

SESSION 2

REVIEW OF THE 500 KW DIRECT CONTACT GEOTHERMAL PLANT AT EAST MESA

Kenneth E. Nichols
Barber-Nichols Engr. Co.
Arvada, CO 80002

1.0 SUMMARY

The concept of a binary power cycle utilizing direct contact heat exchangers was first proposed by Jacobs and Boehm in 1973 for use with geothermal brines. This concept was proposed primarily to overcome difficulties associated with the fouling and scaling nature of many moderate temperature brines. However, thermodynamic analyses and subsequent economic analyses clearly pointed to possible economic advantages over conventional binary cycles even with non-fouling, non-scaling brines.

For a direct contact binary power plant to be economically attractive it is necessary that a small pinch point be obtainable so that a maximum amount of power can be obtained per unit mass flow of geothermal brine. Since the working fluid comes in direct contact with the brine it must be immiscible with the brine, low in cost and, if part of it goes into solution in the brine, easily recoverable. In addition, noncondensable gases from the brine must be controlled to limit their effect on condenser pressure. The 500 kWe DCHX test facility installed at East Mesa was designed to evaluate techniques to provide economical operation.

The choice of the East Mesa test site as a first location to evaluate the DCHX system placed additional requirements on the system. The brine at East Mesa was at low pressure, requiring the use of downhole pumps. The selection of isobutane as a working fluid required increasing the pressure of the brine. The high amount of dissolved CO_2 in the brine required that it be preflashed to prevent the carryover of CO_2 gas through the turbine and into the condenser which would adversely affect the system performance. All of these problems have been met by the system designer and operator, Barber-Nichols Engineering. Further, problems with

isobutane turbine design, supposed state-of-the-art, were encountered and resolved.

2.0 DIRECT CONTACT HEAT EXCHANGER

The direct contact heat exchanger at East Mesa was designed primarily using the procedures outlined in Section 4.2 of the "Sourcebook on the Production of Electricity from Geothermal Brines". The exception to this was in finding the length of the preheater. The length of the preheater was determined utilizing the following correlation for the volumetric heat transfer coefficient:

$$U_v = 45000 (\phi - 0.05) e^{-0.57R} + 600 \frac{\text{Btu}}{\text{hr ft}^3 \text{ }^{\circ}\text{F}} \quad (1)$$

The direct contact heat exchanger is shown schematically, in Figure 1, based on its original design proposed in September 1978 by Barber-Nichols Engineering. This design was essentially the final configuration. It was modified in terms of the brine injector to one where the brine was injected through the top of the vessel. By lengthening the pipe to the injection nozzle, the point along the column where the brine was injected could be altered. In addition, the level control could be adjusted by re-locating the controller.

The testing performed during the last phase of the geothermal direct contact heat exchanger program utilizing the 500 kW pilot plant provided more insight into the capabilities and limits of the direct contact approach and showed that more work needs to be done to understand the inner workings of a large direct contact heat exchanger. Testing of the column demonstrated that the performance was excellent. The system operated smoothly and was readily controlled over a wide range of operating conditions. Performance evaluation showed pinch differentials of 4 to 6^oF at design flows and better than predicted heat transfer capability. Figure 2 shows a typical heating curve.

