures for determining free CO_2 and the three forms of alkalinity in water as outlined in Standard Methods.(8) First, the sample is titrated with a known concentration of dilute sodium hydroxide to the phenolphthalein end point(83) and the amount of hydroxide required is recorded. The main reaction is as follows:

$$CO_2 + OH^- \rightarrow HCO^-{}_3$$

Then, this solution is titrated with dilute hydrochloric acid to the methyl orange end point (4.3) and the amount of HcL recorded. The main reaction is:

$$HCO_3^- + H^+ \rightarrow H_2CO_3$$

These values are then used to calculate the ppm of CO_2. as a function of the excess bicarbonate present.

Barium Hydroxide Method for Determining Free CO_2

The lab set'up used to determine any residual CO_2 in the preflashed brine sample is shown as Figure 7-4. Basically, Nitrogen is bubbled through a predetermined amount of the sample liberating any free CO_2 . This nitrogen purge gas containing the CO_2 is then bubbled through a known concentration of dilute barium hydroxide. The barium hydroxide and CO_2 form the very insoluble precipitate barium carbonate. The barium hydroxide solution containing the barium carbonate is then titrated with HcL to determine the amount of barium hydroxide left unreacted. This information is then used to determine how much CO_2 reacted.

The amount of bicarbonate remaining in the original sample is then determined by adding a known amount of HcL to convert the bicarbonate to CO_2 , purge the CO_2 with Nitrogen into a fresh barium hydroxide solution and proceed to determine the amount of CO_2 as outlined above.

The main reactions are as follows:

$$CO_2 + Ba(OH)_2 \rightarrow Ba\,CO_3 + H_2O$$

$$Ba(OH)_2 + H^+ \rightarrow Ba^{\theta} + H_2O$$

7.2 Results and Discussion

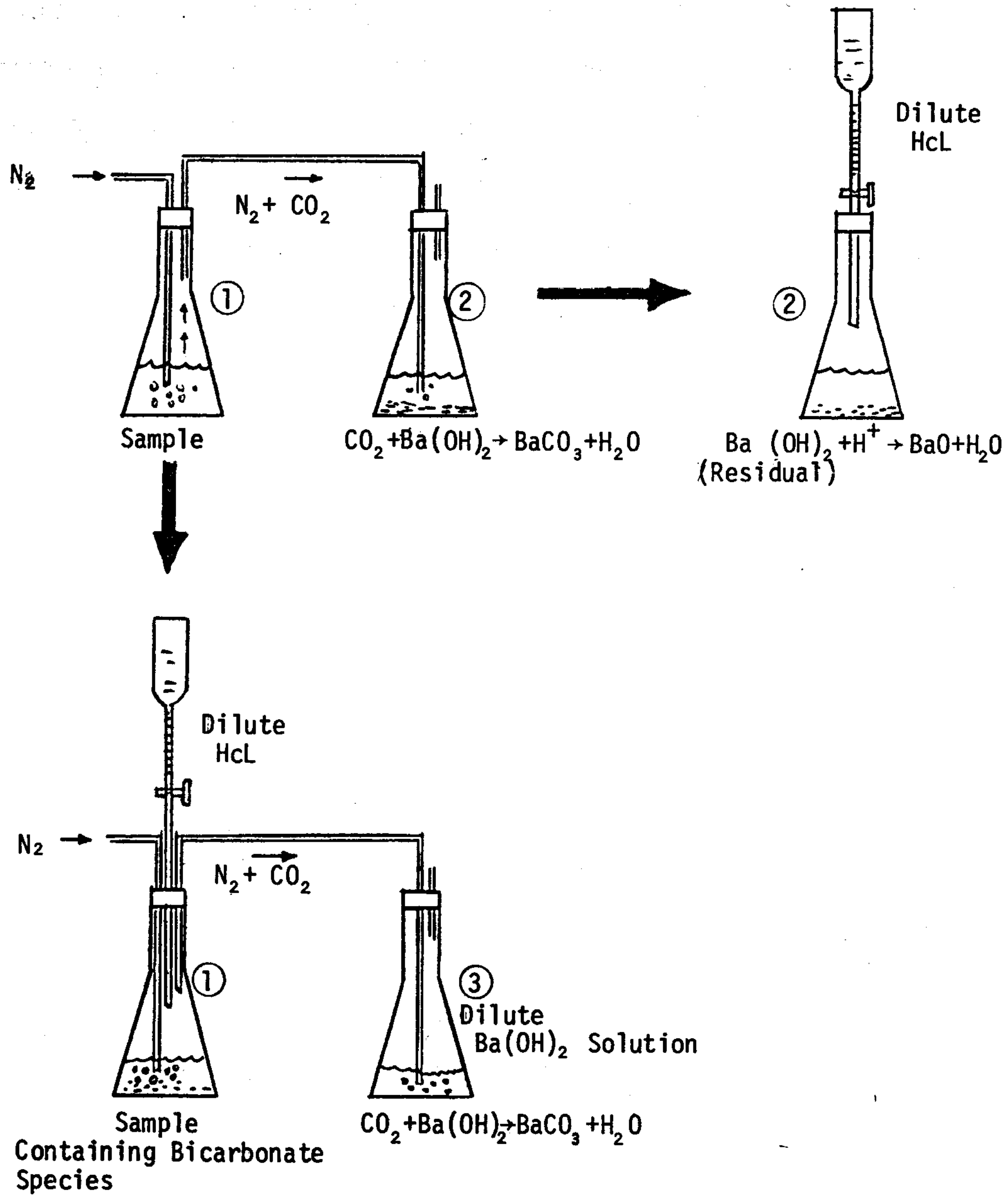
Many mechanical and operational problems with various components of the test loop were experienced during this test program. In addition, only nine sample bombs were available for taking gas or liquid high pressure samples. This was hardly sufficient to cover the scope of test work required in a continuous program. These coupled with the

FIGURE 7-4

LABORATORY SETUP FOR

BARIUM HYDROXIDE METHOD FOR DETERMINING

FREE CO_2



problem of developing an accurate and simple method for determining the total free CO_2 content of the brine within the very short time alloted for the tests, resulted in very limited and sketchy data.

One main objective of the field tests was to determine the amount of CO_2 removed from the brine by flashing in the new separator and relate this to the pH of the separator effluent brine and the temperature differential across the separator. Table 7-1 summarizes the operating data and laboratory results of these tests. An insufficient number of samples were taken of the effluent brine yielding reliable results to really meet the objective. However, the measured "free" CO_2 dissolved in the wellhead brine does appear to show a good correlation with field measured pH values. Figure 7-5 is a graph of the data. A straight line curve was fitted to the data using a linear regression method.

An interesting point to note is that in all cases shown on Table 7-1, "Method 2" or the Barium Hydroxide method for determining the free CO_2 in the residual brine yields higher values than the Back-Titration method. It is believed that these higher values are more accurate.

An attempt was made to compare the few data points obtained for the separator brine outlet with data obtained previously for the East Mesa 8-1 well. Figure 7-6 shows a curve of free CO_2 in the brine on flash-down versus pH values for a single stage flash. Data used to construct this curve was taken from a study of the well chemistry at East Mesa made by H.A.Papazian[9] (compliments of Lawrence Berkley Laboratory). The few points obtained from Table 7-1 and plotted all lie to the right side of the curve. However, as mentioned, the data was insufficient to draw any definite conclusions.

TABLE 7-1

SUMMARY OF OPERATING DATA AND FREE CO₂ CONTAINED IN BRINE FOR THE FLASH SEPARATOR EXPERIMENTS

				FLASH SEPARATOR INLET SAMPLES						FLASH SEPARATOR OUTLET SAMPLES				
						CO_2 CONTENT OF BRINE						CO_2 CONTENT OF BRINE		
DATE	WELL #	WELLHEAD PRESSURE (psig)	WELLHEAD BRINE TEMP. (F)	SAMPLE NO.	pH AT SAMPLE STATION	G.C. (ppm)	METHOD(1)* (ppm)	METHOD(2)** (ppm)	RTD*** (%)	SAMPLE NO.	pH AT SAMPLE STATION	G.C. (ppm)	METHOD(1) (ppm)	METHOD (2) (ppm)
11/14/78	8-1	99	308	1207	6.5	207	343	431	0%					
11/14/78	8-1	99	318	1199	6.2	6	122	130	24%	1200	6.9	N.D	47	61
11/15/78	8-1	98	318	1211	6.3	158	270	303	50%	1212	8.1	N.D	26	25
11/17/78	8-1	99	318	1215	6.3	64	157	226	51%	1216	7.2	85	RESIDUAL NOT RUN	166
11/3/78	6-2	135	330	1173	6.75	SAMPLES USED IN PRELIMINARY LABORATORY EXPERIMENTS				1174A 1174B 1174C	SAMPLES USED IN PRELIMINARY LABORATORY EXPERIMENTS			
11/8/78	6-2	136	329	1190	Not Recorded	59.7	RESIDUAL NOT RUN	469.7	60%	1189	Not Recorded	44	44	132
11/20/78	6-2	128	335	1220	6.15	255	RESIDUAL NOT RUN	255	55%	1221	7.93	SAMPLE NOT RUN		
11/22/78	6-2	128	335	1222	6.2	392	RESIDUAL NOT RUN	RESIDUAL	24%	1223	7.10	SAMPLE NOT RUN		

• Total Values for Method (1) = ppm determined by G.C. plus ppm Residual determined by Back-Titration

** Total Values for Method (2) = ppm determined by G.C. plus ppm Residual determined by Barium Hydroxide

*** RTD % = Percent of Venting where 100% equals a ΔT across separator of 18 F.

FIGURE 7-5

PH vs ppm FREE CO_2 IN WELLHEAD BRINE

(BRINE INLET SEPARATOR)

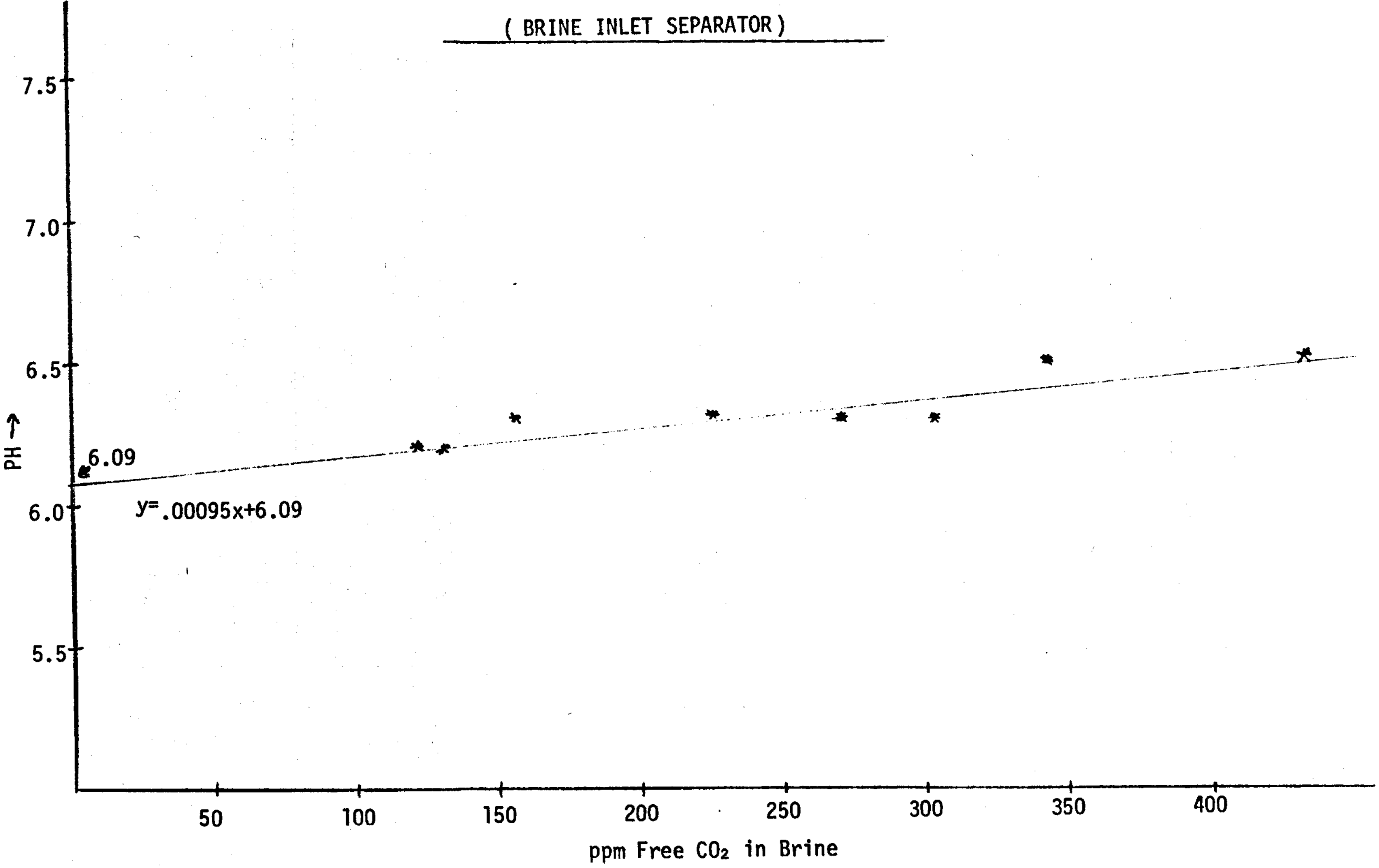
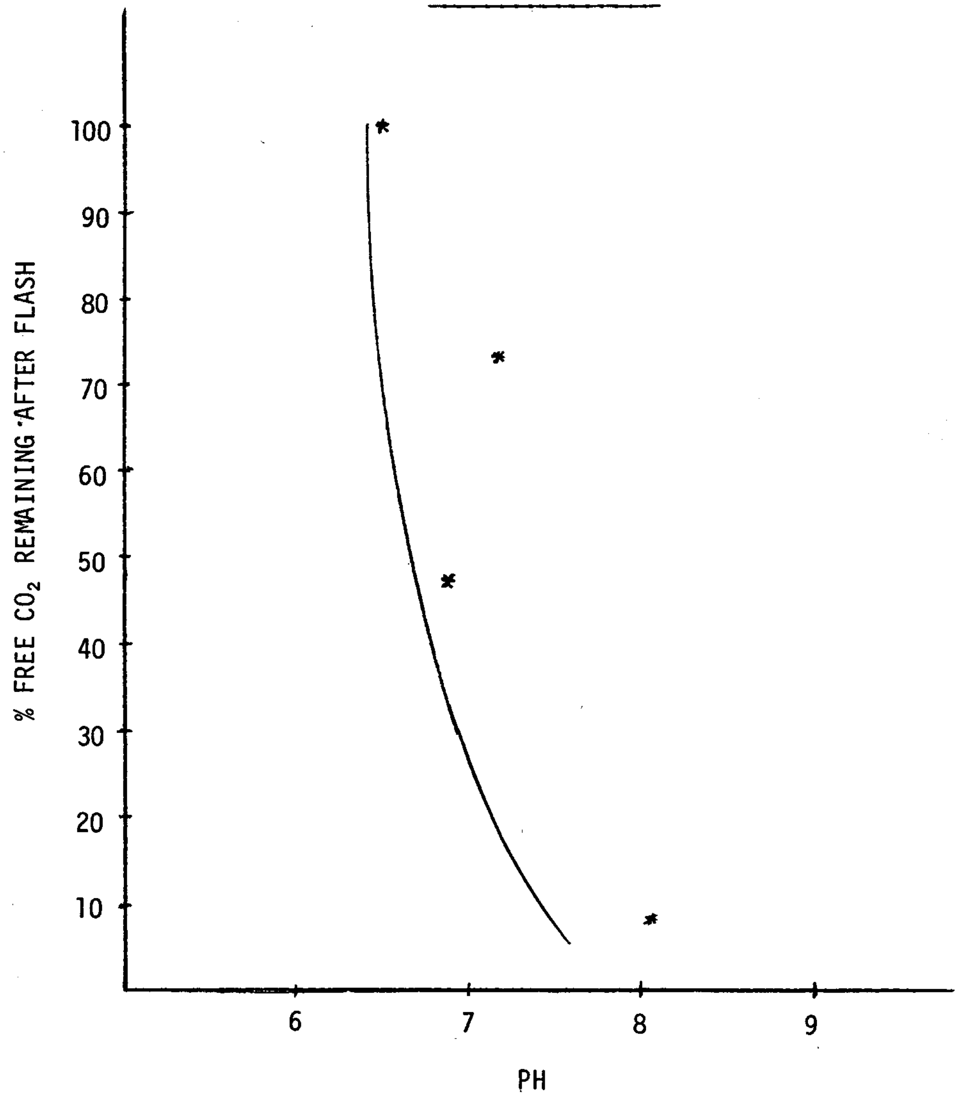


FIGURE 7-6

PERCENT FREE CO_2 IN BRINE
vs
PH AFTER FLASH



Because of the problems mentioned above, very little data necessary to determine what overall affect varying the free CO_2 has on loop performance was obtained. Table 7-2 is a summary of data taken at key sampling points, (see Figure 7-3). However, this was insufficient to construct CO_2 mass balances or indicate any definite trends. It is interesting to note the data indicates a substantial amount of CO_2 being introduced to the DCHX via the liquid isobutane stream (sample point 7). However, we know of no data discussing the solubility of CO_2 in hydrocarbons that could be used for comparison.

TABLE 7-2

SUMMARY OF OPERATING DATA AND FREE CO_2 CONTAINED IN BRINE FOR KEY POINTS ABOUT THE LOOP

TOTAL MEASURE CONCENTRATION (ppm) OF FREE CO_2 AT INDICATED SAMPLE POINT $\frac{\text{ppm by Back-Titration}}{\text{ppm by Ba(OH}_2\text{)}}$

TEST No.	DATE	WELL #	WELLHEAD PRESSURE (psig)	WELLHEAD BRINE TEMP (F)	% RTD	BRINE FLOW (lbs/hr)	C_4H_{10} FLOW (lbs/hr)	1	2	3 (by G.C.only)	4 (by G.C.only)	5	6	7 (by G.C.only)	8	9 (by G.C.only)
1	11/6/78	6-2	135	323	0	2668	2324	$\frac{117}{149}$				$\frac{58.7}{96.7}$	$\frac{21}{21}$			
2	11/8/78	6-2	136	329	60	3188	2370	$\frac{59.7}{469.7}$	$\frac{44}{132}$	64% CO_2	58% CO_2	$\frac{32}{82}$				
3	11/14/78	8-1	99	318	51	3569	2260						$\frac{35}{56}$	214		57
4	11/14/78	8-1	99	318	24	3458	2236	$\frac{122}{130}$	$\frac{47}{61}$				$\frac{106}{74}$ $\frac{30}{40}$	178		11
5	11/15/78	8-1	99	308	0	3456	2257	$\frac{343}{431}$			1000		$\frac{336}{286}$	924		
6	11/15/78	8-1	98	318	50	3550	2250	$\frac{270}{303}$	$\frac{26}{25}$							
7	11/16/78	8-1	100	319	51	3404	2252						$\frac{82}{20}$			
8	11/17/78	8-1	99	318	51	3409	2247	$\frac{157}{226}$	$\frac{85}{166}$				$\frac{35}{44}$			

SECTION 8

BRINE CARRYOVER EXPERIMENTS

SECTION 8

BRINE CARRYOVER EXPERIMENTS

8.0 INTRODUCTION

It has been demonstrated that electricity can be produced from a geothermal source using isobutane vapor generated by direct contact heat exchange to drive a turbo-generator in a closed Rankine power cycle.[1] A common Problem encountered during these experimental runs was the deposition of a silicious scale on the turbine nozzles and blading that resulted in a decrease in overall turbine efficiency. This scaling is caused by tiny droplets of liquid containing dissolved solids and silica being carried to the turbine by the isobutane vapor.

Liquid carryover in conventional boilers is caused by three basic mechanisms: Foaming, misting and priming. This should not be confused with the water content of the vapor present due to partial pressure considerations.

Foaming causes carryover when a stable froth is formed at the free surface of the brine and becomes entrained in the vapors leaving the boiler. This has not been evident in the DCHX test unit, however, since the vapor space is difficult to observe it should not be discounted.

It is theorized that misting may occur when tiny particles of brine are thrown off by the sudden bursting of working fluid bubbles at the water surface.

Priming includes all "carryover" situations caused by significant slugs of brine being lifted upward by the boiling working fluid. This mechanism becomes critical when the boiler is operated with a high free surface level minimizing the vapor release space.

As part of a previous DCHX-isobutane test program, attempts were made at quantifying and qualifying the water or brine present in the vapor leaving the DCHX boiler and correlating the measured rates with operating variables such as isobutane flow rate, brine flow rate, boiler level and superheat. The results of those tests were as follows: (1) no definite correlations between boiler operating level and the amount or salinity of the water content of the isobutane vapor stream were apparent;(2) no correlation between the partial pressure or degrees superheat of the isobutane and weight fraction water content of the vapor phase was found; and (3) the water content of the boiler vapors was greater then theoretically predicted by ideal gas low partial pressure methods. Data from these tests are given in Appendix (C) for reference.

The scope of this task was: (1) to determine the water vapor fraction of the vapor generated by the DCHX using isopentane; (2) compare this with previous isobutane data to determine if there is a constant trend; and (3) formulate an empirical relationship expressing water vapor percentages as a function of vapor temperature or superheat.

8.1 EQUIPMENT AND PROCEDURES

Carryover tests were conducted for both high temperature and low temperature isopentane cycles. Basically, the test loop used to conduct these carryover experiments was the same as previously used for the isobutane loop carryover tests; ie, vapors from the direct contact heat exchanger to the knockout drum to the dummy nozzle to the condenser, liquid isopentane from the condenser to the hotwell and back to the DCHX (see P & I diagrams-Figures 3-2 and 3-3).

For all "low" temperature cycle isopentane tests, brine was introduced above the apparent free surface level through the lower tangetial inlet nozzle. Figure 3-5 shows the installation of this nozzle as well as the level control mechanisms. The tangential inlet configuration allows for concurrent type heat exchange as well as vapor-liquid separation thus increasing overall boiler efficiency. Both inlet nozzles were used during the "high" temperature cycle tests.

To simulate turbine pressure drop a "dummy" turbine nozzle assembly was installed in the vapor piping between the knockout drum and condenser. The same nozzle used for the isobutane tests was employed for a high temperature isopentane cycle. A new nozzle sized for a downstream pressure of 16-20 psi was fabricated from type 316 stainless steel and installed replacing the high pressure nozzle for the "low" temperature isopentane cycles.

Float type water traps, installed on the bottom drain lines of the knockout drum and hotwell provided continuous blowdown of condensed water. The water was collected and measured during the testing as an indication of the total water content of the isopentane vapor stream. Conductivity of the water was also measured for comparison with the isobutane runs.

Twelve "high" temperature cycle isopentane tests and eighteen "low" temperature tests were conducted. The time of each test was set at an arbitrary 2 hours after initially allowing the DCHX - isobutane loop to stabilize for 1 to 2 hours. The general procedure used was as follows:

a) Start the DCHX at a controlled isobutane flow rate, brine flow rate, and boilers level. Allow system to equilibrate.

b) Monitor all pressures, temperatures, flow rates, levels and conductivity at 10 minute intervals.

c) Collect all blowdown effluent from the knockout drum and hot well water traps. Measure and record total flow rates from each at end of test.

d) Change level in boiler holding brine and isobutane flow rate constant. Allow system to equilibrate and repeat b) and c) above.

e) Change flow rates and repeat sequence above.

8.2 RESULTS AND DISCUSSION

Tables 8-1 through 8-5 summarize the "carryover" data collected during the isopentane runs; 8-1 through 8-3 are for "high" temperature cycles and 8-4 and 8-5 are for "low" temperature cycles. The values on the tables represent the arithmetic averages of data collected for each test run.

Much effort and time has been spent trying to relate boiler operating variables to the measured amount and salinity of the water content of the boiler vapors. As previously experienced with isobutane, no definite correlations between boiler operating level, mass flow rates, vapor temperature, vapor pressure or degrees superheat and weight fraction water content are apparent.

A standard chemical engineering practice used in low pressure distillation assumes that a mixture of immiscible liquids will vaporize in direct molar ratio to their partial pressures at the common saturation temperature. This is based on the ideal gas law. However, the generalized gas laws predict that the liquids will vaporize in inverse molar ratio to their saturated molar volumes at the common saturation temperature.

A problem lies in determining this common boiling or saturation temperature. For equilibruim mixtures, the common temperature can be found by trial and error as the temperature at which the vapor pressures of the immiscible components add up to the total measured pressure of the system. Another method is based on a heat balance about the preheater section of the DCHX. The following equation can be used.

$$(\dot{M}_{brine})\ (C_p^b)\ (T_c - T_{bo}) = (\dot{M}_{wf})\ (C_p^{wf})\ (T_c - T_{wfi})$$

Where:

$\dot{M}_{brine}$ = Mass flow of brine (lbs/time)

Cp^b = Heat capacity of brine (Btu/lb-F)

TABLE 8-1

CARRYOVER DATA SUMMARY-FIELD MEASURED VALUES

HIGH TEMPERATURE ISOPENTANE CYCLES

DATE	TEST#	BRINE FLOW (gpm)	BRINE FLOW (lbs/hr)	ISOPENTANE FLOW (gpm)	ISOPENTANE FLOW (lbs/hr)	BOILER LEVEL[1] (INCHES)	KNOCKOUT DRUM FLOW (lbs/hr)	KNOCKOUT DRUM CONDUCTIVITY (Umhos)	HOTWELL FLOW (lbs/hr)	HOTWELL CONDUCTIVITY (Umhos)
8/25/78	1*	5.87	2723	4.38	1361	+ 6	5.6	200	37.0	325
9/ 1/78	2*	4.47	2072	3.63	1112	0	4.1	114	5.9	350
9/ 1/78	3*	4.54	2113	3.59	1103	- 6	3.8	386	7.8	350
9/ 1/78	4*	4.62	2143	3.59	1102	+ 6	-[2]	100	-[2]	300
9/ 1/78	5*	5.19	2413	3.51	1076	+ 6	5.9	40	23.8	250
8/26/78	6*	4.16	1928	3.49	1086	- 1	3.8	250	30.7	380
8/31/78	7*	5.21	2404	3.42	1051	- 6	4.2	100	58.2	285
8/25/78	8*	5.84	2683	4.35	1350	+ 5	3.7	211	33.7	250
9/ 1/78	9**	4.22	1954	3.65	1129	+ 1	2.7	70		260
8/31/78	10**	4.56	2089	3.54	1087	- 5	3.0	200	43.0	280
9/ 1/78	11**	4.68	2159	3.64	1115	+ 6	3.2	230	11.1	250
9/ 1/78	12**	5.35	2487	3.61	1108	+ 6	3.5	100	12.4	300

* Brine was introduced to boiler through new brine nozzle.

** Brine was introduced to boiler through old brine nozzle.

[1] Boiler level reference is at new brine nozzle entrance, (-) indicates level below and (+) indicates level above; see Figure

[2] Incomplete due to plug up in water trap

TABLE 8-2

CARRYOVER DATA SUMMARY-CALCULATED VALUES

HIGH TEMPERATURE ISOPENTANE CYCLES

TEST NO.	OBSERVED BOILER VAPOR TEMP. (°F)	OBSERVED BOILER VAPOR PRESS. (psia)	CALCULATED BOILING TEMP. (°F)	IC_5 PRESS (psia)	H_2O PRESS (psia)	IC_5 SUPERHEAT (°F)	$\frac{\text{lbs } H_2O \text{ (HW)}}{\text{lb } IC_5}$	$\frac{\text{lbs } H_2O \text{ (KO)}}{\text{lb } IC_5}$	$\frac{\text{Lbs } H_2O \text{ (Total)}}{\text{lb } IC_5}$	THEORETICAL[1] $\frac{\text{Lbs } H_2O}{\text{ln } IC_5}$
1	240	120.8	213	105.8	15.0	27	.0272	.0041	.0313	.0296
2	226	120.4	213	105.5	14.9	13	.0053	.0036	.0089	.0296
3	233	126.5	217	110.4	16.1	16	.0071	.0034	.0105	.0305
4	236	118.8	212	104.2	14.6	24	-	-	-	.0294
5	250	122.8	214	107.4	15.4	36	.0221	.0055	.0276	.0298
6	232	117.1	211	102.8	14.3	21	.0283	.0035	.0318	.0292
7	253	133.4	221	115.9	17.5	32	.0554	.0040	.0594	.0314
8	243	123.7	215	108.1	15.6	28	.0250	.0027	.0277	.0301
9	228	126.4	217	110.3	16.1	11	Not Collected	.0024	-	.0305
10	230	128.9	218	112.3	16.6	12	.0396	.0028	.0424	.0307
11	231	124.0	215	108.3	15.7	16	.0099	.0029	.0128	.0301
12	247	123.3	215	107.8	15.5	32	.0112	.0032	.0144	.0301

(1) Based on ideal gas law

TABLE 8-3

CARRYOVER DATA SUMMARY-ANALYSIS OF BLOWDOWN SAMPLES
HIGH TEMPERATURE ISOPENTANE CYCLES

TEST NO.	SAMPLE LOCATION	LAB SAMPLE NO.	LAB CONDUCTED ANALYSIS RESULTS			
			CONDUCTIVITY (ppm)	TOTAL SOLIDS (ppm)	TDS (ppm)	pH
1	Knockout Drum Blowdown	1059	310	Not run	152	6.62
	Hotwell Blowdown	1060	382	Not run	108	5.52
	K.O. Blowdown	1081	164	64	44	6.23
2	HW Blowdown	1082	386	96	44	6.00
	K.O.	1075	477	288	264	6.43
3	H.W.	1078	353	104	76	5.87
4	No Samples Taken		-	-	-	-
5	K.O.	1086	110	48	36	7.05
	H.W.	1087	310	220	180	6.20
6	K.O.	1066	335	Not run	112	5.54
	H.W.	1067	338	Not run	58	5.91
7	K.O.	1076	153	56	64	6.21
	H.W.	1079	322	136	132	5.73
8	K.O.	1064	255	Not run	96	6.5
	H.W.	1065	305	Not run	116	6.3
9	K.O.	1083	143	56	32	6.00
	H.W.	1084	342	72	40	5.85
10	K.O.	1077	234	112	104	6.36
	H.W.	1080	309	56	60	5.75
11	K.O.	1088	157	96	56	7.44
	H.W.	1089	296	252	40	6.40
12	K.O.	1080	115	40	44	6.64
	H.W.	1091	310	256	172	6.61