FIGURE 1
DCHX CONFIGURATION

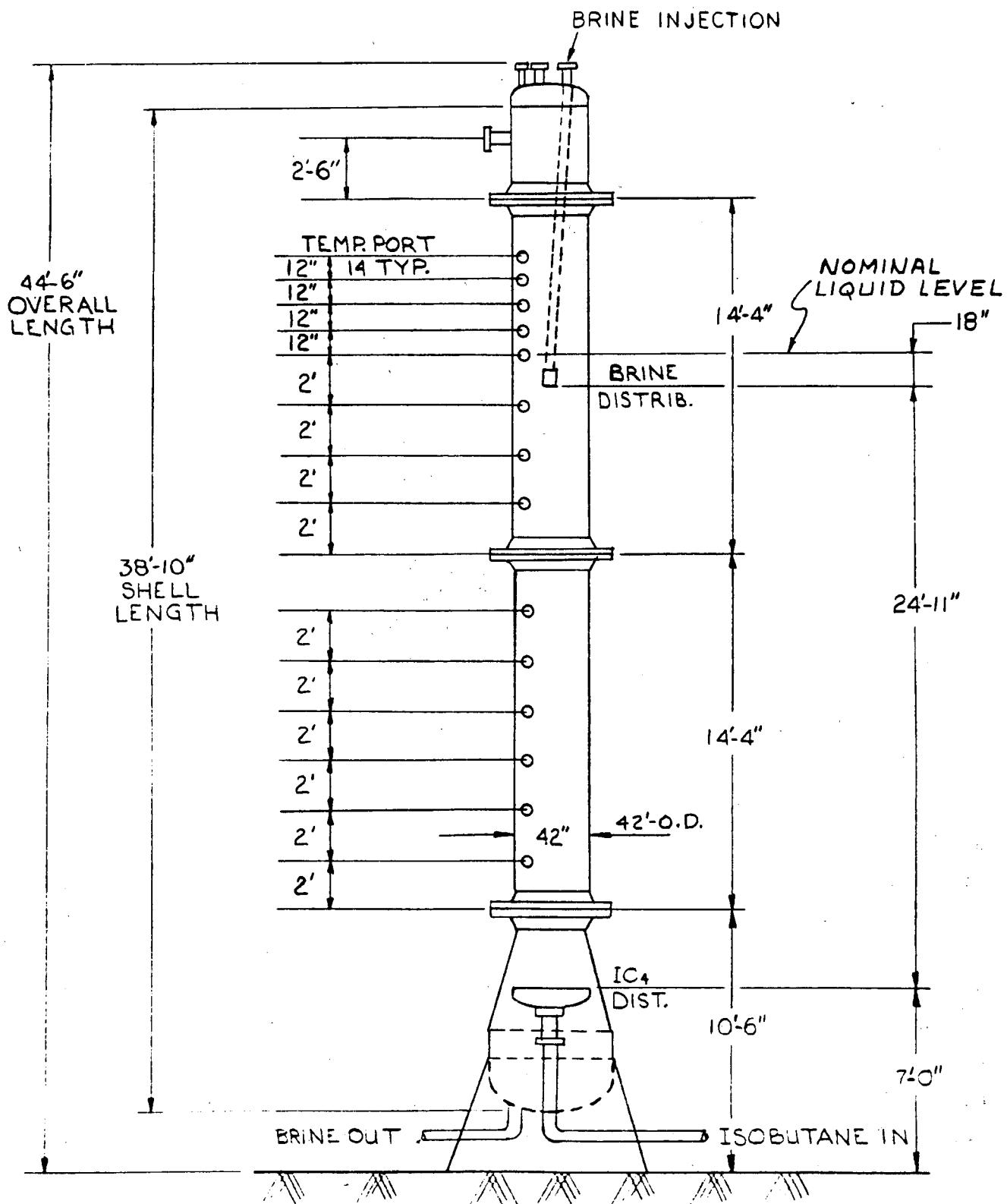
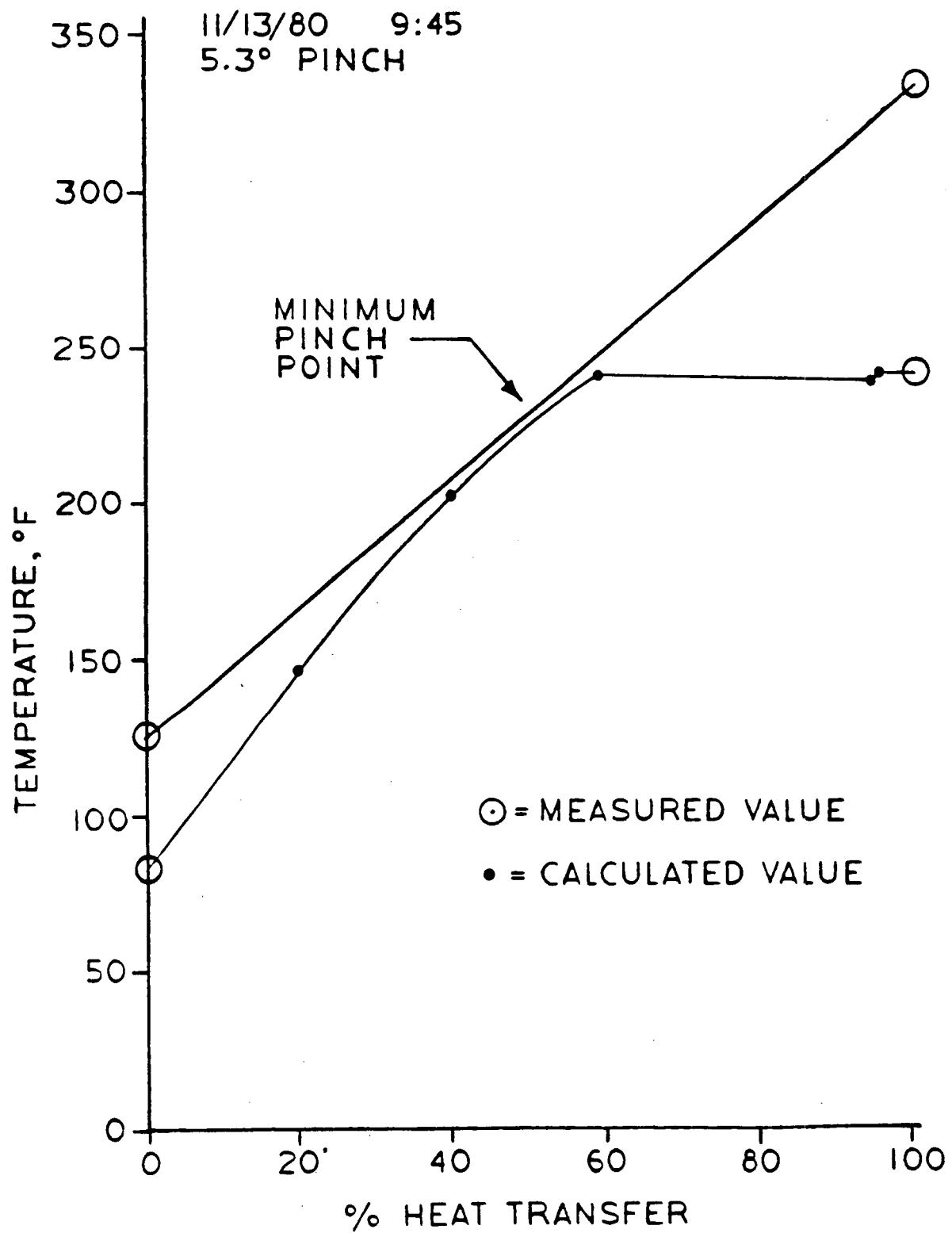


FIGURE 2
TEMPERATURE VS % HEAT TRANSFER



Testing during the final phase was directed towards establishing the limits of the column to transfer heat. The working column height was shortened progressively from the design length of 28 feet to approximately 16 feet. The short column performed as well as a full column and there are indications that the column could have been shortened even more without affecting its ability to transfer heat (see Figure 3). The column's ability to perform as well with shortened lengths as with its original design length indicates that the heat transfer coefficients and criteria derived from the previous smaller size columns are very conservative. Also, the temperature profiles recorded over the length of the column were not characteristic of a typical counter-flow heat exchanger process nor predictable based on previous small scale direct contact experiments. These abnormal temperatures indicate that the flow phenomenon through the preheater is not understood for this configuration.