TABLE 8-4

CARRYOVER DATA SUMMARY-DCHX FLOWS AND OPERATION

LOW TEMPERATURE ISOPENTANE CYCLES

TEST No.	DATE	(BRINE(Mb) FLOW (lbs/min)	ISOPEN-TANE(M_I) FLOW lbs/min.	$\frac{M_I}{M_b}$	DCHX OPERATING PARAMETERS LEVEL (INCHES)	VAPOR TEMP. (F)	PRESS (PSIA)	KNOCKOUT DRUM BLOWDOWN (lbs/min)	HOTWELL BLOWDOWN (lbs/min)	WATER VAPOR LOADING$\left(\frac{\text{lbs } H_2O}{\text{lbs ISO}}\right)$ TO K O DRUM	TO HOTWELL	TOTAL
1	1/18/79	52.29	18.49	.353	-1.3	140	41.9	2.7	2.6	.146	.141	.287
2	1/17/79	64.09	18.47	.288	-1	142	36.0	1.4	0.9	.076	.049	.125
3	1/17/79	60.69	18.46	.304	-1.3	154	40.0	2.7	1.8	.146	.098	.244
4	1/23/79	84.02	28.97	.345	-2.1	147	42.0	0.6	0.9	.201	.031	.052
5	1/22/79	84.02	28.97	.345	-2.8	149	41.0	0.3	1.3	.010	.045	.055
6	1/22/79	84.73	28.99	.342	-2.7	156	43.0	0.6	1.1	.021	.048	.059
7	1/25/79	135.70	39.20	.288	-4.5	145	45.0	5.8	1.3	.148	.033	.181
8	1/24/79	116.70	38.80	.331	-6	151	45.0	4.1	1.8	.106	.046	.152
9	1/26/79	122.01	38.20	.313	-5	158	45.0	7.4	1.3	.194	.034	.228
10	1/29/79	70.80	21.60	.305	-4	144	39.8	0.1	1.2	.005	.056	.061
11	2/1/79	55.07	14.54	.264	-4	152	41.0	0.07	0.1	.005	.007	.012
12	2/1/79	56.00	17.38	.310	-3	142	40.0	0.05	0.2	.003	.012	.015
13	2/2/79	65.18	20.02	.307	-3.5	151	40.0	0.03	0.5	.001	.025	.026
14	2/2/79	65.34	23.13	.354	-5.0	141	40.0	0.02	0.5	.001	.022	.023
15	2/5/79	65.34	16.77	.257	-2.0	166	40.0	0.07	0.6	.004	.036	.040

TABLE 8-4 (cont.)

CARRYOVER DATA SUMMARY-DCHX FLOWS AND OPERATION

LOW TEMPERATURE ISOPENTANE CYCLES

					DCHX OPERATING PARAMETERS					WATER VAPOR LOADING $\left(\frac{\text{lbs } H_2O}{\text{lbs ISO}}\right)$		
TEST No.	DATE	(BRINE(Mb) FLOW (lbs/min)	ISOPENTANE(M_I) FLOW lbs/min.	M_I/M_b	LEVEL (INCHES)	VAPOR TEMP. (F)	PRESS (PSIA)	KNOCKOUT DRUM BLOWDOWN (lbs/min)	HOTWELL BLOWDOWN (lbs/min)	TO K O DRUM	TO HOTWELL	TOTAL
16	2/5/79	65.34	18.31	.280	-3.0	165	40.0	0.07	0.6	.004	.033	.037
17	2/6/79	74.64	24.70	.331	-3.0	145	41.0	0.10	0.6	.004	.024	.028
18	2/6/79	74.64	20.76	.278	-4.0	160	40.0	7.90	5.0	.381	.241	.622

TABLE 8-5

CARRYOVER DATA SUMMARY-CALCULATED VALUES

LOW TEMPERATURE ISOPENTANE CYCLES

TEST No.	OBSERVED BOILER VAPOR TEMP Tb_0 (F)	OBSERVED BOILER PRESS (psia)	CALCULATED BOILING[1] TEMPERATURE Tb_1 (F)	Tb_2 (F)	ISOPENTANE (Tb_0-Tb_1) (F)	SUPERHEAT (Tb_0-Tb_2) (F)	THEORETICAL WATER VAPOR LOADING $\left(\frac{lbsH_2O}{lbsC_5H_{12}}\right)$ @ Tb_1	@ Tb_2	@ Tb_0
1	140	41.9	139	125	1	15	.0162	.0142	.0160
2	142	36.0	130	123	12	19	.0149	.0139	.0167
3	154	40.0	135	124.5	19	30	.0156	.0142	.0185
4	147	42.0	139	125	8	22	.0162	.0142	.0174
5	149	41.0	137	124	12	25	.0159	.0141	.0178
6	156	43.0	141	126	15	30	.0165	.0144	.0189
7	145	45.0	143	126	2	19	.0168	.0144	.0171
8	151	45.0	143	131	8	20	.0168	.0151	.0181
9	158	45.0	143	129	15	29	.0168	.0148	.0192
10	144	39.8	135	119	9	25	.0156	.0134	.0170
11	152	41.0	137	124	15	28	.0159	.0141	.0182
12	142	40.0	135	120	7	22	.0156	.0136	.0167
13	151	40.0	135	122	16	29	.0156	.0138	.0181
14	142	40.0	135	122	7	20	.0156	.0138	.0167
15	166	40.0	135	140	31	26	.0156	.0162	.0206

[1] Tb_1 calculated using ideal gas law and $P_{H_2O}+P_{IC_5} = P_{Total}$
Tb_2 calculated from liquid-liquid heat balance

[2] At saturated (boiling) lbs H_2O/lb $IC_5 = Vg^s(IC_5)/Vg^s(H_2O)$ where Vg^s = specific volumes

TABLE 8-5 (cont)

CARRYOVER DATA SUMMARY-CALCULATED VALUES

LOW TEMPERATURE ISOPENTANE CYCLES

TEST No.	OBSERVED BOILER VAPOR TEMP Tb_0 (F)	OBSERVED BOILER PRESS (psia)	CALCULATED BOILING TEMPERATURE[1]		ISOPENTANE SUPERHEAT		THEORETICAL WATER VAPOR LOADING $\left(\frac{lbs H_2O}{lbs C_5H_{12}}\right)$		
			Tb_1 (F)	Tb_2 (F)	$(Tb_0 - Tb_1)$ (F)	$(Tb_0 - Tb_2)$ (F)	@ Tb_1	@ Tb_2	@ Tb_0
16	165	40.0	135	143	30	22	.0156	.0168	.0204
17	145	41.0	137	124	8	21	.0159	.0141	.0171
18	160	40.0	135	126	25	34	.056	.0144	.0195

---

[1] Tb_1 calculated using ideal gas law and $P_{H_2O} + P_{IC_5} = P_{Total}$

[2] Tb_2 calculated from liquid-liquid heat balance

At saturation (boiling) lbs H_2O/lb IC_5 - $Vg^S(IC_5)/Vg^S(H_2O)$ where Vg^S = specific volumes

$\overset{\circ}{M}wf$ = Mass flow of working fluid (lbs/time)

C_p^{wf} = Heat capacity of fluid (Btu/lb-F)

Tbo = Brine outlet temperature (F)

Twfi = Working fluid inlet temperature (F)

Tc = Common temperature

The theoretical vapor loading $\left(\frac{\text{lbs } H_2O}{\text{lbs Isopentane}}\right)$ for each test was calculated using methods outlined above. These are compared with the measured vapor loading in Tables 8-2 and 8-5. As can be seen, there is no general relationship between measured and theoretical values.

Another observation of the data included in Table 8-3, is the salinity content of the hotwell blowdown. Theoretically, the blowdown should only be condensed water vapor having a very low or negligible salinity. The measured salinity can be adequately explained by acknowledging the premise that ammonium bicarbonate may be formed on cooling in the condenser. The fact that ammonia is present in the vapor has been verified by Bechtel[10]

SECTION 9

ISOBUTANE RECOVERY SYSTEMS

SECTION 9

ISOBUTANE RECOVERY SYSTEMS

9.0 INTRODUCTION

In direct contact systems, the cost of the isobutane dissolved in the rejected brine is considered an additional operating expense over a system using surface type heat exchangers. The amounts of isobutane lost and the annual costs corresponding to anticipated rates of loss for a 50 MW_e (net) plant are tabulated in Table 9-1. As indicated on the table, the measured amount of isobutane in the brine effluent from the direct contact heat exchanger averages 193 ppm. For a 50 MW_e power plant without an isobutane recovery system flowing brine at 9.83 million pounds per hour, this amounts to an isobutane loss rate of 14.96 million pounds per year or an annual cost of $2.24 million. Thus, an isobutane recovery system is imperative.

Based on this need, DSS Engineers, Inc., has designed and operated a degassing or stripping column at the East Mesa Test Facility. This has successfully demonstrated good efficiency (average volatilization efficiency 98%) in removing the dissolved isobutane from the brine. The measured isobutane concentration of the reject brine leaving the degassing column averages 10.4 ppm. Should this type of column be utilized at the 50 MW_e power generating facility mentioned above, the isobutane losses could be reduced to 806,000 lbs/year for an estimated overall annual savings of $2.12 million

TABLE 9-1
ANNUAL OPERATING COSTS
FOR ISOBUTANE LOST VIA REJECT BRINE

Basis:

Plant Size:	50 MW_e
Brine Flow:	9,830,000 lbs/HR
Load Factor:	90%
(1) Isobutane Cost:	15¢/lb commercial grade (estimated 1979 bulk-delivered)

ISOBUTANE CONCENTRATION IN:

Rejected Brine (ppm)	Isobutane Consumption (lbs/YR)	Annual Cost ($)	Cost (MILS/KWH)
5	387,500	58,125	.15
10	775,000	116,250	.29
10.4(2)	806,000	120,900	.31
20	1,550,000	232,500	.59
193(3)	14,960,000	2,244,000	5.7

(1) Supplier: Philips Chemical Company, Texas

(2) Estimated degassing column outlet concentration (Reference 1)

(3) Estimated degassing column inlet concentration (Reference 1)

However, degassing alone does not solve this isobutane consumption problem. The degassing column also liberates carbon dioxide and some water vapor from the brine, and as a result, the vapor leaving the column is a mixture of isobutane, carbon dioxide and water vapor. The problem becomes how to recover the isobutane from the gas mixture. This is the purpose of this section of the report.

9.1 Theory

Many elaborate and intriguing methods of gas-gas and gas-liquid separation may be adapted to this system. However, it is beyond the scope of this study to investigage each alternative. Therefore, the proposed recovery system will be based on state-of-the-art compression and condensation operations.

Table 9-2 shows the composition of the vapors leaving a degassing column being operated at essentially one atmosphere and 137 F at the East Mesa Test Site[1]. The measured mole fractions of the gas averaged: 37.1% isobutane, 38.4% CO_2 and 24.5 % other gas which is assumed to be water vapor. A family of curves was constructed from this data using ideal gas law methods that show the theoretical relationship of pressure, temperature and isobutane mole fraction of the gas phase at saturation. This is shown as Figure 9-1 assuming these would be the conditions in an isobutane recovery condenser. Based on these curves, the recovery of 90% (arbitrary) of the isobutane in the mixture (reducing the mole fraction of the gas phase isobutane to .088) by polytropic compression and condensation would require a condenser pressure of over 850 psi. This would require large compressors and

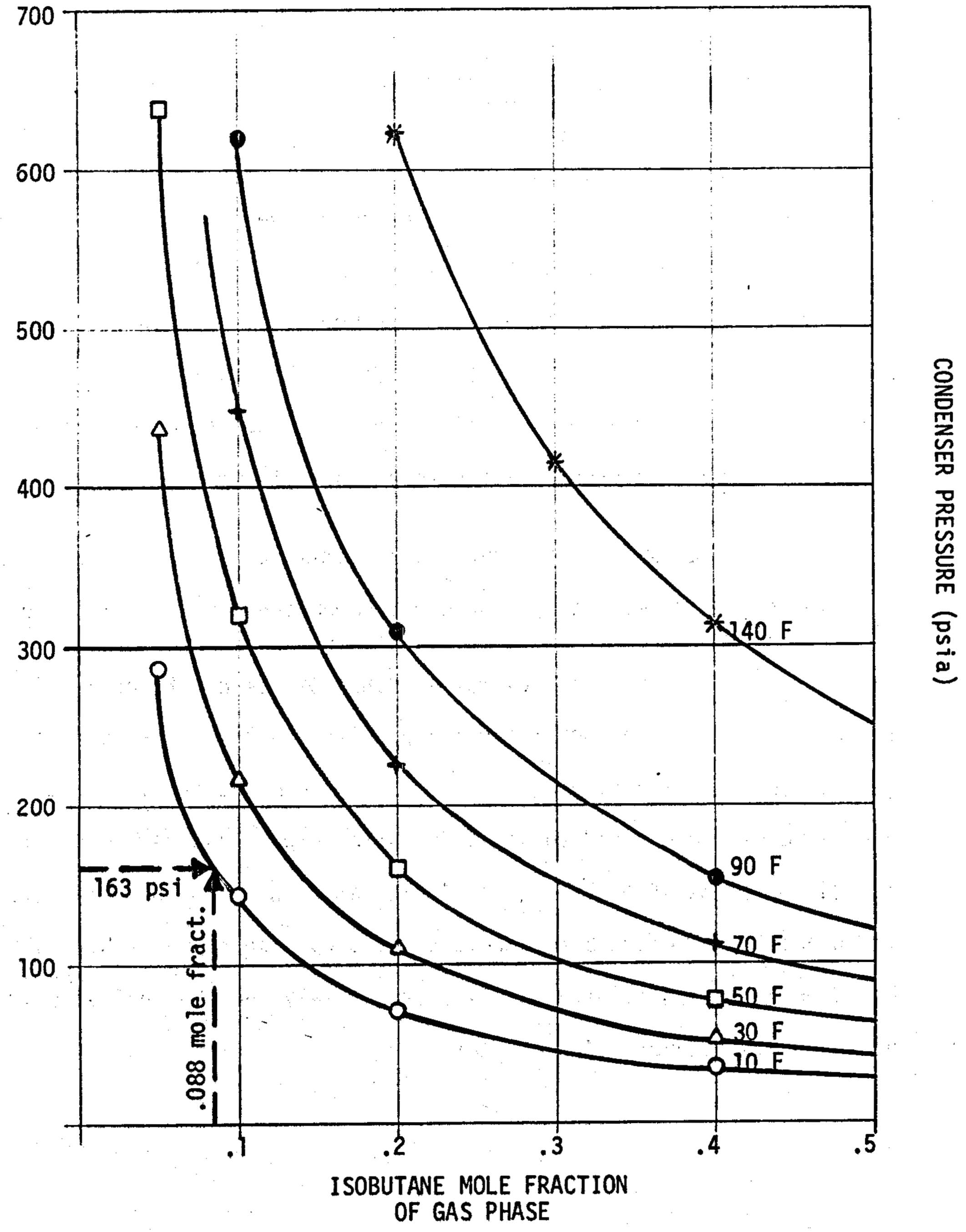
TABLE 9-2

COMPOSITION OF DEGASSING COLUMN VAPORS IN MOLE PERCENT(1)

Test No.	C_4H_{10} %	CO_2 (%)	Other (Assumed H_2O) %
1	43.85	33.68	22.47
2	35.38	35.84	28.78
3	39.34	34.36	26.30
4	33.34	43.75	22.91
5	33.75	44.04	22.21
Averages	37.1	38.4	24.5

FIGURE 9-1

CONDENSER PRESSURE AND TEMPERATURE REQUIRED TO CONDENSE ISOBUTANE FROM A MIXTURE OF ISOBUTANE AND CO_2 (DRY BASIS)



high pressure vessels and piping and would be financially undesirable.

By utilizing compression with refrigeration a lower pressure condenser is possible and the energy requirement of the recovery system will be lower.

9.2 Recovery System for a 50 MWe Facility

9.2.1 Heat and Mass Flow Considerations

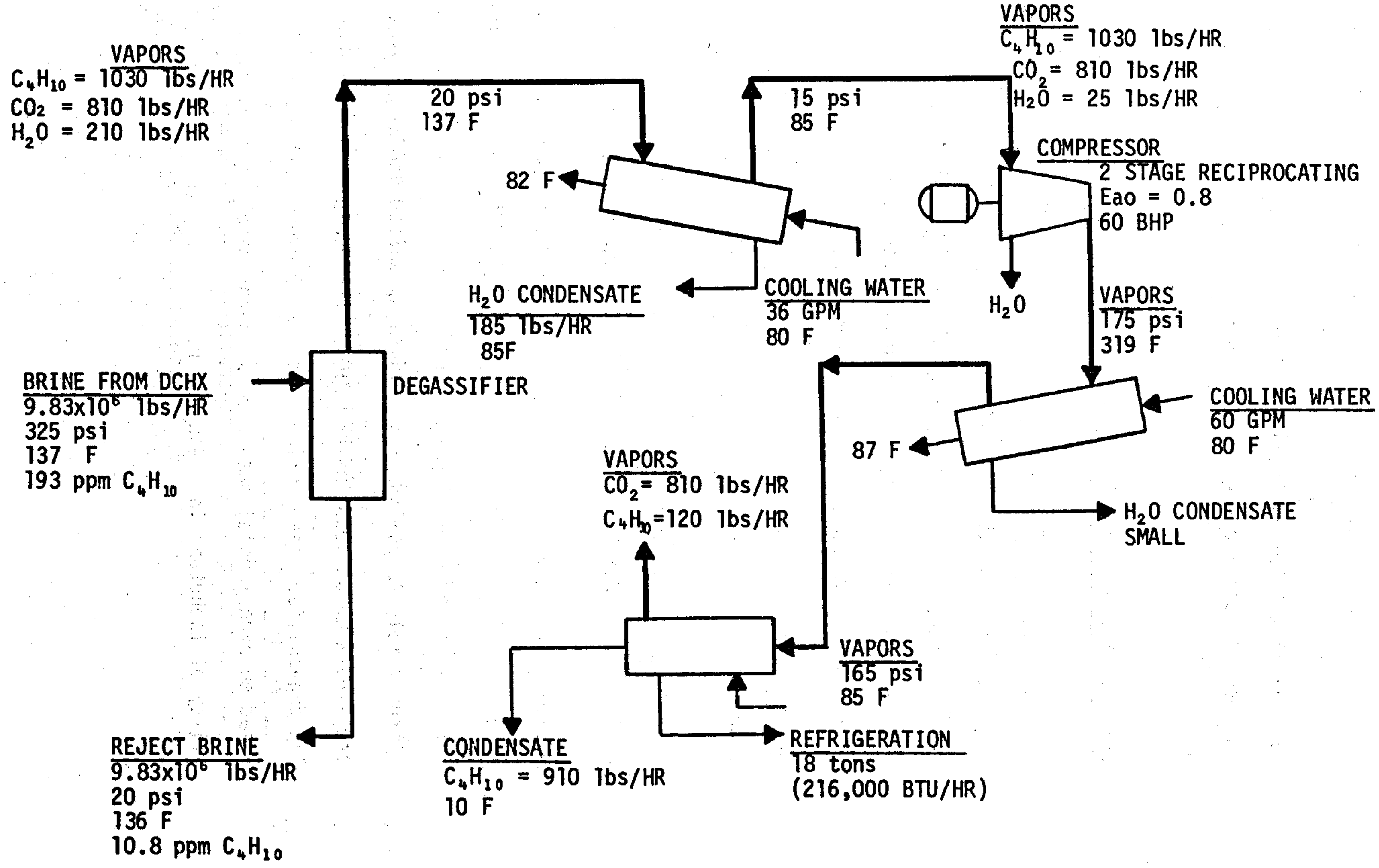
The basis for the systems studied involves recovery of 90% ($^{wt}/wt$) of the isobutane from the degassing column vapor effluent by compression and cooling. At the same time, it is economically advantageous to minimize the pressure in the final isobutane condenser. Based on the curves shown in Figure 9-1, the final condenser operating conditions required for 90% recovery were chosen at 163 psi and 10 F.

Accordingly, heat and mass balances were estimated for a recovery system based on the East Mesa data. These are shown schematically in Figure 9-2.

Referring to Figure 9-2, approximately 9.83×10^6 lbs/hr of brine effluent at 325 psi, 137 F and containing 193 ppm isobutane is introduced into a vertical packed degassifying column. Here the brine is flashed to approximately 20 psi and volatile isobutane, water and carbon dioxide are liberated at 1030 lbs/hr, 210 lbs/hr and 810 lbs/hr respectively. The degassifier reject brine contains 10.8 ppm isobutane. This is based on experimental data and is approximately the theoretical solubility of isobutane in fresh water.

FIGURE 9-2

ISOBUTANE RECOVERY SYSTEM FOR 50 MW_e GENERATING FACILITY
HEAT AND MASS FLOW CONSIDERATIONS



The vapors from the degassifier are then passed through a surface type heat exchanger where they are cooled to approximately 85 F. This requires 36 gpm cooling water at 80 F. Based on the theoretical saturated equilibrium data, approximately 185 lbs/hr water will condense in the heat exchanger.

The precooled vapors are then compressed to 175 psi. Because of the relatively low volumetric flow rate of gas (300 SCFM) a two-stage reciprocating compressor will probably be required. Based on an overall adiabatic efficiency (E_a0) of 0.8 and maximum compressor ratio of 3.5:1 per stage, the power requirements of the compressor are 60 bhp. The final compressor discharge temperature calculates to be 319 F.

The gases leaving the compressor are then precooled to approximately 85 F in a second surface type heat exchanger. This requires 60 gpm water at 80 F. A small amount of water will be condensed but the isobutane should remain in the gaseous state at this temperature and pressure.

The vapors are then cooled to 10 F in the final condenser and it is expected that 910 lbs/hr of isobutane will be condensed. At this temperature, the vapors leaving the condenser should contain 12.9% isobutane and 87.1% carbon dioxide. Approximately 216000 Btu/hr of heat must be removed in the condenser requiring 18 tons of refrigeration.

9.2.2 Refrigeration Considerations

The two general types of refrigeration cycles in common use are: 1) compression refrigeration and, 2) absorption refrigeration.

In compression refrigeration, shown in Figure 9-3, a refrigerant fluid is employed to absorb heat at low pressure and temperature by evaporation. Mechanical energy is then used to compress the refrigerant vapors to a higher pressure and temperature where the heat is rejected in a condenser to a heat sink. Free expansion to the lower pressure and temperature completes the cycle.

In absorption refrigeration, shown in Figure 9-4, no mechanical energy is required except in pumping a liquid absorbent and this is small compared to the energy required for mechanical compression in a compression refrigeration cycle. In this system heat is absorbed at a low temperature and pressure and rejected at an intermediate temperature and high pressure. The refrigerant is compressed from the low pressure (heat absorption) to the high pressure (heat rejection) by intermediate absorption at a low-pressure level. The refrigerant is then released from the absorbent at a high pressure through the addition of heat from a high temperature source.

Absorption refrigeration has the advantage of minimizing moving mechanical parts and may effectively utilize low-pressure steam as the source of heat. Unfortunately, a decision as to which type of refrigeration system to include in the isobutane recovery scheme can not be made based on these or the energy requirements alone. The capital costs of the equipment must also be considered.

9.2.3 Capital Investment Analysis-Absorption Refrigeration vs Compression Refrigeration

Before the isobutane recovery system design can be further developed and optimized, it must be shown that recovery of the isobutane will be

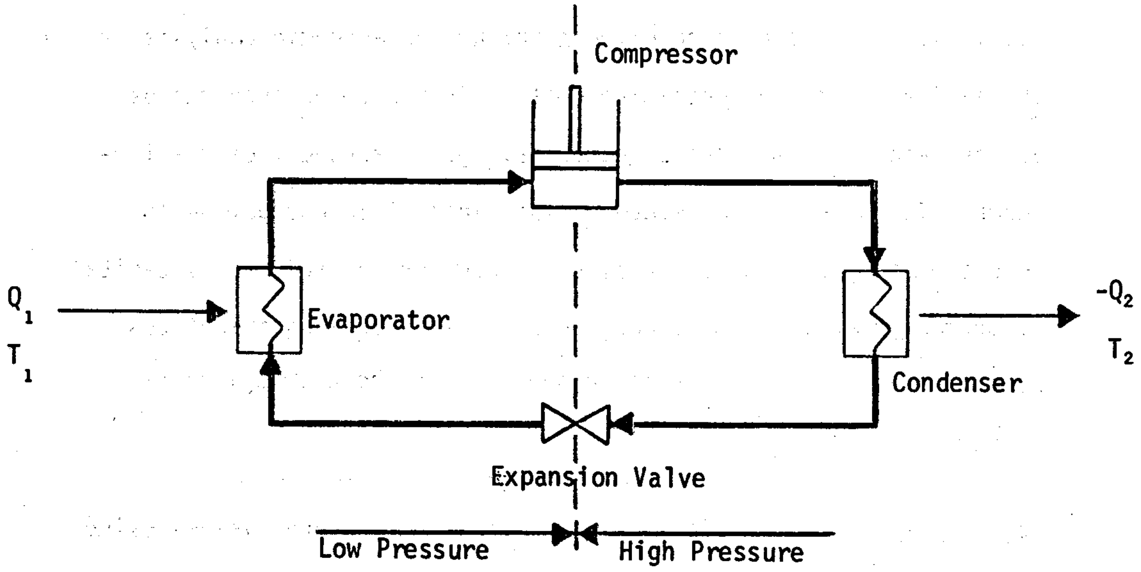


FIGURE 9-3 - Mechanical Compression Refrigeration Cycle (single stage)

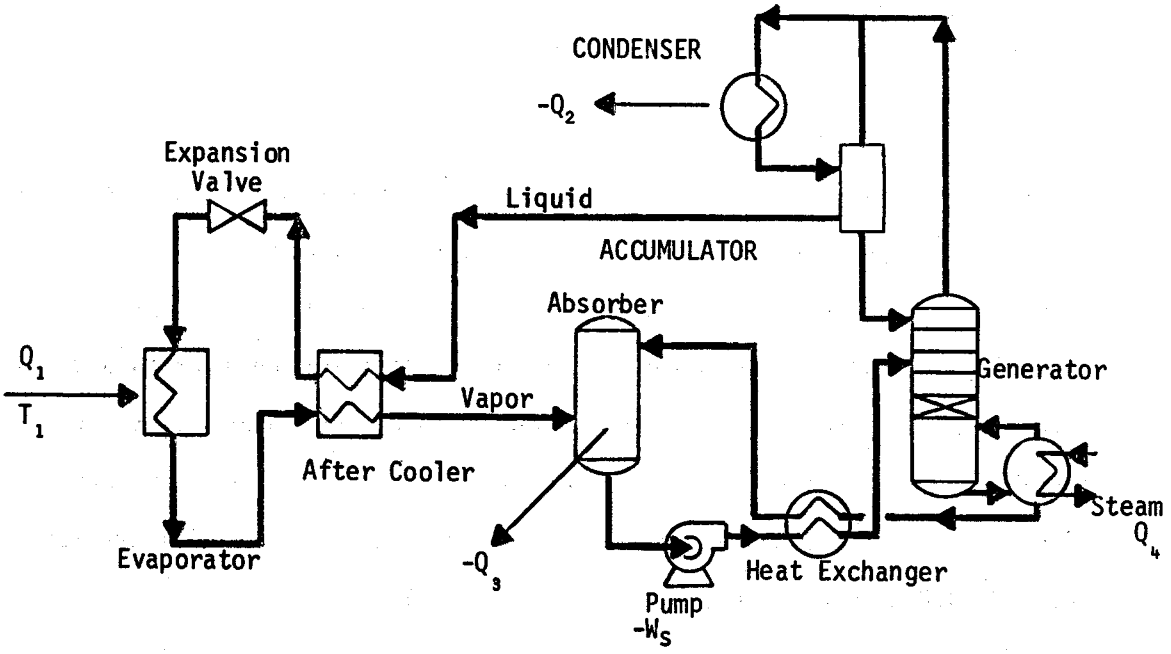


FIGURE 9-4 - ABSORPTION REFRIGERATION

profitable. This can be determined through an economic analysis of capital investment and operating costs. Then a comparison can be made between these and the expected savings on recovery of the isobutane. It is beyond the scope of this initial investigation to present a detailed economic analysis. However, a preliminary capital and operating cost estimate was prepared using base equipment costs and percentage factors considered average for the chemical process industry.

The major capital cost elements for isobutane recovery systems using absorption and compression refrigeration are compared on Table 9-3. Basic modular concept method and factors presented by Guntrie(11) were used to determine these costs. In summary, the capital costs of a complete installed isobutane recovery system (1979 pricing) using absorption refrigeration totals \$240,912 and one using mechanical (compression) refrigeration would be \$277,570.

The major production cost items and values for both systems are listed on Table 9-4. In general these operating costs consist of feedstocks (make up refrigerant); utilities (electricity, steam, etc.); miscellaneous process materials and supplies; operating and maintenance labor, including payroll burdens and plant overhead; and maintenance parts and supplies. To these costs are added fixed costs of insurance, taxes and depreciation to arrive at total operating costs. In summary, the operating costs of an isobutane recovery system using absorption refrigeration total \$154,888 annually and one using mechanical refrigeration is \$173,705.

TABLE 9-3

CAPITAL COSTS SUMMARY[1]

ISOBUTANE RECOVERY SYSTEM

ABSORPTION vs MECHANICAL REFRIGERATION

DIRECT COSTS	ABSORPTION REFRIGERATION	MECHANICAL REFRIGERATION
(E) Equipment (fob)	$ 73,583.	$ 84,779.
(m) Field Materials	45,621.	52,563.
(L) Direct Field Labor	42,678.	49,172.
Total Direct[2] M & L	161,882.	186,514.
INDIRECT COSTS		
Insurance, etc.[3]	11,920.	13,734.
Labor Indirects[4]	29,021.	33,437.
Total Indirects	40,941.	47,171.
Base Module Cost	202,823.	233,685.
Engineering[5]	16,188.	18,651.
Sub Total	219,011.	252,336.
Contingency & Fees[6]	21,901.	25,236.
Total Installed Module Costs	$240,912	$277,570.