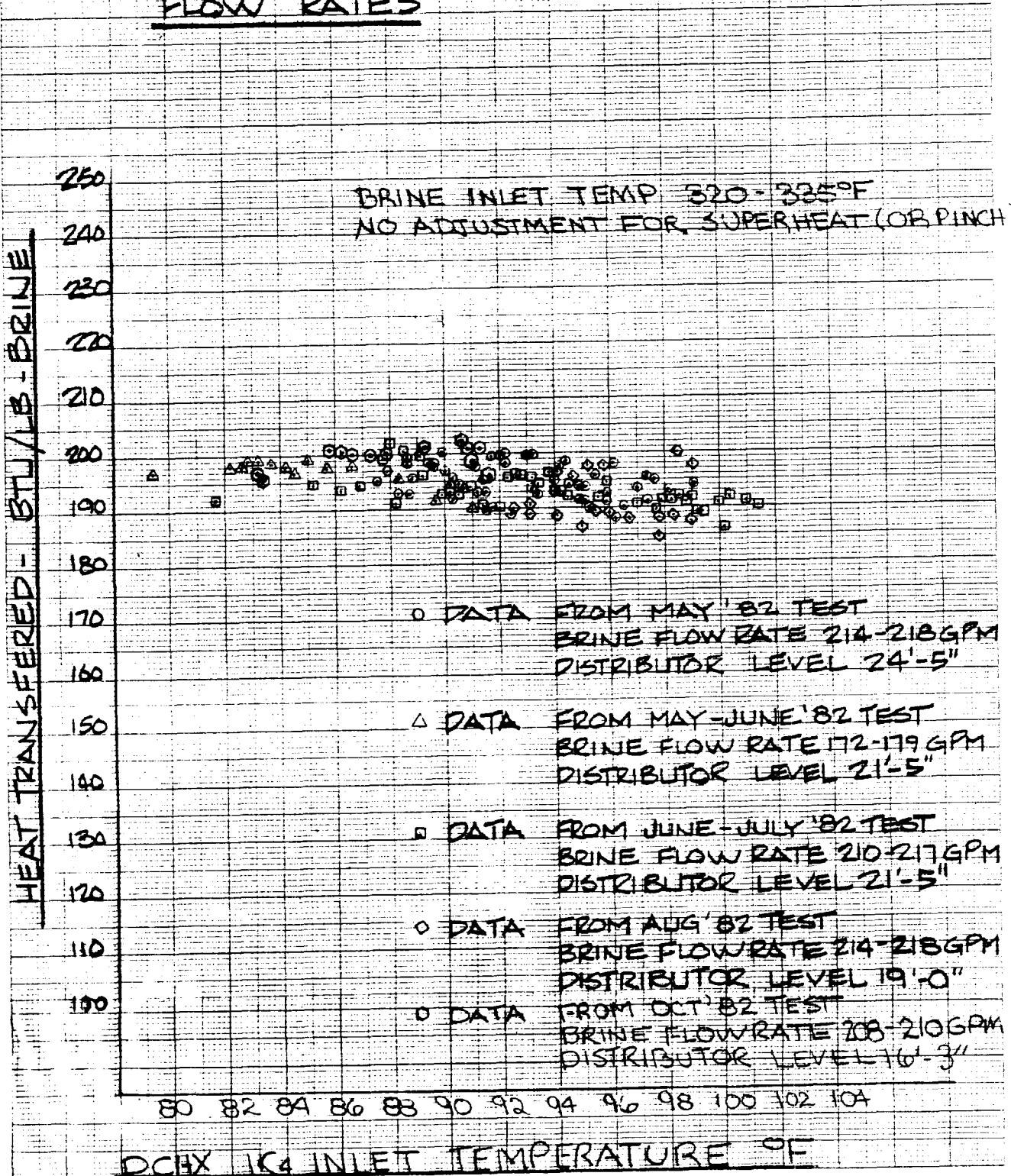
The column was operated at 110% above its rated through-put with no evidence of flooding.

3.0 WORKING FLUID RECOVERY EVALUATION

The use of a direct contact heat exchanger in recovering energy from geothermal brines is advantageous for binary power plants because it eliminates heat exchanger fouling at the heat input side of the power loop. A typical working fluid used in the direct contact approach has low miscibility in brine; however, significant amounts can still be lost from the plant unless working fluid recovery equipment is provided. For a 5 MW geothermal power plant using isobutane as the working fluid, isobutane losses amount to approximately 2¢/kW-hr if equilibrium solubility limits are reached in the effluent brine. A cost effective recovery system reduces the isobutane losses to a few mils per kW-hr, thus reducing operating costs and improving power plant economics.

There are many approaches to recovering working fluids from the brine before it leaves the plant. Both physical and chemical

FIGURE 3
DCHX PERFORMANCE WITH DIFFERENT
BRINE DISTRIBUTOR LEVELS AND
FLOW RATES



separation techniques are available. All the available recovery methods are complicated by the presence of entrained noncondensable gases, such as carbon dioxide and nitrogen. The presence of these gases requires that the recovery method be a two-step process. The first step liberates or strips the working fluid from the brine. The second step separates the working fluid from any noncondensable gases.

The stripping processes evaluated were vacuum flash extraction, gas stripping, salting and absorption with a lean oil. The most practical approach for stripping working fluid from the brine is a vacuum flash technique. The other three stripping processes are ruled out by both practicality and economics.

Four techniques were evaluated for separating the working fluid from noncondensables and returning the recovered working fluid to the power system. Three of these approaches are practical and had operating costs that were competitive. These systems had a payback of less than one year and operating costs ranging from 1.6 mils per kW-hr to 3.3 mils per kW-hr. The optimization of the compression condensation technique showed that minimum operating costs occurred when the liberated working fluid and noncondensable gases were returned directly to the power loop, thus eliminating the need for the separation condenser. Even after accounting for the power loss penalty due to the increased levels of noncondensable gases in the system, the operating cost was 1.4 mils per kW-hr. This approach will work for any working fluid and any noncondensable gases normally present in the brine. Figure 4 shows the optimization of the direct return system. As compressor power is increased, the cost of lost IC_4 goes down but power and equipment costs are increasing. The minimum cost occurs at approximately 13 kW of compressor power. This power and the costs shown in Table I are for a 5 MW DCHX plant.