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[1] Modular Concept and Direct Cost Factors from Guthrie-Reference (11)

[2] Total Direct M & L = E + m + L

[3] Includes Freight, Taxes, Insurance, etc. = 10% of (E +M)

[4] 68% of L

[5] Construction Engineering @ 10% of (M & L)

[6] 10% of Subtotal for Contingency & Contractors Fees

TABLE 9-4

ANNUAL OPERATING COST SUMMARY[1]

ABSORPTION vs MECHANICAL REFRIGERATION

COST ITEM	ABSORPTION REFRIGERATION	MECHANICAL REFRIGERATION
A. Make-Up Refrigerant	$ 922.	$12,850.
B. Electricity	8,600.	13,478.
C. Steam	3,100.	-0-
D. O & M Labor[3]	90,000	90,000
E. Maint. Parts & Supplies[4]	4,818.	5,551.
F. Misc. Process Materials[5]	18,000.	18,000.
Total Operating Costs	$125,440.	$139,879
Insurance & Taxes[6]	7,227.	8,327.
Depreciation[7]	22,221.	25,499.
Total Production Cost	$154,888.	$173,705.

[1] 90% Operating Factor

[2] Make-up refrigerant- Ammonia water for absorption and Freon 12 for compression - Consumption taken at 1/10 of 1%-circulated-priced at 1979 costs.

[3] One operator and one maintenance man per shift plus 25% payroll burden and 50% overhead

[4] 2% of total module cost

[5] 20% of O & M labor

[6] 3% of total capital costs

[7] Calculated at straight line, 15 years, or 5% of total investment less interest

Capital investment is the total amount of money needed to supply the necessary isobutane recovery and support facilities plus working capital required for operation of these facilities. These are tabulated on Table 9-5. In summary, the total investment capital required for the absorption refrigeration isobutane recovery system is $333,315 and for a recovery system using mechanical compression refrigeration is $382,487.

Based on these preliminary cost estimates it appears that an isobutane recovery system incorporating absorption refrigeration is economically more feasible than one using mechanical compression refrigeration.

The overall profitability from recovery of the isobutane as well as the effective value figured on a ¢/kwh electricity basis are key factors in deciding whether capital should be invested in the recovery scheme. The method used in determining the profitability is based on the ROI concept. ROI is a simple financial accounting criteria for measuring profitability of investments that makes no allowance either for time value of money or the time pattern of investment and income. However, it is simple to determine and is accurate as a first hand approximation.

The return on investment (ROI) as used in this context is defined as:

$$\text{ROI} = \frac{\text{Net Annual Savings - Total Production Costs}}{\text{Total Capital Investment}} \text{X100}$$

Figure 9-5 shows the overall return on investment of an isobutane recovery system using absorption refrigeration for various recovery rates and unit costs of isobutane. For a 50 MW_e power generating facility, it is expected that 900-930 lbs/hr of isobutane may be recovered for an ROI of over 200%.

TABLE 9-5

TOTAL CAPITAL INVESTMENT
FOR ISOBUTANE RECOVERY SYSTEMS

	ABSORPTION REFRIGERATION	MECHANICAL REFRIGERATION
Total Installed Module Cost	$ 240,912.	$ 277,570.
Off Sites[1]	24,091.	27,757.
Total Facilities Cost	265,003.	305,327.
Interest During Construction[2]	21,200.	24,426.
Owner's Cost[3]	5,300.	6,107.
Start up Costs	10,453.	11,657.
Total Capital Costs	$ 301,956.	$ 347,517.
Working Capital[5]	31,359.	34,970.
Total Investment Capital	$ 333,315.	$ 382,487.

[1] Allowance for buildings, fences, roads, shops, etc., 10% of Total Module Cost

[2] 8% of total facilities costs

[3] 2% of total facilities costs

[4] One-month operation at full capacity

[5] Three-month operation at full capacity

FIGURE 9-5

ISOBUTANE RECOVERY-50 MWe POWER FACILITY

ROI vs RECOVERY RATE AND UNIT COST OF C_4H_{10}

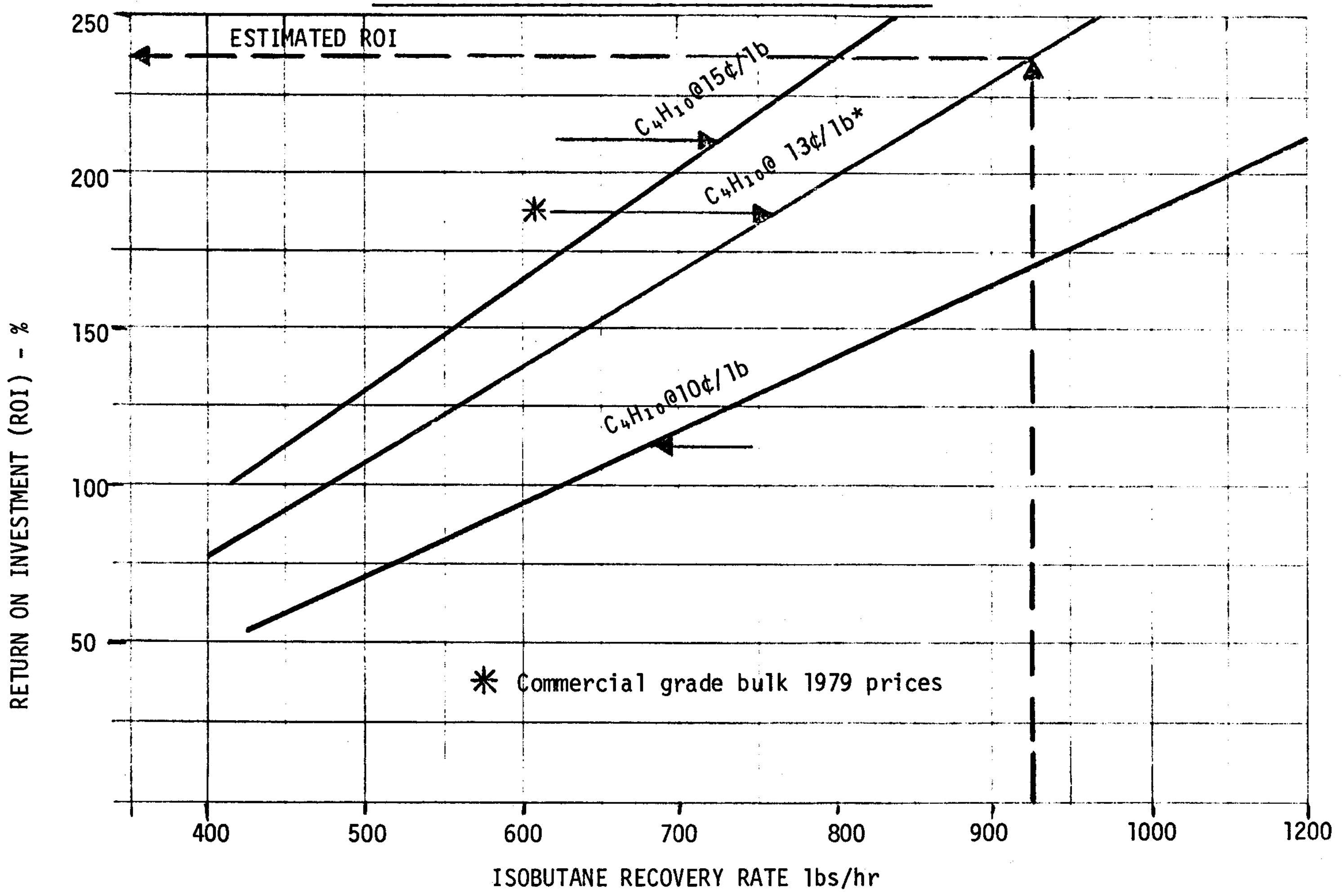
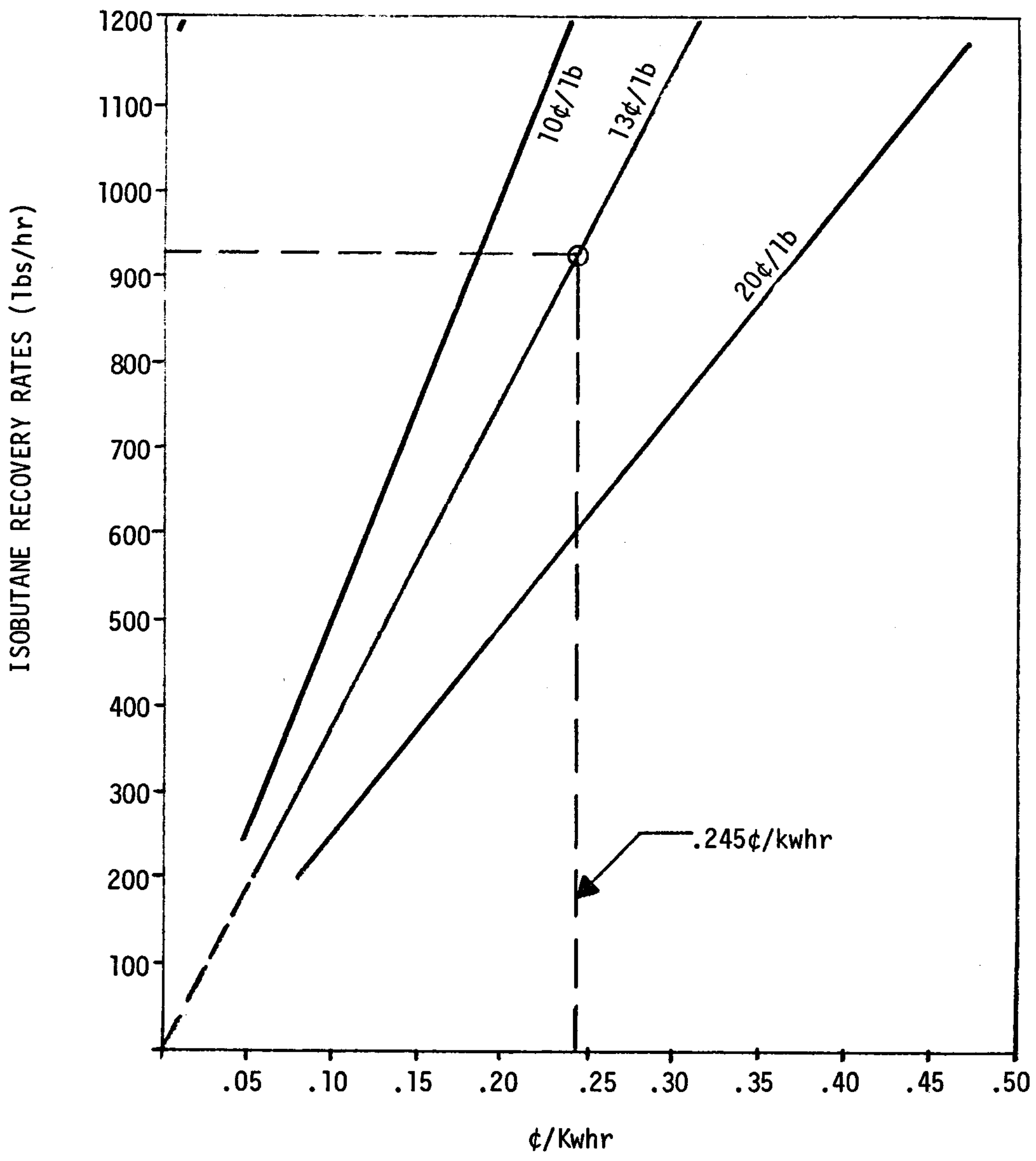


Figure 9-6 is a plot of isobutane recover rate vs net savings in ¢/khr electricity generated and unit cost of isobutane that may be realized by including a recovery scheme at a 50 MW_e facility. Based on the expected recovery rate of 930 lbs/hr at present isobutane costs of 13¢/lb this savings amounts to 2.42 mills per KW/hr.

FIGURE 9-6

ISOBUTANE RECOVERY RATE
VS
¢/KWHR



SECTION 10

DIRECT-CONTACT CONDENSER TESTS

SECTION 10

DIRECT-CONTACT CONDENSER TESTS

10.1 Introduction

In April, 1979, DSS Engineers, Inc., was granted a contract extension (change to P.O. No. 62X-45736V) by Oak Ridge National Laboratory for additional testing at East Mesa. The object of the extension was to operate the direct-contact energy test loop coupled with a direct-contact working fluid condenser in a series of condenser performance evaluations. Specifically, the added contract task provided for (1) two weeks of field preparation and equipment changeover from a previous isobutane run using the original shell-and-tube working fluid condenser, (2), seven weeks of field testing and data collection using the direct-contact condenser (DCC) and isopentane, and (3), four weeks of data reduction and report preparation. The additional work extended the contract through July 1, 1979.

10.2 Field Preparation and Changeover.

The following test loop modifications and maintenance work were completed to assure safe and continuous operation during the DCC testing:

a. A separator was welded into place on the rig and the DCC unit was bolted to it.

b. A support frame was built to hold the DCC unit in place and prevent vibration of the unit.

c. The systems piping was modified to incorporate the DCC unit and separator.

d. The hairpin surface heat exchanger was descaled.

e. Pressure relief valves and pressure switches were checked for proper operation.

f. Rupture discs were inspected for pitting or corrosion and new discs installed where necessary.

g. The system was hydro-tested.

h. Temperature gauges, thermocouples and pressure guages were checked and recalibrated or replaced where necessary.

10.3 DCC Testing Using Isopentane

A first objective was to operate the "hairpin" surface heat exchanger direct-contact condenser loop to collect the baseline data necessary to evaluate the heat transfer performance of the direct-contact condenser and determine the amount of working fluid that would be contained in the water exiting from the separator. The direct-contact heat exchanger-DCC loop could then be operated to collect comparison data. Isopentane was used as the working fluid in all tests. The salient points with regard to actual field testing are outlined below.

Figure 10-1 is a sketch of the DCC and Figure 10-2 shows the flow scheme and loop components. The DCC and separator were piped as a by-pass module to the existing shell-and-tube condenser, which was completely isolated from the circuit by valves on all inlet and outlet lines.

Condensing water was pumped from the cooling tower through a full-cone spray nozzle located within the condenser to make intimate contact with the isopentane vapor. The water and the isopentane condensate then flowed into the new separator where the separation of the two immiscible liquids took place because of the difference in the specific gravities and the interfacial tension.

The water, being more dense, settled to the bottom of the separator,

FIGURE 10-1

DIRECT CONTACT CONDENSER-SEPARATOR MODULE

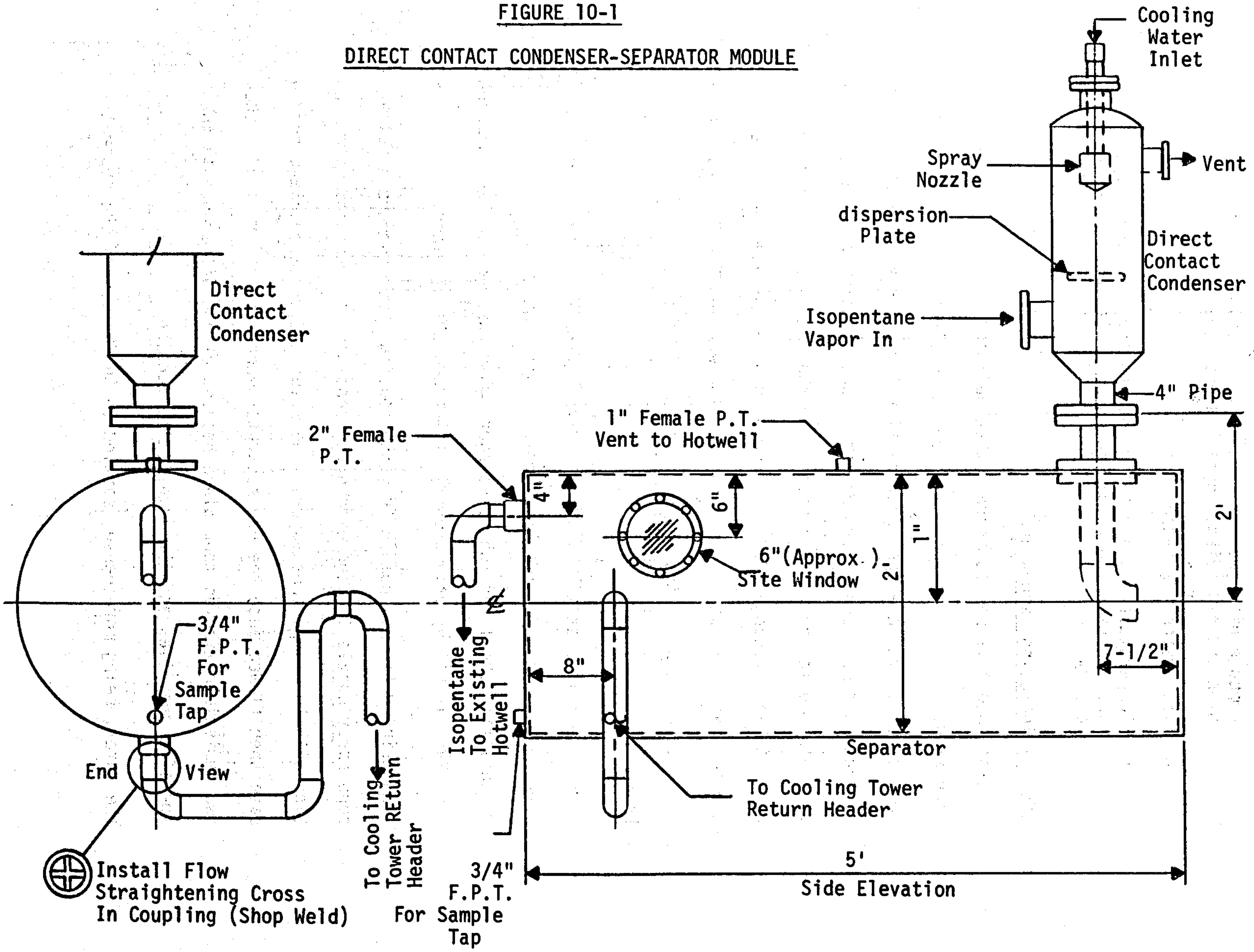
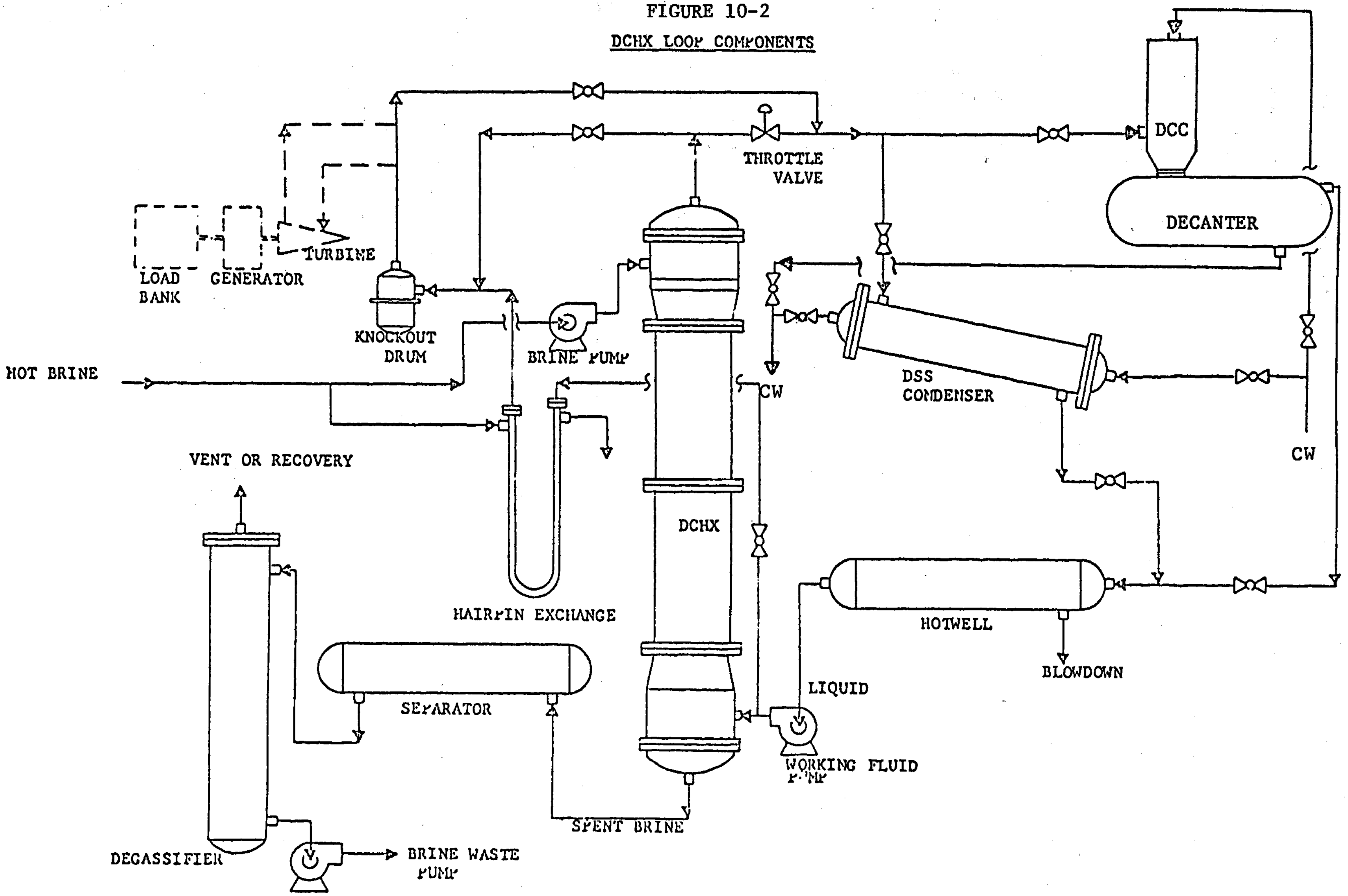


FIGURE 10-2

DCHX LOOP COMPONENTS



was removed as an "undercut", and flowed back to the cooling tower by gravity. The isopentane, being more buoyant, rose in the water and was removed from the separator as a "topcut" and flowed to the working fluid hotwell. Any water flowing from the new separator with the isopentane was removed from the system as hotwell blowdown.

Isopentane certified at a purity of 97+ percent was obtained in sufficient quantity to conduct all required tests. Samples of isopentane were taken and the purity verified at the East Mesa Laboratory facility. The analyses obtained are given below:

COMPONENT	STANDARD	CYLINDER #1	CYLINDER #2
Isopentane	97.2	99.870	99.780
N-pentane	2.7	0.000	0.000
N-butane	0.0	0.078	0.148
Unidentified	0.1	0.012	0.020

The system was charged with the isopentane after the hydrotesting was complete. It was necessary to add makeup isopentane to the system periodically because some isopentane would leave the separator with the condensing water as "carryunder".

Baseline data was first collected using the hairpin surface heat exchanger. During this phase of the task, startup and control of the system and operating procedures were established. It was found that the DCC unit and the separator had to be initially pressurized with isopentane vapor before the isopentane return valve was opened so that flooding of the hotwell with cooling water could be eliminated. A fixed cooling water flowrate was chosen and the cooling water return valve was used to regulate the cooling water level in the separator.

After the cooling water level was adjusted to its proper level, the isopentane return valve was opened and the system was left to stabilize with minor adjustments being made as required. Once stability was achieved, baseline data could be collected for various flow rates of isopentane and water.

After baseline data were collected, attempts were made to obtain comparative data using the DCHX-DCC loop. However, achieving stable conditions for data collection using the DCHX was somewhat more difficult. The DCHX was more sensitive to pressure and temperature fluctuations from the well head, flash tank and DCC module than the hairpin surface-type heat exchanger. Furthermore, any change in the DCHX operation caused unstable conditions in the DCC separator module. Some successful test runs were made, however, and sufficient data were collected for comparison purposes.

Two main problems were encountered during the test period. The first was that the isopentane pump was unable to provide the high flow rate used in sizing the DCC module; the system thus had to operate at less than design capacity. Second, control of the water level in the separator was difficult because of temperature and pressure flucuations from tne well head and inconsistencies in the isopentane flow rate. At times, the water level would rise high enough to flood the hotwell, causing a drop in the isopentane temperature and pressure and result in unstable operating conditions. The system would then have to be shut down and the water drained from the hotwell before proceeding with the test.

10.4 Data Evaluation and Reporting

A. Carryunder

Though the DCC performed well, the separator had some shortcomings. The most crucial problem in the separator was the amount of carryunder. This is defined as the amount of isopentane that is carried out of the separator with the cooling water. Based on work conducted at the University of Utah[12], it could be predicted that a small amount of carryunder (60-130 ppm) would take place due to the solubility of isopentane in water at the temperatures and pressures measured during the test. The theoretical solubility values are based on N-pentane because the thermodynamic similarities with isopentane and the availability of data for N-pentane. (Figure 10-3). Table 10-1 gives the operating conditions, theoretical solubility values and carryunder data as determined from the samples taken from the cooling water return and analyzed by the East Mesa Laboratory.

The laboratory analysis procedure consisted of flashing part of a cooling water sample into an evacuated sample bomb, then injecting this flashed sample into a gas chromatagraph for analysis. The samples showed carryunder amounts well above the predicted values. There is no apparent correlation of the carryunder data with the process variables of flow rate, temperature and pressure.

A possible explanation for the large discrepancy between theoretical and actual values is the small dispersed droplets of isopentane that were observed rising through the cooling water in the separator and collecting at the cooling water-isopentane interface. These droplets

FIGURE 10-3

SOLUBILITIES OF N-PENTANE IN FRESH WATER

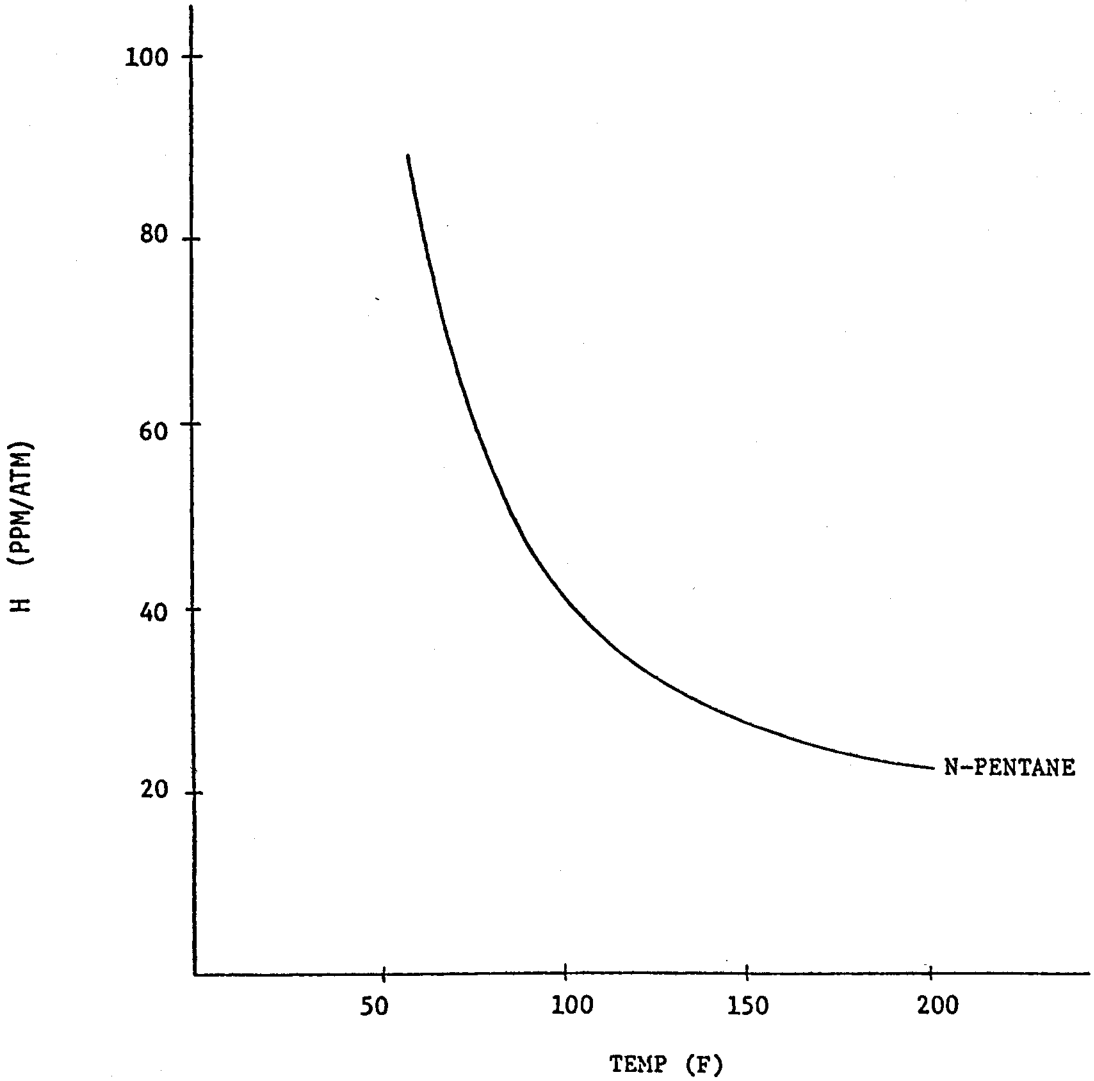


TABLE 10-1

CARRYUNDER TEST DATA AND RESULTS

SAMPLE #	THEORETICAL DATA			ACTUAL DATA			COOLING WATER FLOW RATE (GPM)	ISOPENTANE FLOW RATE (GPM)
	TEMPERATURE (°F)	PRESSURE (psia)	CARRYUNDER (ppm)	TEMPERATURE (°F)	PRESSURE (psia)	CARRYUNDER (ppm)		
1490	77.0	29.7	129.48	77.0	36.7	4525.67	24.0	2.88
1494	77.0	29.7	108.24	100.0	36.2	354.35	26.0	5.52
1497	77.0	29.7	101.98	92.0	33.2	2213.22	24.0	2.88
1500	123.3	30.7	93.43	98.0	32.7	2144.53	23.0	7.44
1501	77.0	35.7	95.39	103.00	34.2	5415.66	22.0	6.00
1502	77.0	35.7	99.76	95.0	31.2	5795.96	33.0	5.28
1503	77.0	35.7	108.90	91.0	31.7	4431.75	41.0	5.52
1504	77.0	35.7	62.04	88.0	35.7	4311.90	23.0	3.60
1505	125.6	34.7	120.84	96.0	38.2	1874.24	11.5	4.80
1506	116.6	36.7	110.95	94.5	34.7	4114.23	34.0	7.32
1508	125.6	37.7	92.08	93.0	28.2	3285.72	10.0	2.40
1509	111.2	35.7	93.24	106.0	34.7	1119.10	10.0	3.60
1512	127.4	30.7	80.82	105.0	29.7	252.33	20.0	3.60
1514	120.2	30.7	91.07	111.0	35.7	1929.91	20.0	4.80
1515	127.4	34.7	92.32	102.0	32.7	4978.03	31.0	4.80
1518	122.0	29.7	111.39	75.0	26.2	4664.43	40.0	2.4
1519	122.0	28.7	103.75	89.0	30.2	4661.32	40.0	7.20

are introduced into the cooling water in the DCC and become dispersed in the water at the separator inlet, where turbulent mixing occurs due to the impinging of the fluid stream on the separator wall.