4.0 CONTROL OF NONCONDENSIBLE GASES

In the direct contact process, noncondensable gases are stripped from the brine by the IC_4 and are carried to the

FIGURE 4

OPERATING COST OF TWO STAGE
FLASH WITH DIRECT RETURN TO
POWER CONDENSER

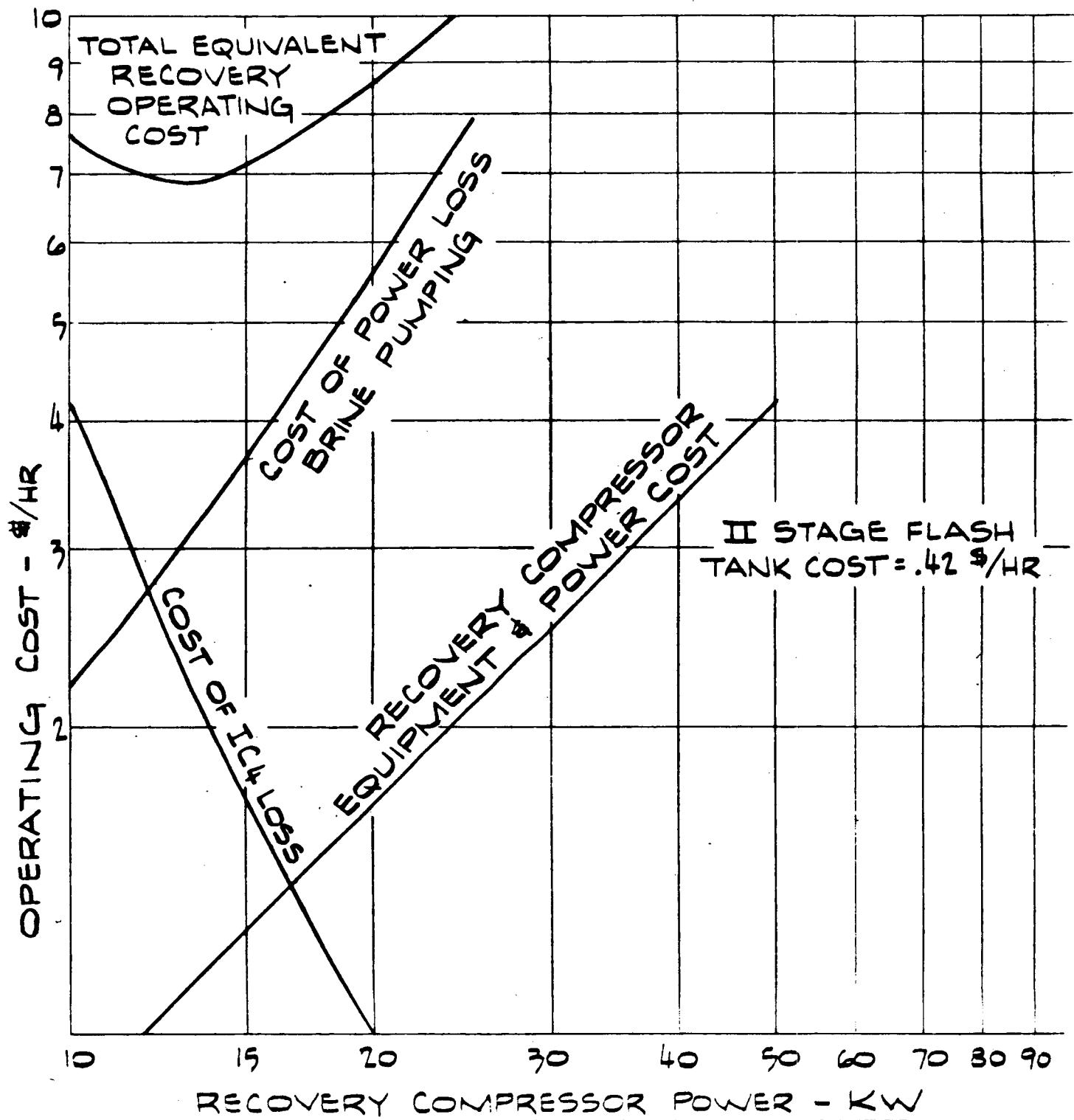


TABLE I

DIRECT CONTACT IC_4 RECOVERY SYSTEMS
COST COMPARISON

	<u>Equipment Costs</u>	<u>Power Costs</u>	<u>Lost IC_4 Costs</u>	<u>Total Equivalent Operating Costs</u>
Two-Stage Flash with Permeable Membrane*	\$181,800	\$4.91/hr	\$4.55/hr	2.8 mils/kW-hr
Two-Stage Flash with CO_2 Absorption*	\$257,400	\$2.44/hr	\$4.55/hr	2.7 mils/kW-hr
Two-Stage Flash with IC_4 Absorption (Lean Oil)*	\$285,800	\$4.69 /hr	\$4.55/hr	3.3 mils/kW-hr
Two-Stage Flash with 95°F Condenser	\$ 29,500	\$4.37/hr	\$2.90/hr	1.6 mils/kW-hr
Two-Stage Flash with 35°F Condenser & Electric Powered Refrig.	\$ 30,800	\$4.11/hr	\$2.90/hr	1.6 mils/kW-hr
Absorption Refrig.	\$ 30,800	\$3.96/hr	\$2.90/hr	1.5 mils/kW-hr
Two-Stage Flash with Direct Return to Power Loop	\$ 27,400	\$3.89/hr	\$2.25/hr	1.4 mils/kW-hr