Droplets of isopentane near the cooling water-isopentane interface rise to the interface due to the difference in the specific gravities of the liquids. However, some isopentane droplets may be caught in the turbulent currents, drawn downward and carried along the bottom of the decanter. These droplets may thus be swept out of the separator through the cooling water drain before they have time to rise to the liquid-liquid interface.

Another hypothesis for the discrepancy is that an emulsion may be formed in the DCC and a double emulsion may exist in the separator. As direct contact is made, the isopentane vapor would yield heat to the water mist and condensation of the isopentane on the water drop would occur. This emulsion would then enter the separator and encounter turbulent mixing as the emulsified drops are impinged on the separator wall. During this mixing process it is possible that a double emulsion would be formed. The double emulsion condition would retard the separation process that occurs in the separator and cause some of the isopentane to pass through the cooling water drain. The randomness of the carryunder data can be attributed to mixing and turbulent flow in the cooling water stream.

The high rate of carryunder resulted in considerable loss of working fluid. These losses should be minimized for economic reasons. It is suggested that the design be modified to: (1) provide a longer retention time, and (2), provide less turbulent flow. These changes would allow the droplets to rise to the interface surface before reaching the cooling water drain.

Another important aspect of carryunder is the safety considerations. The isopentane lost from the separator will flash as the pressure drops in the cooling water lines and at the cooling tower and result in a potential hazard for an explosion.

B. Heat Transfer

The heat transfer properties of the DCC were evaluated from the collected data. The heat duty, which is defined as the amount of heat that is available to be transferred per unit time, was calculated from the following expression:

$$Q_a = m_a \, (\Delta h)_a$$

$$Q_b = m_b \, (\Delta h)_b$$

where: Q_a = total heat duty of water
Q_b = total heat duty of isopentane
m_a = mass flow rate of water
m_b = mass flow rate of isopentane
Δh = change in enthalpy

After determining the total heat duty for isopentane and water, a percent difference in heat duty was calculated using the following equation:

$$\Delta\% = \frac{Q_a - Q_b}{Q_a} \times 100$$

These percent differences were categorized and a select group was chosen for more complete evaluation. The data chosen showed the least amount of error between the isopentane and water heat duties, and included enough data points to be able to plot the necessary curves. Table 10-2 is a summary of the data that was used and the results of the evaluation.

The volumetric heat transfer coefficient for the DCC was evaluated using the following definition:

$$U_v = \frac{Q_{cond}}{V \, \Delta T_{lm}}$$

where:

U_v = volumetric heat transfer coefficient

Q_{cond} = heat duty of isopentane for condensation

V = active condensation volume

ΔT_{lm} = apparent log mean temperature difference for condensation

Column 10 in Table 10-2 shows that higher values of U_v were possible at higher isopentane mass flow rates. (The flow rates are shown in Column 4).

The terminal temperature difference (TTD) is the difference between the isopentane condensing temperature and the calculated DCC temperature of the cooling water out. Figure 10-4 is a plot of the TTD vs isopentane flow for constant cooling water flows. This shows that at lower isopentane flow rates, the TTD is greater. In general, condensing performance was characterized by TTDs of between 27°F and 52°F, leaving considerable driving force for subsequent heat transfer in the form of subcooling in the separator.

TABLE 10-2

DCC TEST DATA AND RESULTS

Test	T_{sat} (°F)	$(T_{H_2O})_{in}$ (°F)	M_{IC5} (Lbm/hr)	M_{H_2O} (Lbm/hr)	$Q_{Superheat}$ (Btu/hr)	$Q_{Condense}$ (Btu/hr)	ΔT_{LM} (°F)	TTD (°F)	U_v (Btu/hr-ft^3-F)	NTU
17-1525	131.0	62	2220	13002	32732	306428	56.4	45.5	2057	0.42
17-1550	132.9	62	2220	13002	36568	305566	58.4	47.4	1983	0.40
17-1610	133.8	62	2166	13002	31771	297794	56.7	43.9	1988	0.49
17-1625	133.8	62	2046	13002	34649	281612	60.4	50.2	1768	0.36
21-1500	134.8	68	2106	13938	32566	289612	55.7	46.0	1968	0.37
22-1235	126.4	68	1278	12498	36006	177433	50.9	44.2	1320	0.28
25-1430	128.2	67	2310	11496	26584	320513	45.9	33.3	2646	0.61
30-1400	131.0	67	2958	11250	0	410398	43.2	27.5	3597	0.84
30-1545	130.1	67	1866	10998	32121	257891	50.5	36.6	1936	0.54
31-1025	138.5	66	3042	11250	14261	415419	51.9	35.6	3034	0.71
31-1050	139.4	66	2676	11502	23993	364896	56.1	41.7	2464	0.57
31-1120	137.5	65	2160	11502	27282	295788	57.1	44.8	1963	0.43
31-1135	134.8	65	1800	11502	36179	247320	57.9	48.3	1617	0.37
31-1212	133.8	65	1536	11502	25371	210941	59.2	50.5	1350	0.31
31-1235	130.1	65	1128	11502	34579	155761	58.1	51.6	1016	0.23
31-1320	129.2	65	1128	11502	28900	156314	57.1	50.6	1037	0.24
31-1530	146.2	68	1482	5502	15913	207141	57.3	40.6	1368	0.66
1-1125	130.1	66	3204	16998	0	443214	49.9	38.0	3410	0.53
1-1145	132.0	66	3030	16998	10878	417322	52.7	41.4	2997	0.47
1-1222	131.0	66	2286	16998	25636	315713	55.2	46.5	2165	0.34
1-1300	129.2	66	1464	16500	31888	203193	56.8	50.8	1356	0.22
1-1410	128.2	66	2178	19998	26092	302241	54.3	47.1	2108	0.28
4-1530*	118.6	67	744	4999	14125	105001	40.2	30.6	991	0.52
5-1320*	131.0	71	1116	4999	6259	153877	42.8	29.2	1362	0.72
5-1335*	128.2	70	1116	4999	25802	154495	40.8	27.3	1434	0.76
8-0955*	126.4	69	1878	9996	24895	261259	43.0	31.2	2303	0.61

* DCHX Boiler Run

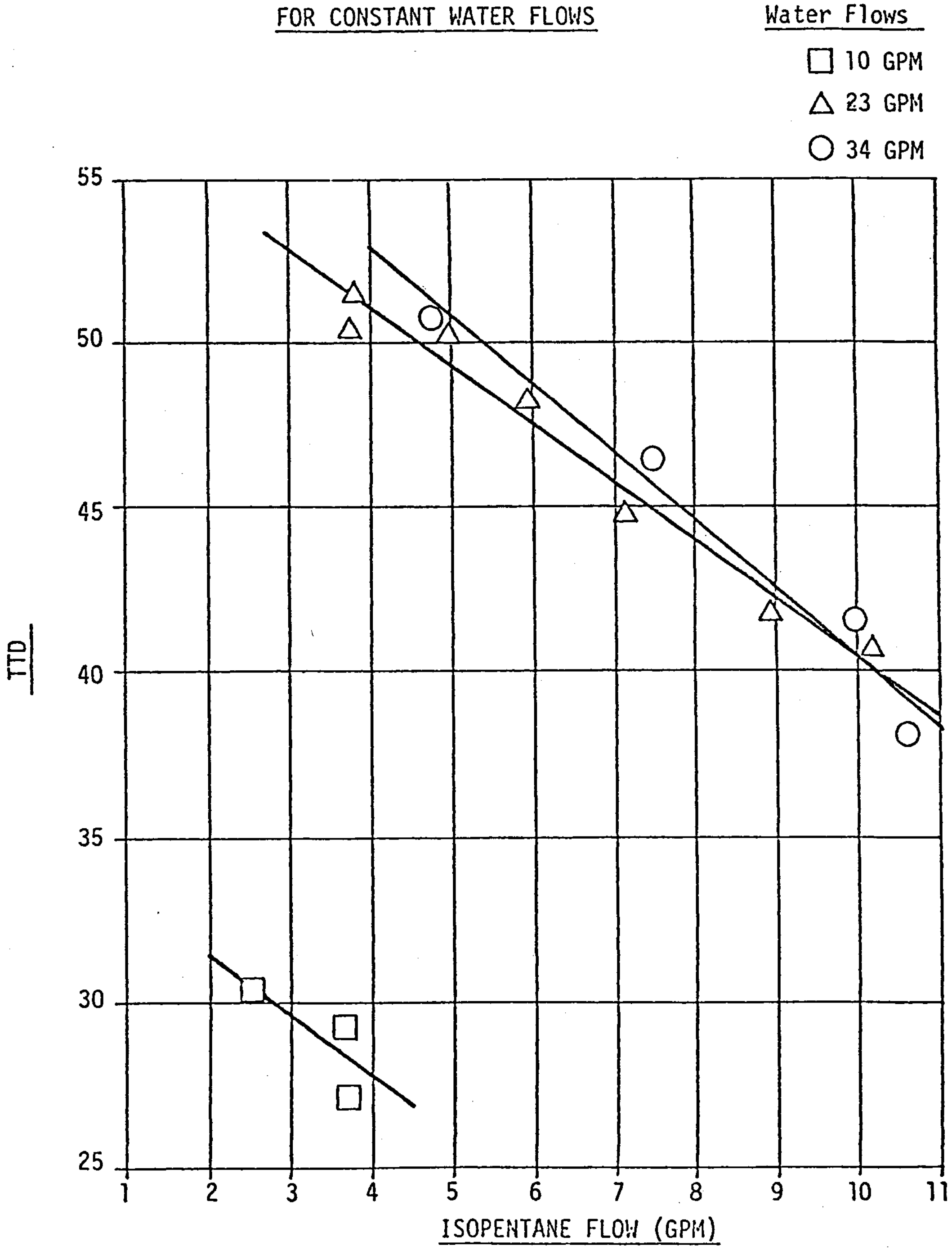
FIGURE 10-4

TERMINAL TEMPERATURE DIFFERENCE

-VS-

ISOPENTANE FLOW

FOR CONSTANT WATER FLOWS



Another parameter, the number of cooling water heat transfer units, or NTU, has been used to express the DCC performance. This parameter is defined as:

$$NTU = \frac{\Delta T_a}{\Delta T_{1m}}$$

where: ΔT_a = change of water temperature for isopentane condensing

NTU vs. isopentane flow for constant cooling water flow rates have been plotted in Figure 10-5. This plot shows that increased DCC performance was achieved with increased isopentane flow rates.

The data seems to indicate that the performance of the DCC has a strong dependence on the isopentane mass flows. Increased isopentane flow rate, at a constant cooling water flow rate could: (1) increase the U_v value, (2) decrease TTD values, and (3) increase NTU values for an overall improvement in DCC performance. To achieve higher isopentane flow rates and optimum performance, the existing isopentane pump should be replaced with a pump capable of generating higher isopentane mass throughput.

Many heat and mass transfer mechanisms may be responsible for the overall performance of the direct-contact condenser. An explanation of these phenomena could be as follows: The region just below the cooling water inlet nozzle would have the coldest temperature and lowest pressure in the DCC. This low-pressure region near the top of the DCC would tend to draw the higher pressure superheated vapor upward. It is this pressure difference that establishes the necessary

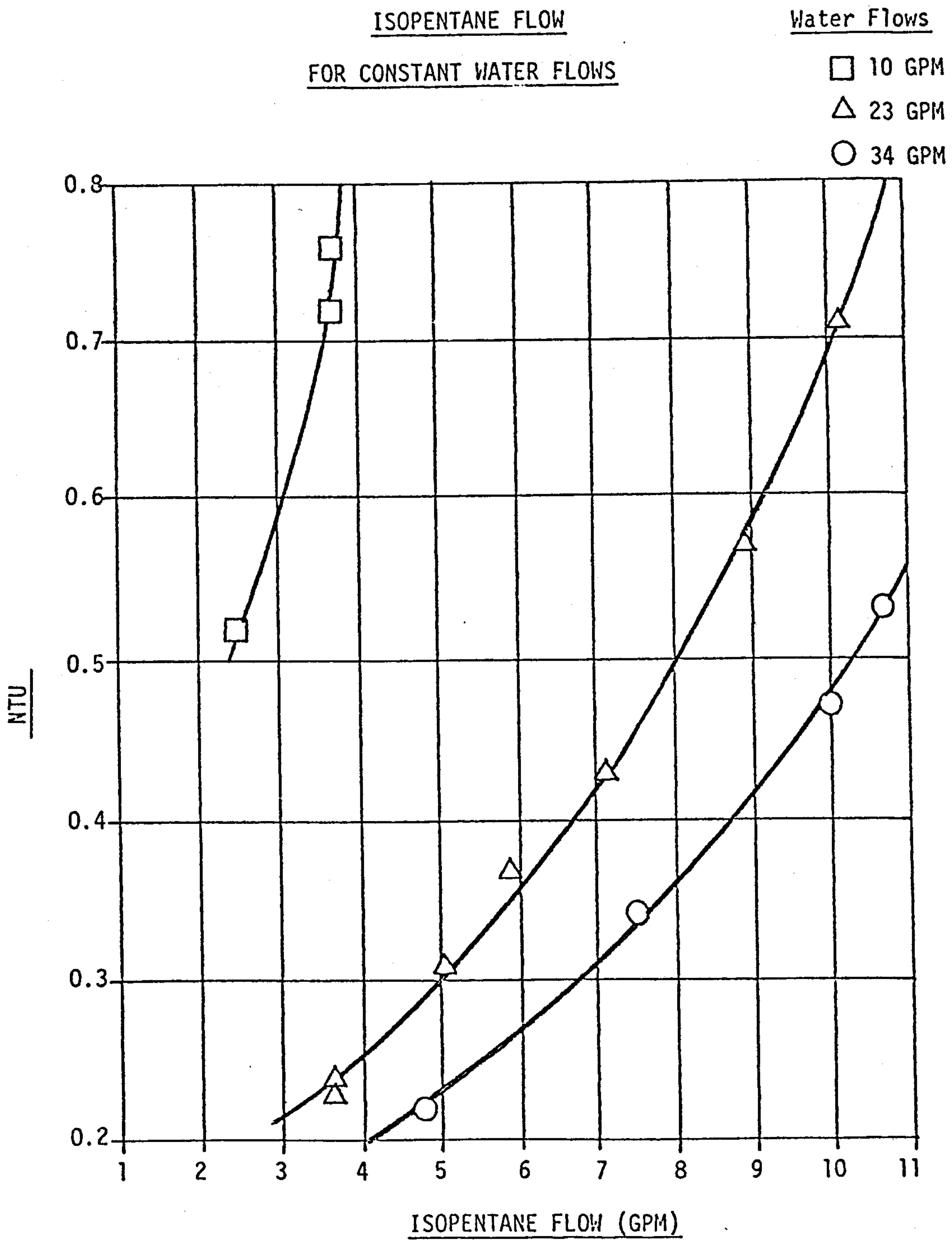
FIGURE 10-5

NUMBER OF TRANSFER UNITS

-VS-

ISOPENTANE FLOW

FOR CONSTANT WATER FLOWS



countercurrent flow that enables intimate contact between the vapor and liquid. Within this low-pressure region the isopentane would probably be condensed as a film enveloping the water drops. The upward momentum of the countercurrent-flowing isopentane should have been lost as condensation occurred and this "emulsion" of water encased in isopentane falls downward through the upwelling vapors. The emulsified drops from this cold, low-pressure region encounter warmer, higher pressure and rising isopentane vapor. In this condensing region heat and momentum are transferred to the emulsified drop from the isopentane vapor, which condenses on the droplets and increases the isopentane film thickness. The condensing region ends where the saturation temperature occurs and the emulsion enters a region where desuperheating of the superheated inlet vapor takes place. As the emulsion passes through the desuperheating region some of the condensate may be re-vaporized from the superheated vapor's sensible heat loss. This desuperheating process prepares the rising vapor for the condensing region. After passing through the desuperheating region the emulsion is directed out of the DCC and into the separator where subcooling of the isopentane transpires.