* Systems so marked were not penalized for CO_2 returned to power loop

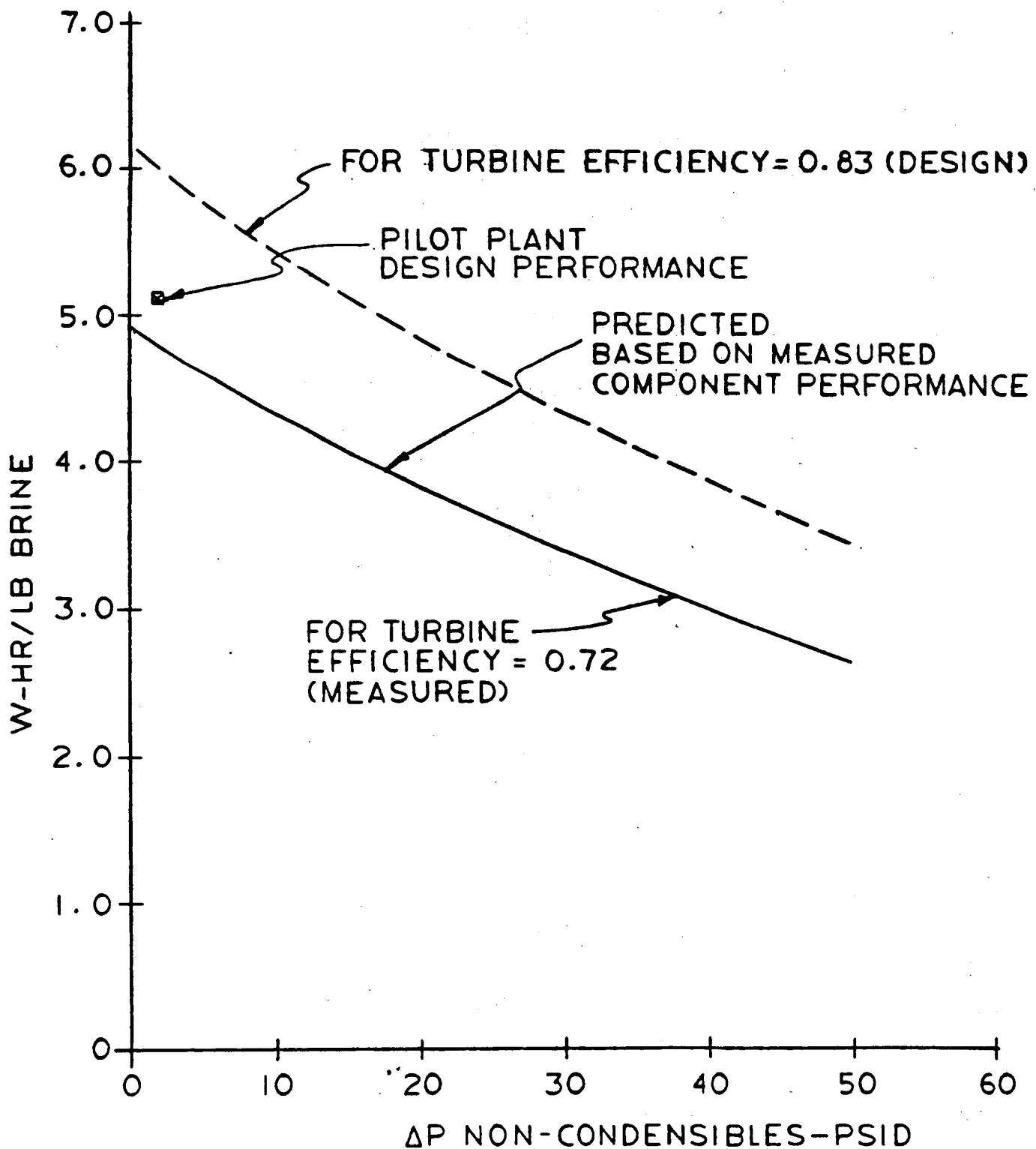
condenser. These gases tend to reduce power production and system efficiency due to increased turbine back pressure. Although the heat transfer process is enhanced by the large surface area generated by the IC_4 droplet swarm rising in the brine stream, the process promotes the mass transfer of noncondensable gases dissolved or entrained in the incoming brine into the working fluid stream. After passing through the power turbine, the noncondensing gases separate from the working fluid in the condenser where the concentration builds until the gases redissolve in the liquid working fluid and an equilibrium condition in the loop is established. The noncondensable buildup elevates the pressure in the condenser, reducing turbine output power and resource utilization. This effect is shown quantitatively in Figure 5. While trace amounts of N_2 , CH_4 , and H_2S can be found in the brine used, by far the most significant noncondensable is CO_2 .

Control of the condenser pressure elevation due to noncondensables can be achieved by flashing the incoming brine to control the quantity of noncondensables remaining in the brine injected into the DCHX. Flashing is accomplished by spraying brine into a vessel through two spray nozzles to provide single-stage flash separation of CO_2 and other dissolved gases. The steam/ CO_2 vapor mixture produced is vented from the vessel and directed to a shell-and-tube heat exchanger where thermal energy is recovered by pre-heating and vaporizing a small sidestream of IC_4 . Early analysis of CO_2 equilibrium in the process loop indicated that reduction of dissolved CO_2 to 50 ppm in the incoming brine would reduce CO_2 buildup in the condenser to approximately 2 psi. For noncondensable levels of 800 ppm, typical of brine from East Mesa well 8-1, this corresponds to a total flash of about 5°F . For the 2000 ppm levels found in well 6-2, a total single-stage flash of about 20°F is required.

Carbon dioxide levels were monitored in the brine out of the primary flash tank, in the IC_4 in and out of the DCHX, and in the condensate vapor phase in the condenser exit manifold over a range of flash conditions.

FIGURE 5

VARIATION OF PLANT PERFORMANCE DUE TO
TURBINE BACK PRESSURE EFFECTS WITH
NON CONDENSIBLE ACCUMULATION IN
CONDENSER



At equilibrium, CO_2 released from the brine at the top of the DCHX is replaced by CO_2 injected into the column with the IC_4 . There is an insignificant amount of CO_2 dissolved in water leaving the hotwell and a minor amount returning to the hotwell from the recovery system; therefore, brine leaving the DCHX has the same level of CO_2 as when it enters. Likewise, the compositions of the working fluid in and out of the DCHX must be the same (except for the water fraction) when the system is at equilibrium.

The measured concentration of dissolved CO_2 leaving the sand trap (flash vessel) is shown in Figure 6 along with a comparative theoretical prediction. While some deviation from the predicted value is apparent, reasonable agreement is shown.

A correlation of the pressure elevation in the condenser, as a function of dissolved CO_2 entering the DCHX, is shown in Figure 7. The pressure elevation is obtained by subtracting the sum of the saturation pressures of IC_2 and H_2O at the measured condensate temperature from the measured total pressure of the condensers. The pressure elevation, therefore, includes the effect of other noncondensables (principally propane found in the IC_4 feedstock) and any subcooling of the condensate. The indicated theoretical line is based on a concentration factor of 7.9 across the DCHX and a residual elevation due to other noncondensables of 2.5 psi and an equivalent 3 psia due to condensate subcooling.

Flashing the brine to reduce dissolved CO_2 levels resulted in reduced pressure elevation due to CO_2 accumulation in the condenser. Extensive flashing, however, adversely affects the electrical generating potential of the resource by reducing the temperature of the brine sent to the DCHX. While this effect is minimized by recovering the latent and sensible energy of the steam/noncondensable mixture in a supplementary IC_4 heater, the overall plant performance test data displays only modest sensitivity to flash pressure, indicating the improvement in condenser pressure elevation was largely offset by the resulting brine inlet temperature loss (see Figure 8).

FIGURE 6
CO₂ CONCENTRATION IN BRINE
LEAVING SAND TRAP AS A
FUNCTION OF SAND TRAP
PRESSURE.

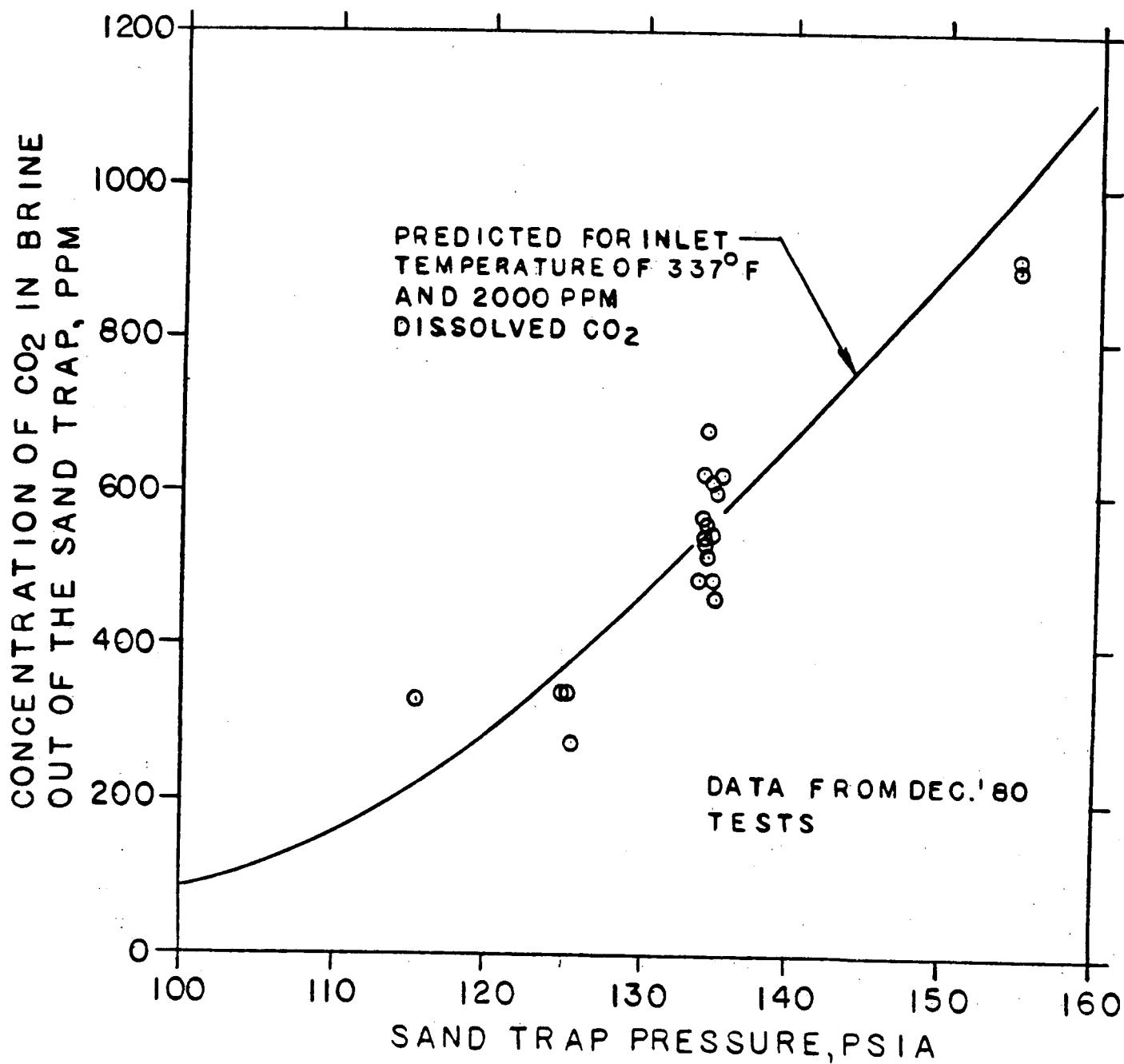


FIGURE 7

EFFECT OF DISSOLVED CO₂ IN THE
BRINE AT THE SAND TRAP EXIT ON
EXCESS PRESSURE IN CONDENSER.