It is thus hypothesized that the positive dependence of the DCC performance on the isopentane flow rate is caused by: (1) increased water holdup (and residence time) due to momentum transfer from the condensing isopentane vapor, and (2) increased local heat transfer coefficients in and around the cooling water drops.

The experimental results of this test demonstrated that direct-contact condensation of isopentane vapor with water was possible and the DCC-separator module provided adequate heat rejection capacity for the DCE test loop. Evaluation of the collected data made it evident that the test results were not completely representative of the DCC modules' potential. Optimum operation was not achieved due to the limit imposed by the undersized isopentane pump, though the tests and data did indicate that the DCC module can operate effectively if needed in the low isopentane flow rate ranges. The results imply that further design modifications and technical development of the DCC-separator module could lead to a condensing module that would rival shell-and-tube condensing units. A DCC-separator module that is designed and operated at the optimum conditions should be as effective as conventional condensers, while having many advantages. Some of these advantages are: (1) less maintenance because there are no tubes to corrode or scale, (2) higher heat transfer rates because there is no temperature drop across tube walls, and (3) less expense due to absence of tube bundles.

A better understanding of the process mechanics was achieved and has led to some ideas for improving the condensing module. The module, consisting of a DCC and a separator, could be operated more effectively if some design changes were made. The diameter of the DCC unit could be reduced and the height could be increased. This could be done in such a way as to maintain the same active heat transfer volume yet provide a longer retention time for the condensation-revaporization process to proceed more completely. Various spray patterns and cascade systems should be tried to establish the most efficient method of mixing the cooling water with the isopentane vapor. Consideration should be given to the use of an impingement plate at the DCC outlet to help slow the fluid before entering the separator.

It is necessary to modify the design of the separator to decrease the large amount of carryunder. The retention time must be increased. This can be accomplished by: (1) increasing the length of the separator, (2) inserting a set of baffles to direct the flow in the separator, or (3), slow the flow rate before it enters the separator. Another potential problem that may be eliminated is the formation of of a double emulsion. The double emulsion should not form if the inlet fluid is allowed to impact on the isopentane layer instead of being impinged against the separator wall in the water layer.

It is recommended that further testing of a DCC and separator module be done with the above design modifications and improvements. The continued testing could yield the needed data to establish design criteria and provide a condensing module for immiscible binary fluids that is relatively inexpensive and maintenance free.

APPENDIX A

LETAN-KEHAT MODEL EVALUATIONS

APPENDIX A

Letan-Kehat Equation (Ref. 1):

$$\theta_d = \frac{t_{do} - t_{co}}{t_{di} - t_{co}} = \exp\left(-\frac{M}{r}\right)\left\{\frac{m}{r}\left[\frac{1+S}{\alpha_1}\left[1 - \exp(\alpha_1 z)\right] - \frac{S}{\alpha_2}\left[1 - \exp(\alpha_2 z)\right]\right] + 1\right\}$$

$$\theta_c = \frac{t_{ci} - t_{co}}{t_{di} - t_{co}} = \exp\left(-\frac{M}{r}\right)\left\{\frac{m}{r}\left[\frac{1+S}{\alpha_1}\left[1 - \left(\frac{r}{m}\alpha_1 + 1\right)\exp(\alpha_1 z)\right] - \frac{S}{\alpha_2}\left[1 - \left(\frac{r}{m}\alpha_2 + 1\right)\exp(\alpha_2 z)\right]\right] + 1\right\}$$

where:

$$\alpha_{1,2} = -\frac{m}{2}\left[\left(\frac{1}{M} + \frac{1}{r} - \frac{1}{P}\right) \pm \sqrt{\left(\frac{1}{M} + \frac{1}{r} + \frac{1}{P}\right)^2 - \frac{4}{Mr}}\right]$$

$$S = \frac{\alpha_1 + \frac{m}{r} - \frac{rm}{PM}\left[\exp\left(\frac{M}{r}\right) - 1\right]}{\alpha_2 - \alpha_1}$$

$$P = \frac{1}{R} + M$$

$$R = V_d/V_c$$

$$r = (\rho c_p)_d/(\rho c_p)_c$$

ρ = density

c_p = specific heat

V = superficial velocity

M = ratio of wake to drop volumes

m = volume of wake elements shed per volume of drop and unit length of column, ft^{-1} (cm^{-1})

t = temperature

θ = dimensionless temperature

sub c = continuous phase

sub d = dispersed phase

z = distance along column, or height. ft (cm)

APPENDIX B

COLUMN DIAMETER CORRELATIONS

APPENDIX B

COLUMN DIAMETER CORRELATIONS

Minard-Johnson Correlation

An empirical correlation has been presented by Minard and Johnson[5] for the relationship between the drop diameter and superficial velocities of the continuous and dispersed phases in a spray column at the flooding conditions. This equation is:

$$V_c^{0.5} = \left[47\mu_c^{-0.07}\ \Delta SG^{0.14}\ SG_c^{-0.5}\right] - \left[1.8d^{0.056}\ \mu_c^{-0.075}\left(\frac{SG_d}{SG_c}\right)^{0.5} V_d^{0.5}\right] \tag{2-1}$$

where:

d = droplet diameter, inch

SG_c = continuous phase specific gravity

SG_d = dispersed phase specific gravity

$\Delta SG = SG_c - SG_d$

μ_c = continuous phase viscosity, centipoise

V_c = continuous phase superficial velocity, ft/hr

V_d = continuous phase superficial velocity, ft/hr

The superficial velocity can be expressed as: mass flow rate/(density x column area). Therefore, the column diameter can be written as a function of droplet diameter by rearranging Eq. (2-1), i.e.

$$D = \frac{\left[\left(\frac{4M_c}{\pi\rho_c}\right)^{0.5} + 1.8d^{0.56}\ \mu_c^{-0.075}\left(\frac{SG_d}{SG_c}\right)^{0.5}\left(\frac{4M_d}{\pi\rho_d}\right)^{0.5}\right]}{\left[47\mu_c^{-0.07}\ \Delta SG^{0.14}\ SG_c^{-0.5}\right]}$$

where:

M_c, M_d = Mass flow rate of dispersed phase and continuous phase respectively. lb/hr

D = Column diameter, ft

If the constant average physical properties, mass flow rates and droplet size are defined, the spray column diameter can be determined.

APPENDIX B (con't)

Sakiadis-Johnson Model

Sakiadis and Johnson[6] have proposed a generalized theoretical equation for the flooding rate with respect to droplet size and superficial velocities. This equation has been used in the following form for correlation of published data on spray columns for the phase combinations solid-liquid, liquid-liquid, and gas-liquid.

$$1 + 1.8\left\{\left(\frac{\rho_d}{\rho_c}\right)^{1/4}\left(\frac{V_d}{V_c}\right)^{1/2}\right\} = 0.565d^{1/4}\left\{\frac{V_c^2}{g_c}\ \frac{\rho_c}{\Delta\rho}\ \mu_c^{1/4}\right\}^{-1/4} \qquad (2\text{-}3)$$

where:

ρ_d = density of dispersed phase, lb/cu-ft

ρ_c = density of continuous phase, lb/cu-ft

g_c = 4.17×10^8, ft/hr

$\Delta\rho$ = $\rho_c - \rho_d$, lb/cu-ft

μ_c = viscosity of continuous phase, lb/ft-hr

For fixed mass flow rates of continuous phase and dispersed phase and a given droplet size, Eq. (2-3) can be rewritten and used to calculate the diameter of the spray column, i.e.:

$$1 + 1.8\left\{\left(\frac{\rho_d}{\rho_c}\right)^{1/4}\left(\frac{4M_d}{\pi D^2\rho_d}\right)^{1/2}\left(\frac{4M_c}{\pi D^2\rho_c}\right)^{-1/2}\right\} = 0.565d^{1/4}\left\{\left(\frac{4M_c}{\pi D^2\rho_c}\right)^2\frac{\rho_c\mu_c^{1/4}}{g_c\Delta\rho}\right\}^{-1/4} \qquad (2\text{-}4)$$

Simplifying Eq. (2-4) and solving for D gives:

$$D = \left\{\frac{\left[1 + 1.8\left(\frac{\rho_c}{\rho_d}\right)^{1/4}\left(\frac{M_d}{M_c}\right)^{1/2}\right]\left[\left(\frac{4M_c}{\pi\rho_c}\right)^2\left(\frac{\rho_c\mu_c^{1/4}}{g_c\Delta\rho}\right)\right]^{1/4}}{0.565d^{1/4}}\right. \qquad (2\text{-}5)$$

Eq. (2-5) is used to calculate the column diameter with the assumption of constant average physical properties. Once the column diameter is determined, the throughputs of brine and isobutane may be obtained by $\frac{4M_c}{\pi D^2}$ and $\frac{4M_d}{\pi D^2}$ respectively.

APPENDIX C

PREVIOUS ISOBUTANE TEST RESULTS

APPENDIX C

PREVIOUS ISOBUTANE TEST RESULTS

CARRYOVER DATA SUMMARY - CALCULATED VALUES

Test No.	Observed Boiler Vapor Temp.(F)	Observed Boiler Vapor Pressure (psia)	Calculated Boiling Temp.(F)	C_4H_{10} (psia)	H_2O (psia)	C_4H_{10} Super-heat (F)	lbs H_2O(HW) / lb C_4H_{10}	lbs H_2O(KO) / lb C_4H_{10}	lbs H_2O(Tot) / lb C_4H_{10}	Theoret lbs H_2O / lb C_4H
1	212	259	198	248.0	11.0	14	.0166	.0018	.0184	.0095
2	225	263	199	251.5	11.5	26	.0160	.0018	.0178	.0096
3	219	213	181	205.3	7.7	38	.0184	.0018	.0202	.0084
4	216	251	195	240.6	10.4	21	.0195	.0059	.0254	.0093
5	232	194	173	187.6	6.4	59	.0248	.0021	.0269	.0080
6	223	202	177	196	7.0	46	.0431	.0289	.0720	.0082
7	163	167	161	162	4.9	2	.0070	.0034	.0104	.0073
8	210	206	178	198	7.2	32	.0161	.0016	.0177	.0083
9	213	205	178	198	7.2	35	.0175	.0014	.0189	.0083
10	208	198	175	191	6.7	33	.0161	.2080	.2241	.0081
11	226	317	216	300	15.9	10	.0162	.0007	.0169	.0105
12	226	316	216	300	15.9	10	.0167	.0007	.0174	.0105
13	223	314	215	298	15.5	8	.0158	.0012	.0170	.0104
14	205	196	174	189	6.6	31	.0165	.0031	.0196	.0081
15	230	197	174	189	6.6	33	.0249	.0017	.0266	.0081
16	225	280	205	268	12.8	20	.0413	.0032	.0445	.0099
17	215	278	204	265	12.5	11	.0077	.0020	.0097	.0098
18	222	283	206	271	13	16	.0162	.0022	.0184	.0099

APPENDIX C (con't)

PREVIOUS ISOBUTANE TEST RESULTS

CARRYOVER DATA SUMMARY - MEASURED FLOW & SALINITY VALUES

						DCHX BOILER	KNOCKOUT DRUM				HOTWELL			
		BRINE FLOW		ISO FLOW		Level[1]	Flow		Conduct.	Salinity	Flow		Conduct.	Salinity
Date 1978	Test #	GPM	lbs/hr	GPM	lbs/hr	(inches)	liter/hr	lbs/hr	(umhos)	(ppm)	liter/hr	lbs/hr	(Umhos)	(ppm)
5/31	1	5.0	2664	8.5	2318	4.5	1.9	4.2	600	338	17.4	38.4	600	354
5/31	2	6.1	2785	8.1	2211	7.2	1.8	4.0	600	324	16.0	35.3	600	358
6/1	3	4.9	2213	7.6	2097	4.0	1.7	3.7	533	288	17.5	38.5	1600	947
6/2	4	5.9	2664	8.2	2274	16.0	6.1	13.4	5200	2725	20.1	44.4	567	335
6/5	5	4.8	2168	7.4	2047	6.4	2.0	4.3	586	333	23.1	50.8	600	355
6/5	6	4.8	2168	7.7	2115	16.3	27.8	61.2	6800	3250	41.4	91.2	3186	1886
6/5	7	3.0	1355	6.8	1870	6.0	2.9	6.4	700	381	5.9	13.1	600	355
5/6	8	4.0	1806	6.9	1926	1.5	1.4	3.1	600	341	14.1	31.0	572	339
6/6	9	4.2	1897	6.9	1888	6.7	1.2	2.6	343	195	15.0	33.0	600	355
6/6	10	4.3	1942	6.5	1787	14.5	168.7	371.7	6400	3059	13.1	28.8	600	355
6/13	11	7.2	3252	9.8	2704	0	.9	2.0	*4000-1100	-	19.9	43.8	400	237
6/14	12	7.5	3387	9.8	2704	1.0	.8	1.9	450	254	20.5	45.1	400	238
6/14	13	7.5	3387	10.1	2779	12.0	1.5	3.3	385	215	19.9	43.8	425	251
6/7	14	4.2	1897	6.8	1875	.5	2.6	5.8	400	227	14.0	30.9	650	338
6/7	15	4.2	1897	6.5	1794	6.0	1.4	3.1	400	226	20.2	44.6	600	355
6/8	16	5.3	2394	7.8	2135	11.0	3.1	6.9	675	354	40.0	88.1	700	411
6/9	17	5.3	2394	8.5	2357	0	2.1	4.6	443	252	8.2	18.1	757	448
6/9	18	5.3	2394	7.9	2165	5.6	2.2	4.8	350	193	15.9	35.0	700	414

* power went off, shutdown rig, salinity up and not sufficient flow thru cell well to purge salts

[1] inches below new brine inlet - see Figure 7.2

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