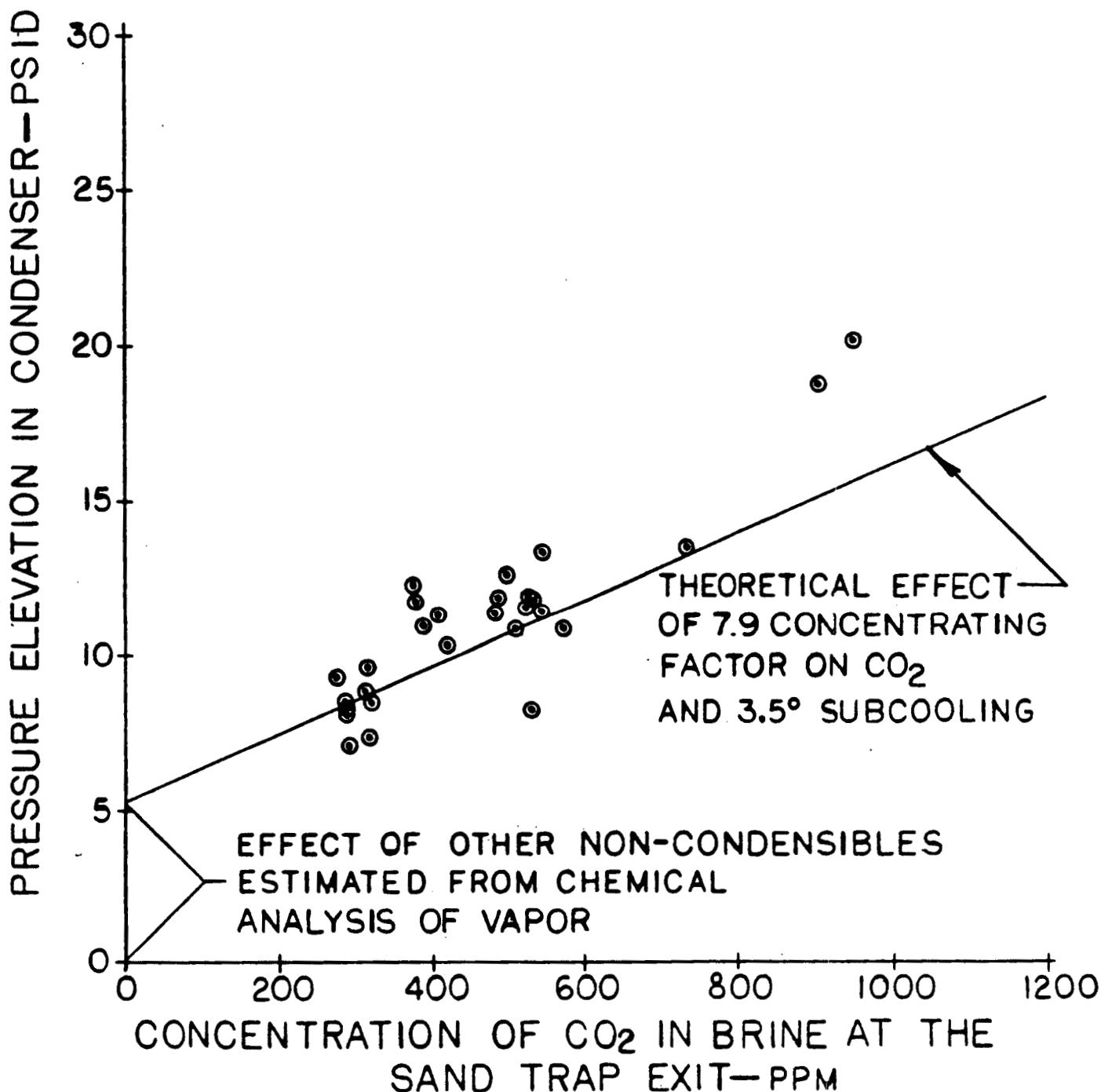
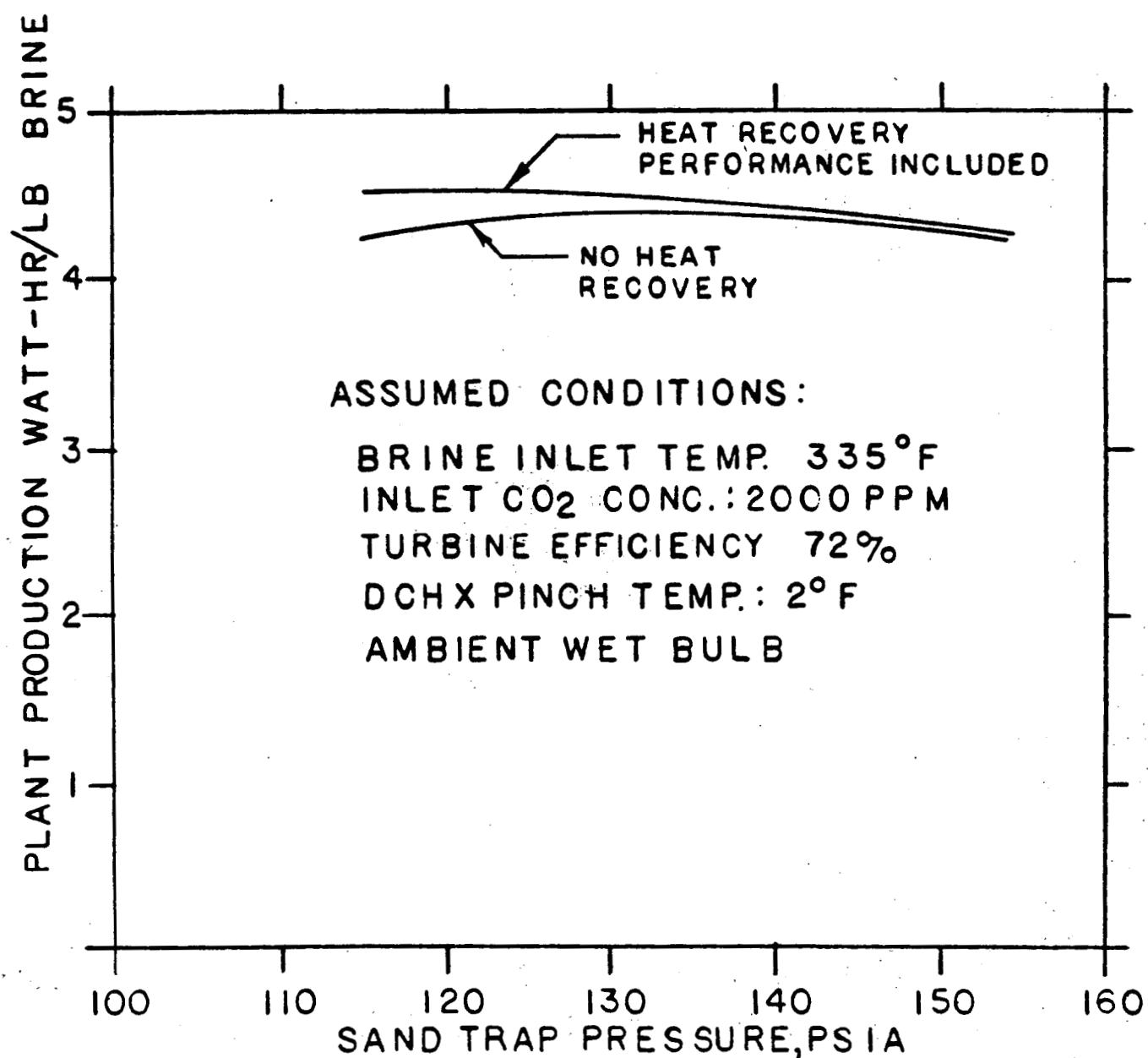


FIGURE 8

PREDICTED EFFECT OF SAND TRAP
PRESSURE ON OVERALL PLANT
PERFORMANCE



5.0 OVERALL PLANT PERFORMANCE

Pilot plant performance has been evaluated in terms of brine utilization. This parameter is defined as the net plant output (watts) divided by the brine flow rate (lb/hr). The net plant output was determined by deducting electrical parasitic losses for the plant including condenser power requirements, IC_4 and brine pumping power, and recovery system power from the gross turbine output.

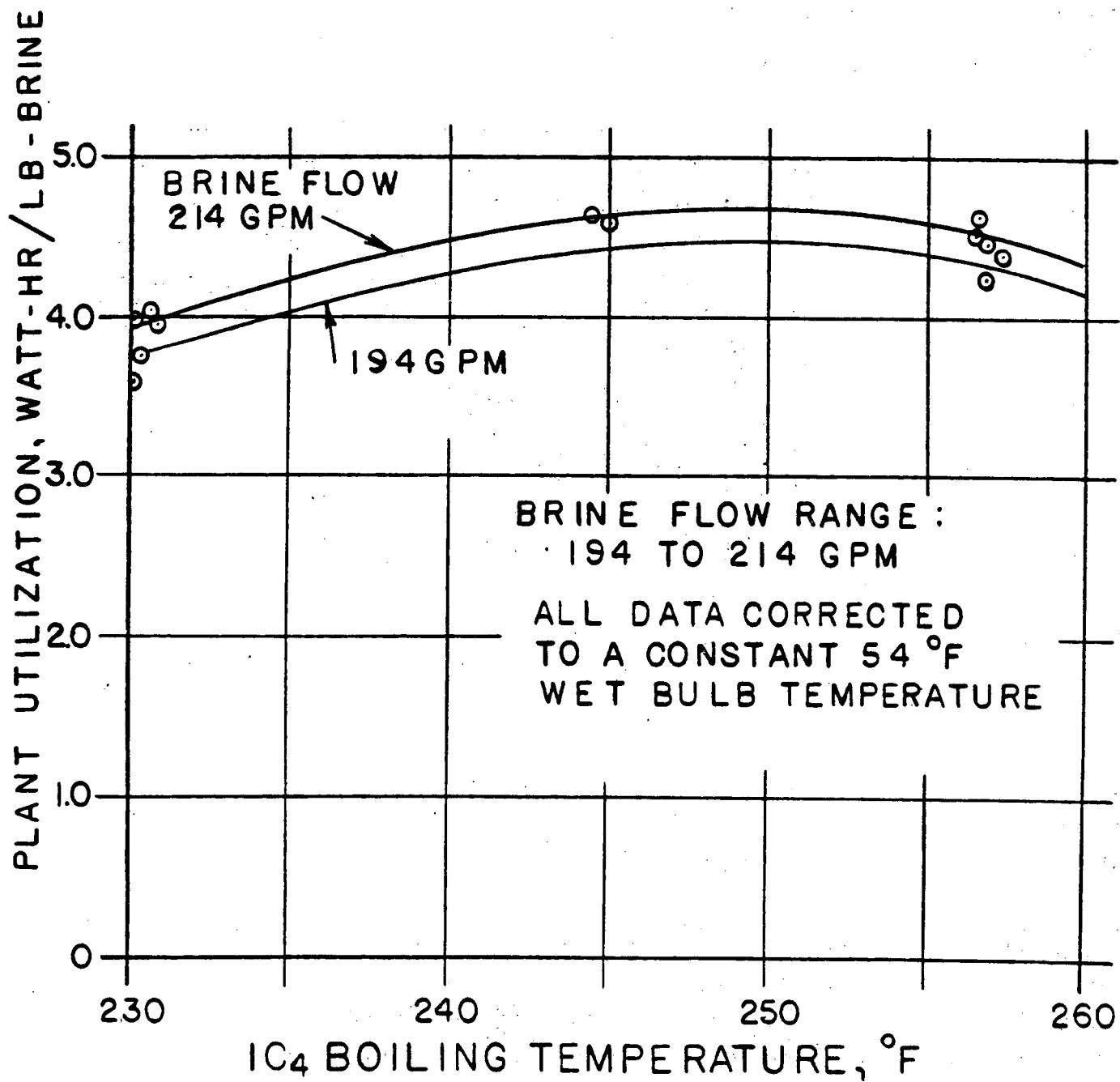
Because of high resource costs, a maximum utilization usually yields a near cost-optimum cycle, so long as reasonable temperature differences are maintained for the heat exchange equipment. For example, the 500 kW pilot plant was designed for a 7.0°F pinch temperature difference in the DCHX and a 30°F difference (condensing temperature minus wet bulb) for the evaporative condensers. With these constraints, 500 kW cycle design conditions were selected at a peak temperature of 255°F , resulting in a cycle utilization of 5.1 watt-hrs/lb. Increasing the boiling temperature above 250°F produces a slight gain in source utilization but it also causes the absolute pressure level in the system to rise, resulting in higher system costs. An increase in the system pressure level also makes both brine and IC_4 pump performance more critical to the achievement of overall performance goals.

Field tests have essentially verified the performance prediction and, with the exception of the power turbine, all components performed as expected. The predicted electrical parasitic power was 246 kW. Based on field tests, the measured loss was 250 kW, which verifies the above prediction.

Figure 9 shows measured plant utilization as a function of boiling temperature for the 500 kW pilot plant. Data shown was taken near design brine flow rate and was corrected to a constant 54°F wet bulb temperature. The maximum utilization appears to be between 4.5 and 4.7 watt-hrs/lb of brine flow at an IC_4 boiling temperature of 250°F .

FIGURE 9

VARIATION OF 500 KW PILOT PLANT
UTILIZATION WITH ISO BUTANE
BOILING TEMPERATURE



The pilot plant was operated over a range of brine flow rates and ambient conditions. The DCHX and condensers exceed design performance. With the exception of the turbine, the remainder of the hardware is achieving design goals. Even with increased CO₂ levels encountered with Mesa 6-2 brine, the plant's tested performance meets design performance objectives if the low turbine performance is accounted for.